



Integration of carbon capture technologies in blast furnace based steel making: A comprehensive and systematic review

Jorge Perpiñán^{a,*}, Begoña Peña^a, Manuel Bailera^{a,b}, Valerie Eveloy^c, Pravin Kannan^c, Abhijeet Raj^{d,e}, Pilar Lisbona^a, Luis Miguel Romeo^a

^a Escuela de Ingeniería y Arquitectura, Universidad de Zaragoza, María de Luna 3, Zaragoza 50018, Spain

^b Graduate School of Creative Science and Engineering, Waseda University, Okubo, Shinjuku-ku, Tokyo 169-8555, Japan

^c Department of Mechanical Engineering, Khalifa University, Abu Dhabi P.O. Box 127788, United Arab Emirates

^d Department of Chemical Engineering, Khalifa University, Abu Dhabi P.O. Box 127788, United Arab Emirates

^e Department of Chemical Engineering, Indian Institute of Technology Delhi, New Delhi 110016, India

ARTICLE INFO

Keywords:

Steel industry
Ironmaking
Carbon capture
Technical comparison
Systematic review

ABSTRACT

Decarbonization of the iron and steel industry, which accounts for 7–9% of global annual emissions, is a strategic objective to achieve carbon emissions reduction targets in line with climate change policies, while maintaining economic competitiveness. Carbon capture (CC) technologies are of critical importance to achieve these goals. This work presents the first systematic review of the integration of CC technologies in the blast furnace-basic oxygen furnace (BF-BOF) steelmaking route, which is expected to maintain a dominant market share over the coming decades. Integration options for post-combustion, looping cycles, oxy-combustion and pre-combustion are described and compared in terms of energy penalty, carbon emissions abatement potential, cost, technology readiness level, and practical deployment considerations. The review yielded 188 studies from peer-reviewed articles and technical papers. Research is mainly focused on chemical absorption, physical adsorption, and oxy-blast furnace technologies, but other carbon capture methods including calcium looping, Sorption Enhanced Water Gas Shift, and membranes appear promising in terms of cost and carbon emission reduction. This article provides an in-depth analysis of the current state of the art and crucial considerations for future decision making in the techno-economic selection and integration of CC technologies. Barriers to overcome for practical implementation are also identified and discussed in this article.

1. Introduction

Decarbonisation of energy-intensive industries is a strategic objective in European and National policies to accomplish the ambitious target posed in the European Green Deal (2020): to reach a Europe climate neutral in 2050 [1]. The iron and steel (I&S) industry is one of the major industrial contributors to global CO₂ emissions, and is responsible for 5 % of total global greenhouse gas emissions (GHG) [2], mainly due to the current dependency on coal-based processes [3]. In addition, it is a relevant economic sector embedded in strategic value chains that involve 1.3 % of GDP and 330,000 direct employments in Europe with an expected increase in production [4]. Current policies, together with the significant rise in the price of electricity, natural gas (NG) and CO₂ allowances in the past years, is fostering the interest of the

sector in the development of low-emission technologies.

Globally, steel is produced through three main routes (Fig. 1): the blast furnace-basic oxygen furnace route (BF-BOF), the scrap-based electric arc furnace (EAF), and the direct reduced iron-electric arc furnace (DRI-EAF). The BF-BOF route consumes mainly iron ore, coal and coke, to cover the 70 % of the global steel demand [2]. Iron ore and coal are processed on a sinter strand and a coke oven, respectively. Then, the hot metal (HM) formed in the blast furnace is sent to the basic oxygen furnace to reduce the carbon content and form molten steel.

The scrap-based EAF route uses recycled steel scrap and electricity, encompassing 23 % of the global production, but the expansion of this technology depends on the scrap availability [2]. This process consists only of one stage, the EAF, where the scrap is melted to form new steel. The emerging DRI-EAF route, accounting for 7 % production share [2], includes a shaft furnace, where iron ore is directly reduced by using

* Corresponding author.

E-mail addresses: jorge.perpinan@unizar.es (J. Perpiñán), bjp@unizar.es (B. Peña), mbailera@unizar.es (M. Bailera), valerie.eveloy@ku.ac.ae (V. Eveloy), pravin.kannan@ku.ac.ae (P. Kannan), abhijeet.raj@ku.ac.ae, raj@iitd.ac.in (A. Raj), pilarlm@unizar.es (P. Lisbona), luismi@unizar.es (L.M. Romeo).

<https://doi.org/10.1016/j.fuel.2022.127074>

Received 21 September 2022; Received in revised form 15 November 2022; Accepted 4 December 2022

Available online 15 December 2022

0016-2361/© 2022 The Author(s). Published by Elsevier Ltd. This is an open access article under the CC BY-NC license (<http://creativecommons.org/licenses/by-nc/4.0/>).

Nomenclature	
<i>Abbreviation Meaning</i>	
BF-BOF	Blast Furnace-Basic Oxygen Furnace
BFG	Blast Furnace Gas
BOFG	Basic Oxygen Furnace Gas
CC	Carbon Capture
CCUS	Carbon Capture Utilization and Storage
CaL	Calcium Looping
ChL	Chemical Looping
COG	Coke Oven Gas
DRI-EAF	Direct Reduced Iron-Electric Arc Furnace
EUA	European Union Allowance
FSCM	Fixed Site Carrier Membrane
GWP	Global Warming Potential
HM	Hot Metal
I&S	Iron and Steel
IEA	International Energy Agency
NGCC PP	Natural Gas Combined Cycle Power Plant
OBF	Oxygen Blast Furnace
Oxy-TGR	Oxy-Top Gas Recycling
OtherL	Other Looping Processes
Pre-adsor	Pre-combustion CC with physical adsorption
Post-chem	Post-combustion CC with chemical absorption
Pre-chem	Pre-combustion CC with chemical absorption
Post-memb	Post-combustion CC with membranes
Pre-memb	Pre-combustion CC with membranes
SPECCA	Specific Primary Energy Consumption for CO ₂ Avoided
SEWGS	Sorption Enhanced Water Gas Shift
TRL	Technology Readiness Level
TGR	Top Gas Recycling
TSA	Temperature Swing Adsorption
VPSA	Vacuum Pressure Swing Adsorption
WGS	Water Gas Shift

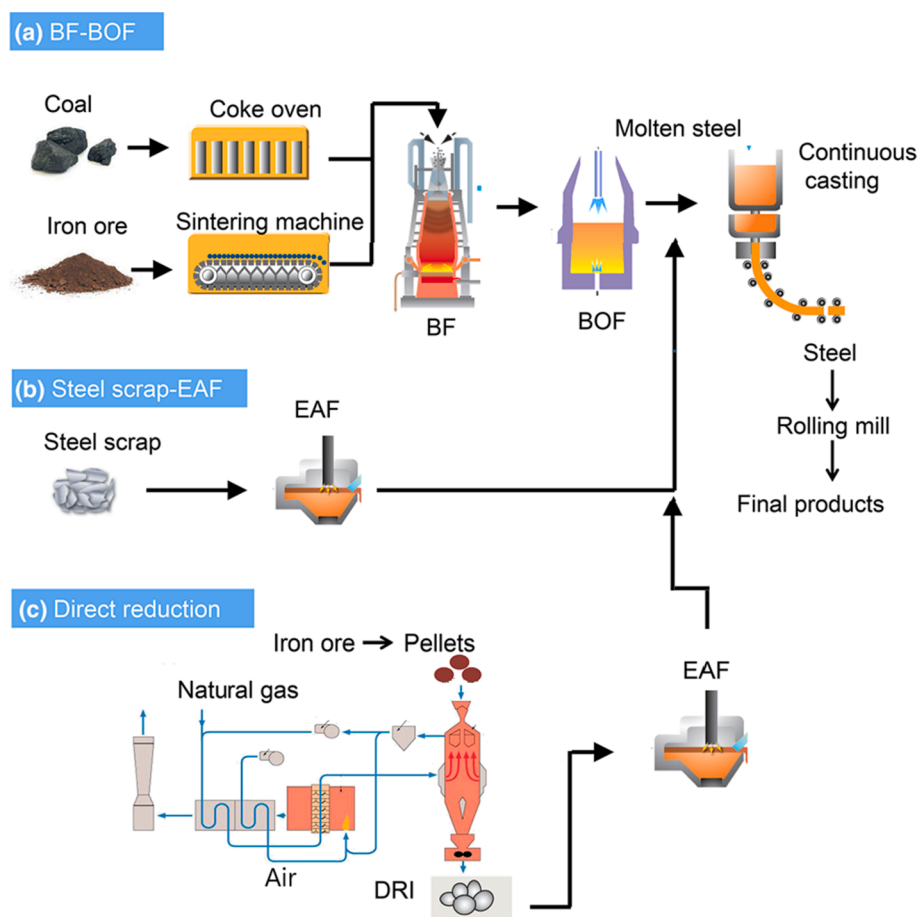


Fig. 1. The main routes of steel production. . Adapted from [7]

reducing gases such as natural gas, coal syngas, or hydrogen. The resulting direct reduced iron (DRI) is melted, together with scrap, in an EAF. MIDREX [5] and ENERGIRON [6] are commercial-scale processes for DRI production.

Specific energy consumption and CO₂ emissions are very different for each route, and contrasted in Table 1. The BF-BOF process is clearly the most energy intensive one because of the energy needs associated with

iron ore reduction in the blast furnace and with the conditioning of the raw materials, iron ore agglomeration and coking processes. The main energy inputs come from coal, giving rise to very high GHG emissions. Despite its high emissions, the BF-BOF route is still under expansion. In China, the I&S industry has developed rapidly over the past two decades, producing almost half of the current global production. Approximately 90 % of China’s steel is produced via BF-BOF route [8]. Since the

Table 1
Main steelmaking production routes: BF-BOF, Scrap-EAF and DRI-EAF [2].

Steelmaking production route	BF-BOF	Scrap-EAF	DRI-EAF
Steel world production	70 %	23 %	7 %
Main inputs	Iron ore, coal and coke	Scrap and electricity	Iron ore, NG, coal syngas, H ₂ , scrap and electricity.
Energy consumption (GJ/tHM)	13–14	4–6	10
CO ₂ emissions (tCO ₂ /tHM)	2.0–2.2	0.3–0.5	0.7–1.3

global steel demand cannot be covered through recycled scrap, the BF-BOF route is expected to maintain its dominance in the market, despite its high emissions. Besides, blast furnaces are only phased-out at relining, which typically takes places every 20–35 years, or up to 40 years for newly commissioned plants. Thus, at least 20 % of today’s blast furnaces are anticipated to still be in operation by year 2050 [9].

The reduction of greenhouse gas emissions from steel production has been the topic of extensive research worldwide. Several alternative ultra-low-CO₂ steel production technologies are under development to replace existing carbon-based steel manufacturing. The project Carbon2Chem [10] studies possible uses for steel gases, while ULCORED [11], ULCOWIN [11], ULCOLYSIS [11], COREX [12], HIs melt [13], and FINEX [14] are projects and programmes aiming at the development of new routes for steel production. ULCOS-BF [11] and COURSE50 [15] are the only projects focused on capturing CO₂ in the BF-BOF route through physical adsorption and chemical absorption methods from the BFG.

Carbon capture technologies will be necessary to achieve significant carbon emission reduction in the I&S industry. Because of the much larger steel production and specific emissions of integrated steel mills (BF-BOF route) compared to EAF or DRI-EAF, the integrated mills have been the main focus of capture technology development and cost analysis. Besides the huge carbon volume to be captured, another major drawback to integrate CO₂ capture in steel mills is the number of

different point sources (Fig. 2). The largest single point source is the blast furnace, from which 70 % of the emissions can be captured, followed by the power plant and the sinter strand. Apart from these processes, other sources emit smaller CO₂ amounts including the coke oven, the basic oxygen furnace or the casting and rolling. Since the emission rate and the mole fraction of CO₂ are very different among the various sources, multiple carbon capture plants or combining flue streams would be needed for each large mill in order to capture a large proportion of the total emissions. I&S gases, unlike flue gases from power production, have the particularity that besides CO₂ and N₂, other gas species are present at important concentrations, notably CO, H₂ and CH₄ (Table 2). However, steel gases have a higher CO₂ concentration than flue gases from power production.

Previous reviews related to low carbon ironmaking have not focused on CC or the BF-BOF route [18–21]. The main objective of this article is to present a systematic review of the state of the art related to carbon capture (CC) within the I&S industry, with special focus on the BF-BOF route. A global and holistic picture is presented to assist decision making regarding the most promising CC technology for the BF-BOF steelmaking route, by evaluating and comparing the technologies in terms of energy penalisation, economic cost, efficiency, Technology Readiness Level (TRL) [22], and the volume of CO₂ captured.

The paper is organised as follows: Section 2 presents the methodology of the systematic review; Section 3 comprises the post-combustion carbon capture systems; Section 4 include the looping carbon capture

Table 2
Typical composition for steel gases [17] and power plant flue gases (%vol).

Units	BFG	COG	BOFG	Coal PP	NGCC PP
H ₂ %vol	4	45–64	1.5	–	–
CH ₄ %vol	–	20–30	–	–	–
CO %vol	22–25	5–10	60–70	–	–
CO ₂ %vol	20–22	2–5	15–20	13	4
O ₂ %vol	–	1–4	2	5	14
N ₂ %vol	49–54	0–5	10	82	82
LHV MJ/Nm ³	3–3.8	16–19.3	7.5–8	–	–

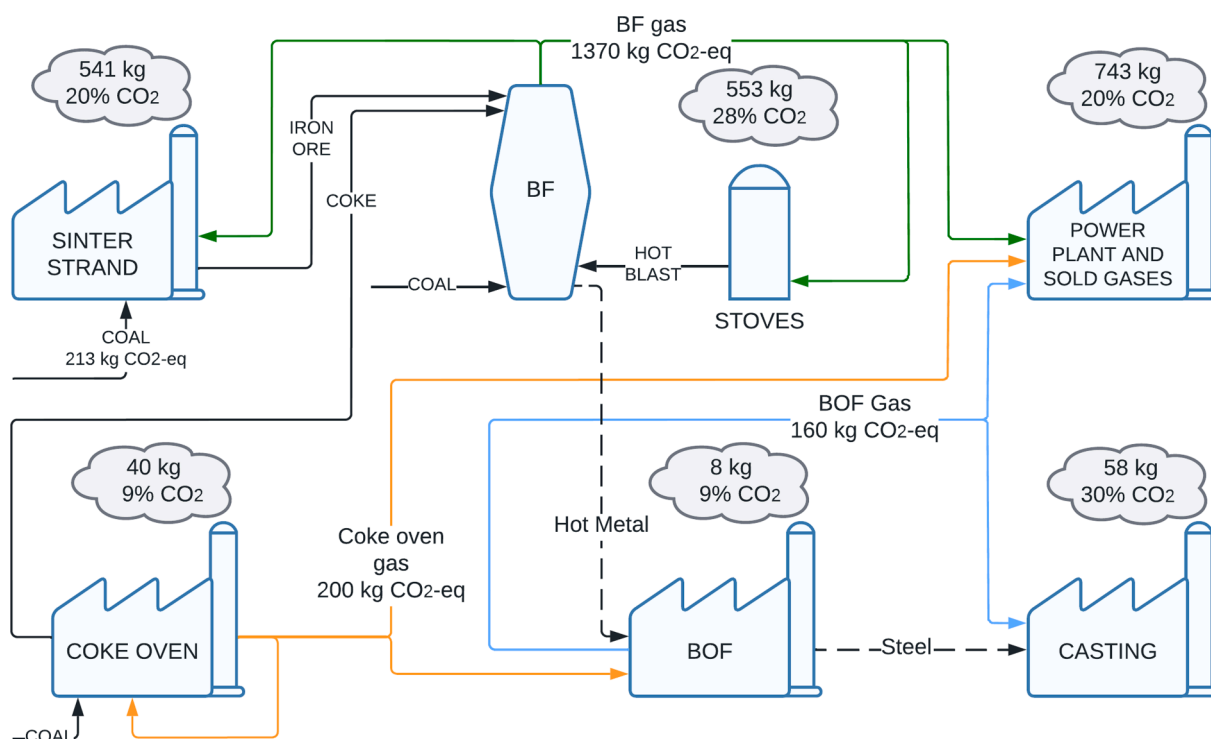


Fig. 2. Simplified flow sheet of an integrated steel mill, showing CO₂ emissions (kg/t steel) and concentration in flue gas (volume %) [16].

Table 3

Search terms, number of hits returned and number of papers reviewed.

N°	Search term	Hits	Reviewed
0	("Iron plant" OR "Iron industry" OR "Ironworks" OR "Ironmaking" OR "Iron-making" OR "Iron production" OR "Iron refining" OR "Steel plant" OR "Steel industry" OR "Steelworks" OR "Steelmaking" OR "Steel-making" OR "Steel production" OR "Steel refining" OR "Blast furnace" OR "Basic oxygen furnace" OR "Coke Oven" OR "Sinter plant" OR "Sinter strand")	Common I&S terms for all the CC searches	
1	("Carbon capture" OR "Carbon dioxide capture" OR "CO2 capture" OR "CC" OR "Carbon utilization" OR "Carbon dioxide utilization" OR "CO2 utilization" OR "Carbon utilisation" OR "Carbon dioxide utilization" OR "CO2 utilisation" OR "CCU" OR "CCUS" OR "Carbon recycling" OR "Carbon dioxide recycling" OR "CO2 recycling" OR "Carbon reduction" OR "Carbon dioxide reduction" OR "CO2 reduction" OR "Carbon mitigation" OR "Carbon dioxide mitigation" OR "CO2 mitigation" OR "Low carbon" OR "Decarbonization" OR "Decarbonisation" OR "Emission reduction" OR "Emission mitigation" OR "Emission recycling")	971	35
2	("Amine gas treating" OR "Amine scrubbing" OR "Amines" OR "MEA" OR "DEA" OR "MDEA")	92	15
3	("Oxy combustion" OR "Oxycombustion" OR "Oxy-combustion" OR "Oxyfuel" OR "Oxy-fuel" OR "Oxy-fuel combustion")	44	3
4	("Top gas recycling" OR "TGR" OR "Top gas injection" OR "BFG injection" OR "Hot Reducing Gases" OR "HRG")	118	37
5	("Precombustion capture" OR "Pre-combustion capture" OR "Pre combustion capture" OR "CO2/H2 separation" OR "Partial oxidation")	130	2
6	("Membranes" OR "Membrane separation" OR "Membrane-based separation" OR "Membrane technology")	84	7
6	("Calcium looping" OR "Calcium-looping" OR "Ca-Looping" OR "CaL" OR "Carbonate looping" OR "Chemical looping" OR "Chemical-looping" OR "Combustion looping" OR "Oxygen carrier")	68	15
8	("Pressure Swing Adsorption" OR "PSA" OR "Temperature Swing Adsorption" OR "TSA" OR "Vacuum Swing Adsorption" OR "VSA" OR "VPSA" OR "PTSA")	62	6
	Total	1569	120

systems; Section 5 contain the oxygen blast furnaces and top gas recycling technologies; Section 6 comprises pre-combustion carbon capture systems; and Section 7 presents a carbon capture technologies comparison.

2. Systematic review approach

To identify the most relevant literature on the topic, a systematic review approach was adopted. Web of Science was selected as the search engine because of its advanced search capabilities and wide range of databases. The search is limited to studies presented in scientific journals and conferences included in the selected search engine. A set of 7 searches were performed, returning a total of 1569 entries (Table 3). Each search combined a term related to the I&S process (N° 0) and another term concerning carbon capture and storage (N° 1 to 7). The hits returned from the search were screened based on their abstract, and selected for review when applying any type of CC on the BF-BOF route of the I&S industry, leading to a total of 120 articles.

The 120 research articles were reviewed and classified into four different topics, according to the typical CC technology classification: post-combustion, looping cycles, oxy-combustion, and pre-combustion [23]. Fig. 3 illustrates the different options available for each CC technology. Each type of technology, if applicable, is divided according to the specific CC technology applied within the category: chemical absorption, physical adsorption, membranes, calcium looping, chemical looping and sorption enhanced water gas shift (SEWGS). For example,

for oxy-combustion CC technology, only oxygen blast furnace technology was found (i.e., oxy-combustion CC technology of BOFG or other gases were not found in the literature).

Articles purely related to the improvement of energy efficiency, fuel consumption reduction, H₂ production without carbon capture, injection of hot reducing agents in the BF without carbon capture, steam methane reforming, and carbon dioxide reforming of methane were not included in this review since CO₂ is not captured or used directly. Articles related to CC in other steelmaking routes (i.e., DRI or EAF) were also not considered.

3. Post-combustion CC in BF-BOF

Post-combustion technologies include those systems where CO₂ is captured from the flue gases. This section reviews chemical absorption and membrane separation techniques for carbon capture in the I&S industry, and is focused on those previous works related to decarbonization of gas streams with no presence of combustible gases or including a stage of water-gas shift reactor to convert CO into CO₂. Research works, where the resulting cleaned gas is considered as a useful fuel, are classified in Section 6 as pre-combustion capture.

3.1. Post-combustion CO₂ capture by chemical absorption

Chemical absorption systems are based on chemical solvents that react reversibly with a certain target species. This method allows for a very selective separation compared to other technologies, and high purities and recoveries. Difficulties arise from the handling of the solvent systems, often exhibiting corrosive or volatile characteristics, as well as from the regeneration process of the saturated absorbent, usually endothermic, necessary for the economical operation of the separation.

The typical operation of industrial absorption consists of two stages: absorber to capture the CO₂ from the acid gas, and stripper to release the captured CO₂, as shown in Fig. 4. In the latter, the rich solution (amine + CO₂) is heated up to 110 °C in order to recover the CO₂ in a concentrated gas stream. The lean solution exiting the stripper is recycled back to the absorber. The global efficiency is increased by including a heat exchanger, where the rich solution is pre-heated before entering the stripper using the outgoing lean solution.

3.1.1. Solvents and associated energy penalizations

In the context of the I&S industry, chemical absorption has been investigated for carbon capture from the flue gases of BF hot stoves [24–28], power plants [29–32], and coke ovens [24,33]. Concurrent carbon capture from different sources has been also studied in other works, aiming for a bigger CO₂ emission reduction [34,35,44,36–43].

The most widely used chemical solvents are single-component alkanolamines in an aqueous solution. These are classified into three categories depending on the type of the amine functional group: primary such as monoethanolamine (MEA) and diglycolamine (DGA), secondary such as diethanolamine (DEA), isopropylaminoethanol (IPAE) and aminoethylethanolamine (AEEA), and tertiary such as piperazine (PZ) and methyldiethanolamine (MDEA). They exhibit different behaviour with regards to reaction rates, heats of reaction, and corrosion. Primary amines are the most reactive, possess the highest heat of reaction, and are usually the most corrosive. These properties are less pronounced for secondary and tertiary amines. In addition to single-component amines, some multi-component solvents (RITE-A and RITE-B) have been developed by the Research Institute of Innovative Technology for the Earth (RITE) [28]. These novel solvents are IPAE-based mixtures of amines.

The most commonly used amine solvent is monoethanolamine (MEA), while other solvents have barely been studied in the literature (see Table 8). The high heat of reaction results in energy penalties ranging from 2.9 MJ/kgCO₂ to 6.5 MJ/kgCO₂ for MEA (30 %wt MEA) [24,25,34,36–38,40,42,43,45,26–33] and between 2.9 MJ/kgCO₂ and 3.1 MJ/kgCO₂, for MDEA (50 %wt MDEA) [35,39]. Yang et al. [44]

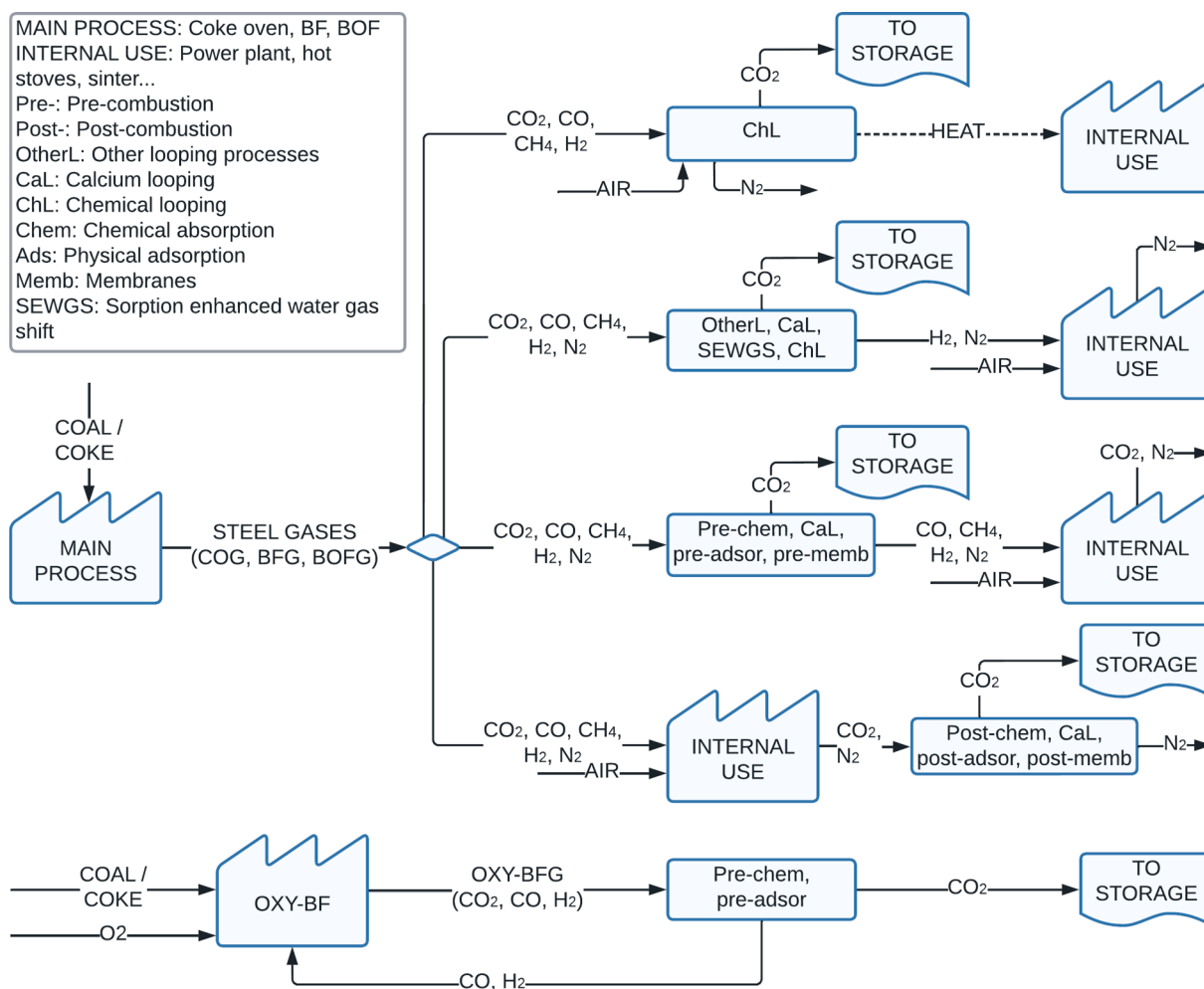


Fig. 3. Carbon capture routes in the Iron and Steel industry (BF-BOF route).

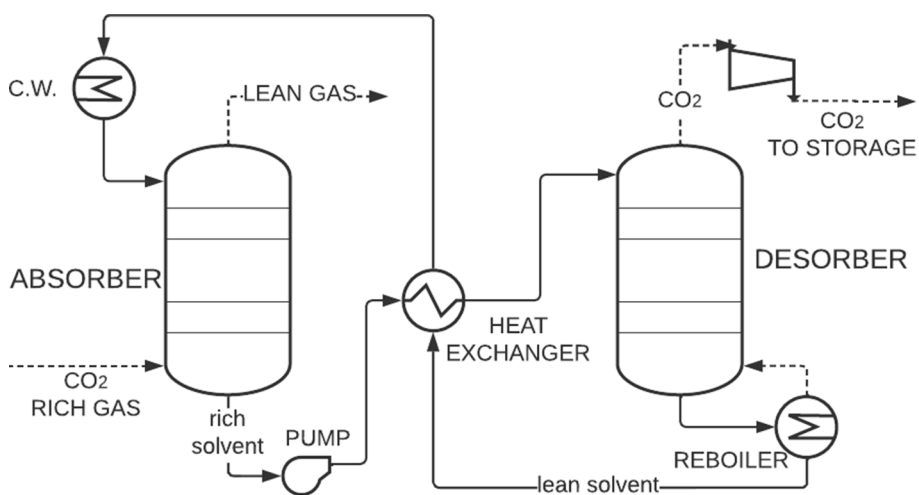


Fig. 4. CO₂ capture with chemical absorption.

investigated a mixture of 40 % MDEA with 10 % PZ obtaining a thermal consumption of 2.3 MJ/kgCO₂. For the solvents RITE-A and RITE-B, the standard thermal penalizations are 3.3 MJ/kgCO₂ and 3.1 MJ/kgCO₂, respectively, but after optimization for RITE-B, 2.5 MJ/kgCO₂ can be achieved. [28]. Cheng et. al. [25] compared the heat of reaction for MEA, AEEA, PZ, and their mixtures with respect to CO₂ capture in the

hot stove flue gas. The study showed 10 % less consumption for PZ than for MEA and AEEA, which exhibit similar thermal penalization (3.7 MJ/kgCO₂).

Among these articles, only a few works [26,30,37,38,42–44] studied the electricity consumption of carbon capture, ranging from 0.28 to 1.5 MJ/kgCO₂. This electricity consumption is related to the required CO₂

Table 4
Main KPIs for pre-combustion CC with membranes in the I&S industry.

Source	Baker (2018) [59]		Luca (2021) [41]	Yun (2021) [42]									
Membrane	Polaris		PVAm	Polyimide									
Permeab. (gpu)	1500		371	500									
Selectivity CO ₂ /N ₂	25		200	20									
Membrane configuration	Two-stage A	Two-stage B	Two-series	Two-stage B						Two-stage A			
CO ₂ Source	BF stoves	BF stoves	BOFG	BF stoves		Utility PP		Lime		Coke		Sintering	
Inlet %CO ₂	25,0	28,0	21,1	27,3		26,4		19,4		14,8		4,8	
Purification and compression	Cond + rec	Mem + Cond + rec	-	-	Cond + Mem + rec	-	Cond + Mem + rec	-	Cond + Mem + rec	-	Cond + Mem + rec	-	Cond + Mem + rec
Pressure of CO ₂ stream (bar)	150	150	120	1–2	152	1–2	152	1–2	152	1–2	152	1–2	152
Spec. Consump. (kWh _e /tCO ₂)	286	235	327	207	318	235	320	302	393	330	473	1063	1220
Spec. Consump. (MJ/kgCO ₂)	1,03	0,846	1,18	0,745	1,15	0,846	1,15	1,09	1,42	1,19	1,7	3,83	4,39
Membrane area (m ² /tCO ₂ /h)	3400	2780	5114	2543	1953	3079	2659	4080	4515	4724	5504	37,894	13,321
CO ₂ inlet (tCO ₂ /h)	41,7	220	6,8	189	189	448	448	32,7	32,7	87,4	87,4	15,3	15,2
Captured CO ₂ (tCO ₂ /h)	33,3	176	6,4	171	170	404	404	29,5	29,5	78,4	78,4	13,7	13,7
Capture rate of CO ₂ (%)	80	80	94	90	90	90	90	90,1	90,1	89,7	89,7	89,7	90
CO ₂ purity in outlet stream (%)	99	99	95,5	90									
CapEx (\$/(kgCO ₂ /h))	1482	1248		1325	1651	1537	1709	2119	2932	2319	3094	9190	8912
Cost of electricity (c\$/kWh _e)	5	5		6									
Plant life (y)				30									
Membrane life (y)	3	3	5	5									
Capacity factor (%)			93	85									
Cost of capture (\$/tCO ₂)	47	36		41,7	52,5	46,6	54	64,7	83,3	70,7	91,2	271,7	252,7

compression work for transport and storage. Both consumptions, thermal and electrical, give an overall equivalent electricity penalization between 1.1 MJ/kgCO₂ and 3.8 MJ/kgCO₂ for this specific technology.

3.1.2. CO₂ emission reduction, technology readiness level and costs

The CO₂ capture process by amine solvents is mature (TRL 9) and already commercialized in natural gas or fertilizers processing plants [46]. Nevertheless, no large-scale project has been found in the literature regarding post-combustion capture with chemical absorption in the I&S industry. This could be due to the large CO₂ amounts, the high costs, and the geological storage difficulties. Most of the published works are theoretical studies. The reported emission reductions range between 11 % and 77 %, corresponding to 230 kgCO₂/tHM and 1700 kgCO₂/tHM, respectively [24,26,36–45,27,29–35].

Only two experimental studies were identified. Cheng et al. [25] investigated the CO₂ capture in a rotating packed bed from a hot stove flue gas flow of 33 l/min containing 30 vol% CO₂. The research was focused on thermal energy consumption with different solvents (MEA, AEEA, PZ and their mixtures), but the CO₂ emission reduction was not assessed. The alkanolamine solutions with 10 %wt PZ and 20 %wt MEA or AEEA were found to be the most effective absorbents to capture CO₂. Goto et al. [28] presented a pilot plant of 1 tCO₂/day installed at the Kimitsu works of Nippon Steel Co. The CC plant was fed with the blast furnace flue gas and tested with MEA, RITE-A, and RITE-B. RITE solvents were developed with IPAE, and the best regeneration energy calculated was 2.5 MJ/kgCO₂ for an ideal operation with RITE-B.

Since post-combustion carbon capture does not offer any

technological or economical advantage (such as increased energy efficiency or fuel-saving), it will always result in a financial burden. The key factor for the installation of a carbon capture stage is the CO₂ tax price. If this price is higher than the carbon capture cost, capturing the CO₂ will be profitable as compared to emitting it into the atmosphere. Until 2018, the European Union Allowance (EUA) price was always below 16 \$/tCO₂, but since then, it has increased to reach 91 \$/tCO₂ in February 2022 [47]. According to the literature, capture costs are between 38.2 \$/tCO₂ and 204 \$/tCO₂ for MEA solvent (with an average of 77.4 \$/tCO₂) [24,26,32,36–38,40,42,43], 76–81 \$/tCO₂ for MDEA [35,39], and about 97 \$/tCO₂ for a mixture of MDEA and PZ [44]. Therefore, based on the current scenario of EUA prices, establishing a carbon capture stage is profitable in most cases.

3.2. Post-combustion CO₂ capture with membranes

Membrane is a cost-effective emerging technology for carbon capture in power plants and energy intensive industries. Gas separation with membranes takes advantage from the different permeation rates of certain materials to specific species, when both sides of the membrane are subjected to a differential pressure. The few theoretical works published in the literature related to I&S industry (TRL-2) are reviewed in this section.

3.2.1. Materials and characterization of membranes

Different materials including polymers [48], inorganic compounds [49,50], metal-organic frameworks [51], zeolite imidazolate

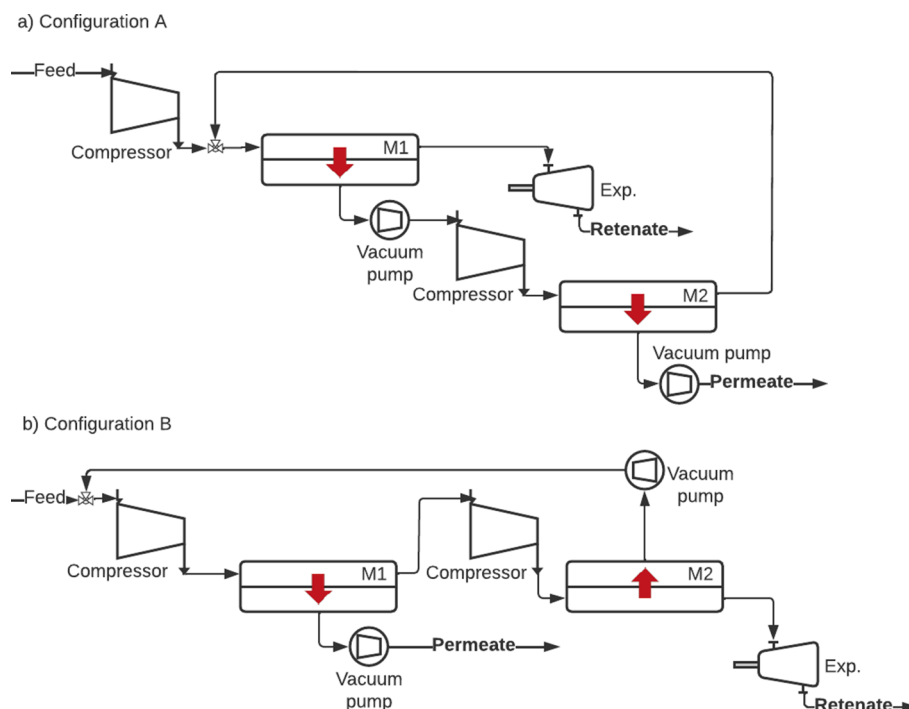


Fig. 5. Configuration of two-stage membrane-based CO₂ capture. . Adapted from [42]

frameworks [52,53], graphene oxide [54], and micro porous hollow fibers [55] have been investigated for CO₂ separation. Membrane performance is mainly assessed through the properties of selectivity and permeability, which traditionally have involved a trade-off among them. The gas permeance is the permeated flow per membrane area and per differential pressure across the membrane (gas permeance unit: $1\text{gpu} = 2.7 \cdot 10^{-3} \text{m}^3(\text{STP})/(\text{m}^2 \cdot \text{h} \cdot \text{bar})$). Permeability is the product of permeance and membrane thickness, while the selectivity is the ratio of permeability coefficients of two gases defined greater than unity [56]. Other important characteristics to be considered are stability, mechanical resistance, and easy and cost-effective manufacturing.

Gas transport mechanism mainly consists of three steps: dissolution, diffusion, and desorption of permeate (solution-diffusion transfer mechanism). Additionally, the presence of reactive carriers in the membrane may enhance mass transfer and selectivity (facilitated transfer mechanism). Accordingly, selectivity and permeability of a membrane to a certain gas depend on the relative size of molecules with respect to pore size and on the difference in the solubility, diffusivity and/or reactivity with certain mass carriers as compared to that of other species present in the gas stream. Solubility selectivity increases with the critical temperature, while diffusivity selectivity favors small molecules [48]. Higher permeability results in lower differential pressure and membrane area, and therefore lower costs, while higher selectivity produces higher purity in the permeate, but higher costs.

Commercially available membranes already exist for the separation of H₂ and CO₂ [57], i.e. Polaris™ (MTR, Inc.), the first commercial membrane developed specifically for post-combustion. Nevertheless, this capture technology is currently under development stage (TRL-6), with only one small pilot plant operating in Norcem Cement at Brevik (Norway), where 70 mol% CO₂ purity was achieved with a single membrane stage [58]. Current developments are mainly focused on power plants because of the global volume of the involved emissions [59]. However, since the costs of CC strongly decrease with CO₂ concentration in the inlet flow, steel and cement plant gases (20–30 vol%) are favored with respect to power plant gases (12–15 vol%) [59]. In the presence of CO, it is desirable to include a stage of water-gas shift reactor to increase CO₂ concentration. Additionally, since membranes are easily

poisoned by sulphur components, a previous desulphurization stage is required to lower the SO₂ concentration to levels of 10–25 ppmv [57].

3.2.2. Configurations and KPIs

There is no experimental study published in the literature related to the integration of membrane-based CO₂ capture systems in the I&S industry. Only a few theoretical research works can be found. Membranes operation is based on a sufficient partial pressure difference between the feed and the permeate stream sides. Since the BFG is at ~3 bar, there are two options: (i) to pressurize the BFG before entering the separation system [57], or (ii) to install a vacuum pump on permeate-side [59]. The latter option is preferred as it consumes less energy.

Given that the optimal range for a low-cost capture based on membranes is that where the rate of CO₂ capture is from 50 to 70 % in the first step [59], at least two membrane stages are needed for higher capture rates. Different configurations have been simulated for the principal CO₂ emission sources in I&S industry. The main results found in the literature are summarized in Table 4. Luca & Petrescu [41] investigated a CO₂ capture system with two spiral wound membranes connected in series with recirculation of 30 % and 5 % of permeate and pressure ratios of 100 and 80 for the first and second stage, respectively. The flue gas stream contained 21.1 vol% CO₂, while the output CO₂ stream reached a purity of 95.5 %. The capture rate was 94 % with a specific energy consumption of 327 kWh/tCO₂.

The two-stage concepts of Fig. 5 have been also explored in [42,59]. Retentate or permeate after the second membrane is recycled in configurations A and B, respectively. The former is more suitable for 20–25 vol% CO₂, while streams containing more than 25 % are better treated by the latter. Baker et al. [59] used Polaris™ membrane [60] in both cases including a CO₂ condensation column for purification (>99 %, 30 bar, –25 °C) and the subsequent compression up to 150 bar. Configuration A was applied to a stream of 1000 t_{CO2}/day with a typical composition of cement industry (25 vol% CO₂), which is similar to that of the flue gas of BF stove or of utility power plant in I&S industry. The first stage increases CO₂ concentration up to 40 vol%, capturing about 60 % of incoming CO₂. The second stage enriches permeate gas above 75 vol%, reaching a global capture rate of 80 %. The electric

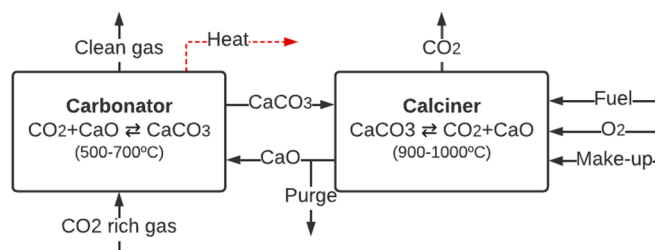


Fig. 6. Configuration of a Calcium Looping-based CO₂ capture process.

consumption is 286 kWh/tCO₂, the required membrane area is 3400 m²/(tCO₂/h) and the total cost of capture is 47 \$/tCO₂. Configuration B was applied to a typical composition of BF stoves flue gas [61] with 28.5 vol % CO₂. The first and second membranes reduce CO₂ concentration from 28 % to 17 % and 8 %, respectively, representing a global capture of 80 % of incoming CO₂. The resulting permeate stream is further enriched (above 80 %) through a third membrane module before the condensation column. The electric consumption is 235 kWh/tCO₂ and a membrane interface of 2780 m²/(tCO₂/h) is required. The total cost of capture is 36 \$/tCO₂.

Yun et al. [42] used a polyimide membrane in both the configurations of Fig. 5, analyzing separately the effect of purification (dehydration and stripping column) and final compression up to 152.7 bar. For the sake of comparison, the capture rate and the mole fraction of CO₂ were set to 90 %. Configuration A was applied to sinter strand (4.81 vol %) and coke oven (14.77 vol%) gases, while configuration B was implemented for flue gases of lime production (19.41 vol%), power plant (26.43 vol%) and BF stoves (27.3 vol%). Electricity consumption and the required membrane surface area sharply increase below 10 vol% CO₂ in the inlet stream. Carbon capture costs increase with lower CO₂ incoming content from 42 \$/tCO₂ in the latter case (BF stoves) to 272 \$/tCO₂ in the former case (sintering process).

According to these theoretical results, membrane-based capture systems could be a competitive solution for CO₂ rich streams. Nevertheless, further experimental research is required for such systems to become available at commercial scale.

4. Looping processes for CC in BF-BOF

Carbon capture with looping processes include those systems where CO₂ is captured either in combustible gases or in flue gases. This section review calcium looping (CaL), chemical looping (ChL) and other looping processes (OtherL), including Ca-Fe or Ca-Cu for carbon capture in the I&S industry.

4.1. Calcium looping

Calcium looping (CaL) is a CC technology where calcium oxide (CaO) is reacted with CO₂ to form calcium carbonate (CaCO₃), and CaCO₃ is then heated to regenerate CaO and CO₂ in a calciner (see Fig. 6). The calcination process is endothermic, and a fuel input to the process is required. The captured CO₂ can then be used or transported for storage. The few theoretical works published in the literature related to I&S industry (TRL 2–3) are revised in this section.

Cormos performed a techno-economic evaluation of a Calcium Looping (CaL) carbon capture process in post-combustion as the alternative to decarbonise an integrated steel mill [35]. A conventional integrated steel mill (4 million tonnes Hot Rolled Coil per year output) with CaL carbon capture was analysed under two scenarios which accounted for the most important CO₂ emission sources within the steel mill (e.g., power plant, hot stoves, lime plant, coke ovens). The carbon capture efficiency, specific CO₂ emissions, and CO₂ capture investment costs of the integrated steel mill with CaL capture process were evaluated (see Table 8). The evaluated captive power plants and the carbon

capture process were thermally integrated using pinch method. Heat integration is of importance considering the multiple heat sources within the steel plant to be used to cover the energy duties. Heat integration is especially critical for CaL processes to quantify the high temperature heat recovery potential. Chisalita et al. explored the environmental aspects of this configuration through a Life Cycle Assessment using the functional unit of 1 tHRC [34]. Global warming potential (GWP) of the integrated steel mill was reduced by 64–75 % when CaL was coupled (emitting 516–754 kg CO₂ eq/tHRC). The GWP even shows negative values given the negative CO₂ emissions allocated to the exported electricity. Nevertheless all other environmental indicators (acidification potential or eutrophication potential) have a more or less significant increase compared to the benchmark case, associated to the extraction and transportation of the natural gas required in the calciner (71.8–119 kg of natural gas/tHRC). The economic penalty of CaL capture leads to a 6 % increase of specific capital investment and a 12 % increase of steel production cost [39].

CaL capture process can be also applied in precombustion for the production of hydrogen from raw coke oven gas (COG) by means of CO₂ sorption enhanced steam reforming using CaO as CO₂ sorbent (CaL-SESR). Xie et al. thermodynamically analysed this process, and estimated equilibrium compositions, energy consumption, and CO₂ emissions [62]. This configuration can lead to concentrations over 95 vol% H₂ in the reforming gas while reducing the temperature of the reforming processes, the CO₂ emission, and the total energy demand per unit of generated hydrogen in comparison to conventional reforming. The CaL-SESR process avoids 99 % of CO₂ generated from the COG after a conventional reforming.

Halmann and Steinfeld also studied the enhanced reforming of blast furnace gas (CO₂, CO, N₂ and H₂) through CaL capture process from a thermodynamic modelling approach. Partial or complete carbon capture can be achieved depending on the operating conditions of the SESR reactor and a relatively pure CO₂ produced. The implications of such reactions with respect to hydrogen production, CO₂ emission avoidance and process efficiency are presented in this work [63].

Liu et al. experimentally explored the sorption enhanced water gas shift (SEWGS) reaction to improve the carbon capture from BFG. The CO in the BFG was first converted to CO₂ and, then, captured through carbonation reaction. The lab-scale experimental results from this study highlighted the importance of the sorbent type on the SEWGS reaction, with both CaO and MgO in the sorbent catalyzing the reaction. WGS reaction becomes the rate limiting step for calcined limestone, while for calcined dolomite, WGS reaction is not a limiting step and most of the CO₂ for the CaO carbonation comes from WGS reaction on MgO surface. Besides, it was observed that not only CaO but also MgO experienced the decay in the catalytic reactivity after multiple cycles [64]. Tian et al. deepen the study of this concept combining CaL CO₂ capture and waste recycling into I&S production. The sorbents used in this experimental study were CaO-based, 90 wt% CaO, obtained from steel slag which showed better reactivity and slower deactivation than commercial CaO. The preparation of steel slag as sorbents co-produces high-quality iron ore with iron content of 55.1–70.6 % which compensate the larger costs when compared to naturally derived CaO. The cost of the steel slag-derived CO₂ sorbent achieves 57.7 €/t, which makes this material a cost-effective option for CO₂ removal in I&S industry. [65].

Tian et al. also performed a techno-economic study of this CaL decarbonisation concept for steelmaking showing a substantial and cost-effective CO₂ emission reduction and the superiority of CaL in comparison to other technologies [66]. The cost of avoided tonne of CO₂ is reduced down to 12.5–15.8 €/tCO₂ (below one third the cost obtained using amine scrubbing technology) while the emissions can be reduced up to an 80 %.

Calcium looping technology has a demonstrated TRL 6, achieved through experimental experimentation with fluidized bed reactors at significant scale and relevant operating conditions for industrial applications [67–69]. Over fourteen CaL experimental plants of different

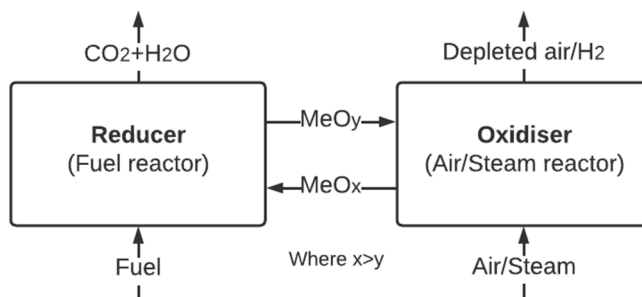


Fig. 7. Configuration of a Chemical Looping-based CO₂ capture process.

sizes from few kWth to MWth have been constructed and operated worldwide [70] while its feasibility at 1–2 MWth scale has been demonstrated through the operation of two large pilot plants [71–73]. The concept itself has been completely demonstrated and its applications to industry has been explored at lower TRLs through the integration with power and cement plants [69,70]. Several ongoing projects target to demonstrate the technical feasibility of industrial integration of CaL capture processes at large scale in relevant environments. CLEANKER project aims at demonstrating at TRL 7 the CaL concept with 90 % of CO₂ capture in a fully integrated configuration with a cement production process. Instead of fluidized beds, this project proposes the use of entrained flow reactors which are well-known equipment in the cement industry [74]. C4U project also targets to elevate two modified CaL carbon capture processes, DISPLACE and CASOH, from TRL 5 to 7 and to design for optimal integration in the steel industry [75].

4.2. Chemical looping

The Chemical Looping (ChL) concept is based on the transfer of oxygen from air to the fuel by means of a metal oxygen carrier, avoiding direct contact between fuel and air (see Fig. 7). Fig. 8.

The utilisation of chemical looping (ChL) as an option of CO₂ capture in the I&S industry has focused on the development of alternatives for the combustion coke oven gases (COG) without CO₂ emissions, or H₂ production for the steel process or other uses such as NH₃ production.

In the case of power production with steel gases, a chemical looping combustion (CLC) combined cycle using COG as fuel was proposed in [76]. Low TRL levels (2–3) are reached as the technology concept has been just formulated using ASPEN simulation considering Ni-based oxygen carriers NiO/NiAl₂O₄ (60:40 wt%) due to good results for CO₂ capture applications [76]. Nevertheless, others oxygen carriers composed of Fe₂O₃/CuO and MgAl₂O₄ have been investigated using

thermogravimetric analysis (TGA) and a laboratory pressurised circulating fluidised bed system for this specific applications in a 3–10 kWth test rig [77].

H₂ production based on CLC has been also suggested for decarbonising I&S production. For Chemical-looping water-splitting (CLWS), an additional reactor is required in a three-stage process that combines the concepts of CLC and steam-iron process [7879]. In this new stage, compared with CLC, steam reactor, the oxygen carrier is partially oxidised by steam to yield H₂ that provides a decarbonised fuel source for producing direct reduced iron (DRI) [78]. Low TRL are also suggested (2/3), with the concept formulated and some integration schemes analysed to improve state-of-the-art options by reducing the amount of coke used in the sintering plant, increasing thermal energy efficiency, CO₂ capture ratio [80] and exergy efficiency [79]. Moreover, some researchers have also proposed to use H₂ generated to produce ammonia, integrating this scheme with pressure swing adsorption technology [81]. In all these alternatives, 100 % direct CO₂ capture efficiency and improved economic figures, compared to the present technologies, could be achieved.

Regarding CLC, there is a general consensus that the maturity of the technology has reached a TRL of approximately 6 [46,82–84]. At least 46 CLC pilot reactors have been successfully constructed and operated with gaseous, liquid, and solid fuels [85,86]. Although the CLC concept and its performance has been completely demonstrated, its applications to industry remain unexplored. Fortunately, recent planned research projects have been proposed to fill this gap. CHEERS project is going to develop a CLC system, with inherent carbon capture, to increase CLC TRL to 7 [87]. The system is designed for deployment in a refinery environment, with petroleum coke as fuel, and to produce high temperature steam. It will have a thermal power of 3 MW and will combine oxygen polisher, a heat recovery steam generator, CO₂ capture, as well as compression and purification of the CO₂ in compliance with

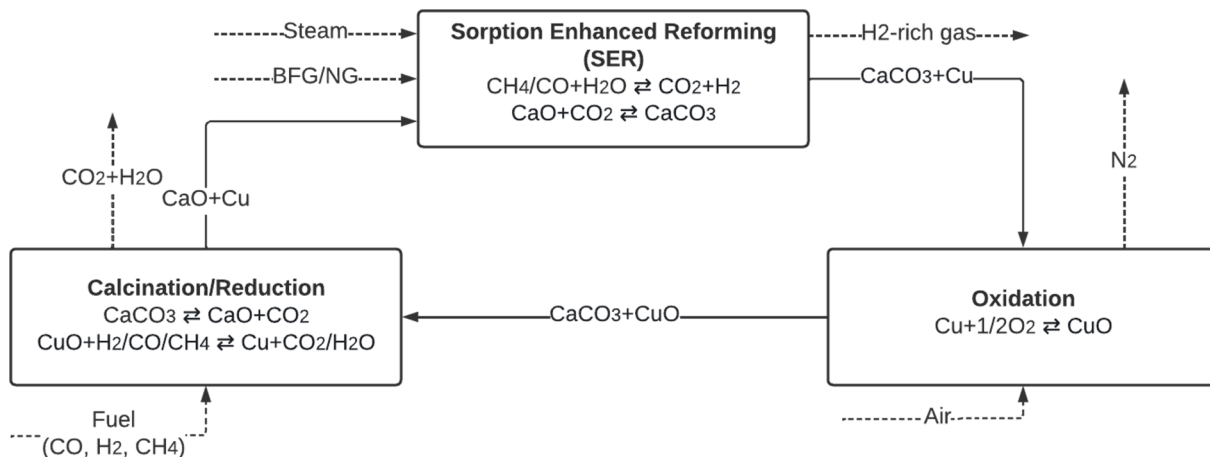


Fig. 8. Configuration of a Ca-Cu based CO₂ capture process, adapted from [90,91]

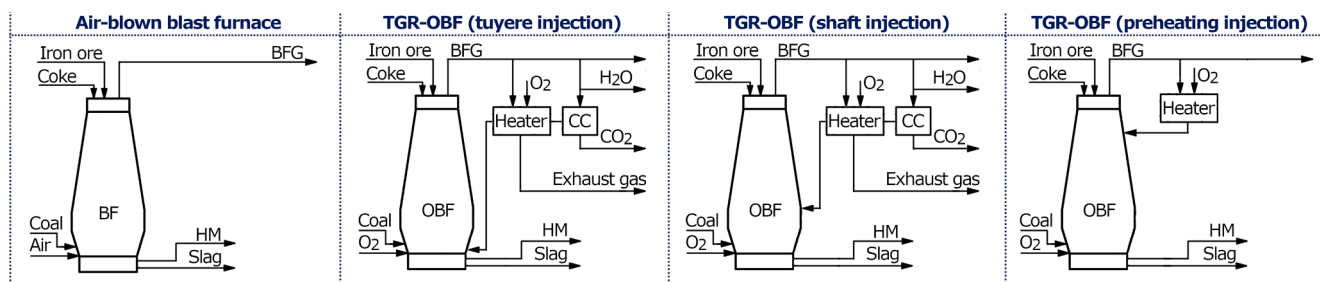


Fig. 9. Process flow diagrams of oxygen blast furnaces with top gas recycling, compared to an air-blown blast furnace.

specifications for enhanced oil recovery [88]. Also, LOUISE project [89] aims to demonstrate CLC of solid refuse fuels using ilmenite as the oxygen carrier at a TRL 6 (150 kWth and 1 MWth) and to accelerate the deployment of CLC by providing the basic design of a 10 MWth CLC demonstration plant (TRL 7) for waste-derived fuels. There are no specific projects or applications of CLC for the steel sector, but the challenges that ongoing projects have to face would be similar to those of CLC in the steel sector.

4.3. Other looping processes

The combination of calcium Looping (CaL) and chemical looping combustion (CLC) has also been proposed for the blast furnace gas (BFG) utilisation. There are two main routes for this process depending on the metal that is combined with calcium: Ca-Fe or Ca-Cu. Both alternatives are promising and have similar level of development (TRL 2/3).

Combined CaL processes have been investigated by Tian et al., who developed a new material to be used in a combined Ca-Fe chemical loop using steel slag as feedstock. In the case of iron, using steel slag as feedstock, the CaO-based, Fe-functionalized CO₂ sorbent (Fe₂O₃ and CaO) have shown promising potential with stable cyclic CO₂ uptake of 10 realistic combined Ca-Fe looping cycles [92]. This is also a three-stage process, where the heat released due to FeO oxidation provides the heat to drive the endothermic CaCO₃ decomposition reaction.

For the copper alternative, there are two schemes proposed. The first one, based on packed-bed, converts mainly blast furnace gas (BFG) into H₂/N₂ together with the release of a large amount of high temperature heat [8]. Around 30 % of the BFG can be upgraded via calcium looping-assisted water gas shift (WGS). Decarbonised fuel gas could be used for onsite power generation or to obtain sponge iron by a Direct Reduced Iron (DRI) process, thus increasing the overall capacity of the steel plant and avoiding the CO₂ emissions. Low values for the specific energy consumption of around 1.5 MJLHV/kgCO₂ and CO₂ capture efficiencies higher than 95 % support the further development of the proposed Ca-Cu looping [91]. The second scheme uses a sorption enhanced water gas shift (SEWGS) reactor with blast furnace gas (BFG) of steel mills. CO₂ is simultaneously removed from the gas using a CaO based sorbent. A Cu/CuO chemical loop supplies the energy required for the regeneration of the sorbent via the exothermic reduction of CuO with coke oven gas (COG) [93]. Approximately 27 % of the BFG can be decarbonised in the SEWGS reactor while also producing H₂. A CO₂ capture ratio of 31 % (CO₂ capture efficiencies of up to 95 %) with respect to the total carbon emissions in the steel mill can be achieved. More than 60 % of the thermal input can be recovered as high-temperature heat, which could be efficiently recovered for producing electricity [93]. This represents a clear improvement when compared with a steelmaking plant with a post-combustion MEA absorption system [31] and allows the utilisation of the excess of H₂-rich gas in the Ca-Cu steel plant to produce additional iron through a direct reduction iron (DRI) process. It has been demonstrated that the production capacity of the steel plant can be increased by around 10.5 % without installing a purpose-built and high-cost NG reforming plant [31,90].

Sun et al. proposed a system for coupled decarbonisation and

desulfurization of BFG via a magnesium-molybdenum looping process [94]. The system includes a desulfurization-decarbonization reactor and a regenerator. Four carbon-sulfur carriers were screened, highlighting MgO-MoO₃ as the most capable sorbent for decarbonisation and desulfurization. Results showed that temperatures over 240 °C decrease in carbon capture efficiency. Carbon removal efficiency reached values up to 99.70 % and complete sulfur removal.

5. Oxygen blast furnaces and top gas recycling for CC in BF-BOF

The most spread strategies for carbon mitigation in the I&S industry are the oxy-fuel combustion in the oxygen blast furnace (OBF) and the top gas recycling (TGR). The former uses O₂-enriched air as hot blast for the combustion of coke and the auxiliary fuels, producing a top gas with very little content of nitrogen [95], while the latter recirculates part of the top gas to the blast furnace, usually after a carbon capture stage. Both strategies are usually combined to make use of the synergies among them.

5.1. Type of configurations and associated penalties

The difference between oxy-fuel systems for power production, and oxygen blast furnaces, is that the former performs complete combustion (i.e., exhaust gas mainly composed by CO₂ and H₂O) [96]. In oxygen blast furnaces there is incomplete combustion to obtain CO and H₂, which are then used to reduce the iron oxides (i.e., top gas composed mainly of CO, CO₂, H₂, and H₂O) [97].

Conventionally, in oxy-fuel combustion, a part of the exhaust gas is recirculated to palliate the lack of nitrogen and its sensible heat, aiming for keeping similar fluid- and thermo-dynamic behavior than in air-blown combustion [98,99]. However, injecting CO₂ or H₂O to the blast furnace is not desirable as it increases the coke consumption and hinders the reduction of iron oxides because of displacing the chemical equilibrium. Therefore, a carbon capture stage is usually included before recycling the top gas, in order to reject the 90 % to 100 % of the CO₂ [100–105] (water is also typically removed through condensation stages). Moreover, recirculating CO and H₂ (reducing agents) diminishes the required amount of coke in the blast furnace, what lowers the CO₂ emissions. Typical capture technologies are vacuum pressure swing adsorption (VPSA) and amine scrubbing. The VPSA presents lower electricity consumption 0.5–0.6 GJ/tCO₂ [103,106], but the CO₂ streams obtained is not of sufficient purity for permanent storage, thus requiring additional cryogenic distillation to reach 99 % purity (cooling necessities of 0.1 GJ/tCO₂) [106]. In the case of amine scrubbing, the main disadvantages are the degradation/replacement of the amines, and the requirement of a non-negligible amount of steam to desorb the CO₂ (2.0–4.0 GJ/tCO₂ of thermal energy, equivalent to 0.7–1.4 GJ/tCO₂ of electricity penalty) [100,107].

In blast furnaces, the top gas can be recirculated not only at the combustion region (i.e., raceways, at the lower tuyeres zone), but also at the beginning of the indirect reduction zone (i.e., shaft, mid zone), at the preparation zone (i.e., preheating, upper zone), or at a combination of them (Fig. 9). Injecting the recirculated gas at the lower tuyeres allow

Table 5

Range of values found in the literature for the inlet and outlet streams that can be considered in a typical TGR-OBF process boundary. Not all the inlet streams will necessarily exist in a given process flow diagram.

TGR-OBF process flow diagram	Flow	Temp. (°C)	References
Iron ore (kg/thm)	1450–1730	25	[100,103,106,111,112,127,139]
Coke (kg/thm)	159–468	25	[100,102,103,106,108,112,113,136]
Auxiliary tuyere injections			
PCI (kg/thm)	50–320	25	[100,102,103,106,112,114,136,141]
Natural gas (kg/thm)	42–150	25	[100,109]
COG (Nm ³ /thm)	140	25	[112]
Recycled top gas (Nm ³ /thm)			
Injected in the preparation zone (after combustion)	200–294	1000	[100,109]
Injected in the shaft (after CO ₂ separation)	82–600	900–1000	[100,103,108,111,112,114,117]
Injected in the tuyeres (after CO ₂ separation)	60–620	25–1250	[100,103,106,108,111–116]
O ₂ /Enriched-air			
O ₂ (Nm ³ /thm)	183–353	25–1150	[105,108,112]
H ₂ O (g/Nm ³)	2–60		[100,108,111,140,142]
N ₂ (Nm ³ /thm)	0–118		[100,106,111–113,117]
Hot metal production (thm/h)			
Slag (kg/thm)	150–570	1350–1500	[21,100,143,102,105,106,108,112,114,116,140]
	189–364		[100,105,108,111,142]
BFG (before recycling) (Nm ³ /thm)			
CO (%vol)	1030–1447	100–326	[100,102,136,139,103,105,106,109,111,112,116,128]
CO ₂ (%vol)	23–52		[102,103,137,105,106,109,111,112,114,116,136]
CO ₂ (%vol)	22–48		[102,103,137,105,106,109,111,112,114,116,136]
H ₂ (%vol)	8–26		[102,103,137,105,106,109,111,112,114,116,136]
H ₂ O (%vol)	4–6		[103,111,112,116,136,137]
N ₂ (%vol)	1–15		[102,103,105,106,109,111,112,136]
Operating hours (h)			
	8600		[106]

for a proper interaction between the reducing gas and the solids, but it decreases the flame temperature. Contrarily, injections at mid shaft avoids decreasing the flame temperature, but the peripheral injection leads to a limited penetration and diffusion of the gas to the center of the descending burden (poor solid-gas interaction) [100]. In the case of upper injections, the role of the recirculated gas is only to preheat the descending solids, mitigating the lack of sensible heat that would provide N₂. As the gas is not aimed to be used as a reducing agent, the diffusion of the gas towards the center is not critical, and the presence of CO₂ is not a problem [100]. Depending on the configuration, the recirculating ratio of the top gas may vary between 30 % and 100 % [100,103,104]. It implies a decrease in the available gases downstream for the I&S plant, and therefore a penalization of 1.5–4.0 GJ/thm in terms of available thermal energy (or 0.5–1.4 GJ/thm in terms of electricity) [100,108]. A potential strategy to palliate this effect is to replace pulverized coal by natural gas, what will increase the H₂ content of the BFG and its calorific value [100]. Typical calorific value of top gas in OBF is 5.7–9.2 MJ/Nm³, while in conventional BF is around 3.0 MJ/Nm³ [102,103,109,110].

To fulfill its role as provider of sensible heat, the recirculated gas is normally injected at 900–1250 °C (in the case of mid and upper injections, at 1000 °C maximum to avoid coke gasification) [100,103,115–117,105,106,108,109,111–114]. To preheat the recirculated gas, pebble-heaters fueled by top gas itself are used (typical heat losses to the ambient are around 5 %) [106]. About 7 % to 12 % of the BFG is used as fuel in the preheating, mixed with other fuels if necessary (e.g., liquefied petroleum gas, natural gas, and/or COG) [103,104,106,110,114]. Hence, this implies an additional decrease on the available thermal energy downstream for the I&S plant of 0.5 GJ/thm (equivalent to 0.17 GJ/thm of electricity) [106]. Another option could be to heat the O₂-enriched blast instead of the recirculated top gas, but this solution is not adopted in practice due to technical limitations in the hot stoves (maximum 32 %vol O₂ in the stoves, in order to avoid

equipment damage by oxidation), so the O₂ is injected cold at 25 °C [118]. Furthermore, it was shown that at high recycling ratios, is more cost-effective to heat the recycled gas rather than the oxygen blast [118].

In addition to the impact on the available gases (less energy available for self-sufficiency), oxygen blast furnaces also imply greater electricity consumptions in the air separation unit (ASU). The volume fraction of O₂ in the enriched hot blast changes from 0 % to 5 % in conventional BF to 70 % – 100 % in OBF [100,104–106,110–113,117], so the O₂ produced in the ASU increases by 150–280 Nm³/thm [106,109,116,119]. Depending on how modern is the technology of the steel plant, the ASU consumption may vary between 260 and 400 kWh/tO₂ [106,109], meaning increments of 0.2 to 0.6 GJ/thm in the electricity consumption.

In total, accounting for all the penalizations (CO₂ separation stage, top gas recycled in the OBF, top gas consumed for heating, and ASU consumption), the thermal energy penalization is 2.0–8.6 GJ/thm, and the electricity penalization is 0.7–1.2 GJ/thm (both combined give an overall equivalent electricity penalization of 1.4–4.2 GJ/thm). Taking into account that integrated steel mills produce only 8.6 to 10.7 GJ/thm of thermal energy in the form of COG and BFG, and that they use between 3.9 and 8.0 GJ/thm of this thermal energy in internal processes (i.e., in coke oven, hot stoves and power plant) [95,109,120–126], the fact of having penalizations associated to OBF of 2.0 to 8.6 GJ/thm may lead to situations of non-self-sufficiency. Therefore, the shortage of available gases is one of the main handicaps of this technology.

5.2. CO₂ emission reduction

Besides providing sensible heat, the recycled top gas acts as reducing agent inside the blast furnace as it is mainly composed by CO and H₂. This allows to reduce the coke consumption by 14–150 kg/thm, at coke replacement ratios of 0.17–0.25 kg_{coke}/Nm³ (5–34 % decrease with respect to conventional BF) [11,108,110,112,114,127,128]. Although

Table 6
Range of values found in literature for different characteristic data of oxygen blast furnaces.

	Value	Reference
Characteristic data of TGR-OBF		
Volume fraction of O ₂ in enriched air (%)	70–100	[100,104–106,110–113,117]
Percentage of BFG recycled into BF (%)	30–100	[100,103,104]
Percentage of CO ₂ captured from the recycled gas (%)	90–100	[100–105]
Percentage of BFG used for heating the recycled gas (%)	7–12	[103,104,106,110,114]
Calorific value of top gas (MJ/Nm ³)	5.7–9.2	[102,103,109,110]
Thermal reserve zone temperature (°C)	700–950	[100,103,108,127,140]
Chemical efficiency (-)	0.94–0.98	[100,109]
Heat losses (MJ/thm)	230–630	[100,108,110,111,127,137,139]
Percentage of direct reduction (%)	4–19	[103,109,110,129,135,136]
Minimum flame temperature (°C)	1800	[100,108,113,137,138]
Decrease in flame temperature by the recycled gas (°C/Nm ³)	1.4–5.0	[108,110,114]
Replacement ratio (kg/Nm ³)	0.17–0.25	[11,108,110,112,114,127,128]
Gas utilization (%)	47.7–54.3	[109,112]
Furnace inner volume (m ³)	2500–2800	[103,109,140]
Productivity (t/m ³ d)	1.6–8.1	[100,108,112,113,116,135,140,141,144]
Technology Readiness Level (TRL)	2–6	[7,113,131,135,141,144,145]
Cost of retrofitting (M€)	900	[102]
Pay-back (y)	20–30	[146]
Change in performance with respect to air-blown BF		
Increase in O ₂ production (Nm ³ /thm)	150–280	[106,109,116,119]
Decrease in coke consumption (kg/thm)	14–150	[11,108,110,112,114,127,128]
Decrease in CO ₂ emissions (kg/thm)		
Due to recycling	100–500	[100,106,108,112,114,119,129,130]
Due to permanent storage	700–770	[105,106,108,119,131]
Increase in electricity consumption (GJ/thm)		
Due to ASU	0.2–0.6	[106,109,116,119]
Due to carbon capture (VPSA)	0.5–0.6	[103,106]
Decrease in available thermal energy downstream (GJ/thm)		
Due to recycling itself	1.5–4.0	[100,108]
Due to heating the recycled gas	0.5	[106]
Due to carbon capture (amines)	2.1–4.1	[100,107]
Increase in productivity (%)		
Increase in CAPEX (incl. carbon capture) (M€/(thm/h))	25–75	[100,108,112,143,147]
Increase in OPEX (€/thm)	1.25	[146]
	76	[146]

some studies report coke consumptions as low as 159 kg/thm, experimentation in prototype-scale oxygen blast furnaces showed 200–230 kg/thm as the minimum feasible coke rate [111].

A direct consequence of this coke saving is the cut in CO₂ emissions. This decrease is inherent to the TGR-OBF technology, and it is in the range 100 to 500 kgCO₂/thm (typically, 10 % – 40 % reduction with respect to BF) [100,106,108,112,114,119,129,130]. On average, the CO₂ emission reduction found in the literature is around 20 % (Table 8). Additionally, as CO₂ is rejected from the recycled top gas via a capture stage, a non-negligible amount of highly-concentrated CO₂ gas is available for underground storage. Considering that this CO₂ is sent to permanent storage, the total CO₂ emissions reduction can be as high as 800–1270 kg/thm (i.e., up to 86 % reduction) [105,106,108,119,131]. It is worth to mention that all the CO₂ emissions presented in this section are calculated after complete combustion of the gas (i.e., the CO content is accounted as equivalent CO₂).

An alternative option to take advantage of the captured CO₂ is to combine OBF with Power to Gas (PtG) [121]. PtG technology consumes renewable electricity to produce H₂ via water electrolysis, which is then combined with the CO₂ emissions of the ironmaking process to obtain synthetic methane [132,133]. This synthetic fuel is used in the blast furnace to keep carbon in a closed loop and avoid geological storage [120,134]. The amount of CO₂ that can be avoided through this method

is limited by the mass flow of natural gas that can be injected in the blast furnace, which depends on the flame temperature. About 70–90 kg/thm of CO₂ could be kept in closed loop and prevented from being permanently stored, which represents 5–7 % of the emissions of a conventional BF. However, it implies additional electricity penalizations of 1.3–2.1 GJ/thm, because of the production of renewable H₂ [121].

5.3. Operating parameters in OBF

In addition to the process mass flows (Table 5), the operation of a blast furnace is characterized by the percentage of direct reduction, the flame temperature, the heat losses, the temperature of the thermal reserve zone, the chemical efficiency, and the gas utilization (Table 6). All of them present some differences between oxygen blast furnaces and air-blown blast furnaces.

The percentage of direct reduction is between 4 % and 19 % in OBF (in air-blown BF is between 20 % and 40 %) [103,109,110,129,135,136]. The higher the O₂ concentration in the blast, the lower the direct reduction. To a lesser extent, increasing the volume of recycled gas also diminishes the percentage of direct reduction [103]. Moreover, for each Nm³ of recycled top gas injected at the tuyeres, the flame temperature will be reduced by 1.4–5.0 °C, depending on the temperature of the injected gas [108,110,114]. The amount of gas

Table 7
Physical dimensions of experimental oxygen blast furnaces [7,113,131,135,141,144,145].

Institution	JFE Steel	Minmetals	LKAB	Baowu	Tulachermet
Height (m)	5.1	n/a	6.0	n/a	n/a
Hearth diameter (m)	0.95	1.0	1.4	n/a	n/a
Throat diameter (m)	0.7	n/a	1.0	n/a	n/a
Inner volume (m ³)	3.9	8.0	8.2	430	1033
Productivity (t/m ³ d)	5.1	5.4–8.1	4.4	n/a	1.6

injected will be limited by the minimum adiabatic flame temperature (AFT) that allows keeping the proper operation of the furnace. The AFT assumed in the OBF studies found in literature is between 1900 and 2800 °C [100,108,137]. This range is actually taken from conventional BF experiences, whose minimum AFT is traditionally assumed at 2000 °C [97]. However, in practice, the lower the percentage of direct reduction, the lower the minimum limit for the flame temperature can be [138]. In fact, some actual blast furnaces have been successfully operated with 1600–1900 °C flame temperatures [113,138]. Since OBFs present remarkably lower percentages of direct reduction than BFs, 1800 °C could be considered as a more realistic technical limit for the flame temperature in OBF [113]. Heat losses for OBF reported in literature are in the range 230–630 MJ/thm [100,108,110,111,127,137,139], which are lower than for BF thanks to the higher productivity and smaller size of the furnace [109]. The temperature of the thermal reserve zone in OBF (700–950 °C) is usually assumed lower than for conventional BF (850–1000 °C), in literature [100,103,108,127,140]. The chemical efficiency used for OBF in the literature is between 0.94 and 0.98 [100], which is higher than that for conventional BF (0.85–0.95) [114] thanks to the downsized hearth diameter, what makes gas channeling effect less relevant [109]. Lastly, the gas utilization for OBF is in the range 47.7 % to 54.3 %, for the studies reviewed [109,112].

5.4. Size and productivity

The productivity of a blast furnace is usually restricted by the flooding phenomena, which limits the downward mass flow of a liquid in a bed of solid particles with countercurrent gases. The liquid will stop flowing downwards (or will even rise upwards as reflux), when the gas exerts a limiting force in the opposite direction to the descent of the liquid [148]. Since the OBF presents lower bosh gas flows than conventional BFs, the flooding limitation is less restrictive. In the literature, productivities in the range of 1.6 to 8.1 t/m³d are reported (thm per m³ of inner volume per day) [100,108,112,113,116,135,140,141,144]. In general, the productivity of the OBF is 25 % to 75 % greater than those for conventional BF, what allows to downsize the furnace while keeping the same production rate [100,108,112,143,147]. Conventional BFs have 5000 m³ inner volume, so OBFs are expected to have around 2500–2800 m³ inner volume [103,109,140]. Despite of the smaller diameter of the shaft, analysis based on discrete element method showed that the compressive stress on the solids is similar or even lower than in conventional BF [109].

5.5. Technology readiness level, and costs

The TGR-OBF technology is not yet commercial. Most of studies found in the literature are TRL 2, and only a few articles mention experimental tests between TRL 3 and TRL 6 (Table 7; note that commercial OBF should have 2500 m³ inner volume). Institutions that have performed experimental research on OBF include (in the ascending

order of size) JFE Steel [7,141], the State Key Laboratory of Advanced Steel Processes and Products of China in cooperation with Minmetals Yingkou Medium Plate [135,144], the ULCOS project at LKAB's experimental blast furnace [7], the Baowu Group [135], and Tulachermet [113,135].

The viability of future research and investments in OBF is ruled by the cost of producing 1 ton of steel [146]. The price of steel (without accounting CO₂ taxes) was around 500 €/t during the last decade, but it rose to 1400–1800 €/t last year due to the global scenario [149]. According to the studies based on pre-pandemic scenarios, the price of steel under OBF technology equals the price of conventional steel when the CO₂ taxes are in the range 40–150 €/tCO₂, depending on the price of electricity and the carbon capture technology (VPSA or amine scrubbing) [102,104,118,146,150]. It should be noted that the electricity price and the CO₂ taxes are correlated, and therefore, it is recommended to consider that electricity price increases by 0.85 €/MWh per each 1 €/tCO₂ increase in the CO₂ price [146].

Retrofitting BF to OBF is barely considered in literature [102], since the required size of the blast furnace decreases, and O₂-enrichment above 30 % causes equipment damage in hot air blowing systems (oxidation of metallic pipes, lances, etc.) [100]. Nevertheless, the cost of retrofitting the furnace and piping is estimated at approximately 900 M€ (cost of the carbon capture plant not included) [102]. The investment cost for a new standalone OBF may be smaller than that for a new conventional BF because of its smaller dimensions. However, it implies additional costs related to the larger ASU, compressors, and carbon capture stage. Hence, the overall investment for OBF may be 1.25 M€ greater than that for conventional BF, per thm/h (including carbon capture) [146]. Pay-back on blast furnace investments is assumed to be between 20 and 30 years in the literature [146].

Regarding operating costs, the major differences between the OBF and BF are the amount of coke consumed (up to 150 kg/thm less) and the electricity consumption (up to 4.2 GJ/thm more). Typical costs for coke and electricity are 300 €/t and 80 €/MWh, respectively [146]. Thus, the OPEX would decrease by up to 45 €/thm because of coke savings, but increase by up to 93 €/thm due to the electricity consumption (extreme values). Moreover, the transportation and permanent storage of CO₂ can be assumed as an additional cost of 27 \$/tCO₂ [146] what translates into 28 €/thm. In total, the net increase in OPEX is 76 €/thm. At the current price of steel, this represents a 5 % increment in the price of steel (one decade ago, it would correspond to a 15 % increment in the price).

6. Pre-combustion CC in BF-BOF

Pre-combustion technologies for carbon capture in the I&S industry can be applied to the CO₂ emissions from raw materials preparation, blast furnace, and blast oxygen furnace. The resulting gas stream after CC can be used as fuel in internal combustion processes. The design and integration of chemical absorption, physical adsorption, membrane separation, and sorption enhanced water gas shift (SEWGS) technology

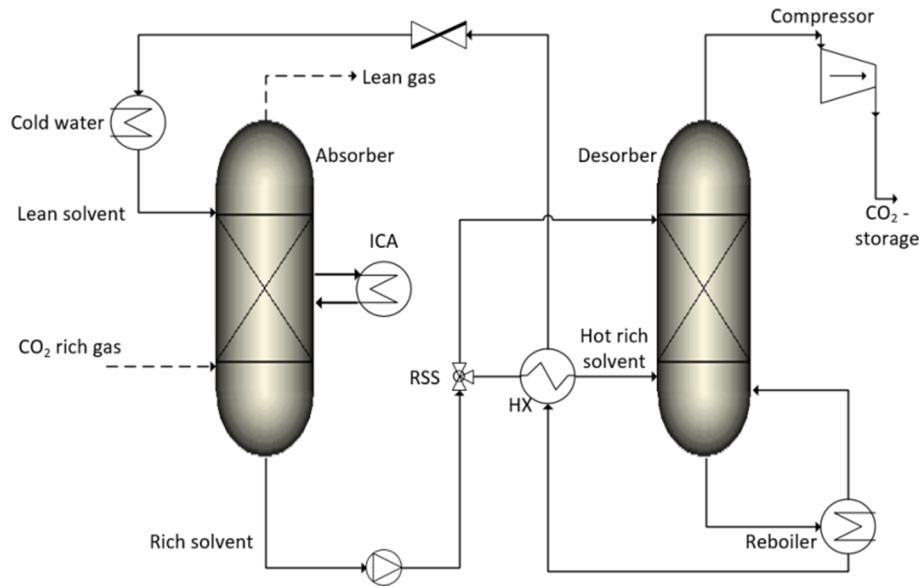


Fig. 10. CO₂ absorption with intercooled absorber (ICA) and rich solvent splitting (RSS), . adapted from [30]

for pre-combustion CC in the BF-BOF route are reviewed in this section.

6.1. Pre-combustion with chemical absorption

Certain chemical solvents, such as amine-based (presented in Section 4.1) or aqueous ammonia, can be used for pre-combustion carbon capture. Tobiesen et al. [151] presents one of the early studies based on modelling CO₂ absorption from the off-gases of conventional and oxygen-based blast furnaces using monoethanolamine (MEA), methyl-diethanolamine (MDEA)/piperazine (Pz), and 2-amino-2-methyl-1-propanol (AMP) as primary, tertiary/secondary mixture, and sterically-hindered primary amine solvents, respectively. It is unclear why secondary amines were not tested despite both primary and secondary amines undergo hydrolysis and carbamate mechanisms for CO₂ absorption, as shown in reactions (1) to (7), while tertiary amines only

undergo slow hydrolysis reaction with CO₂ [152].

Hydrolysis mechanism:

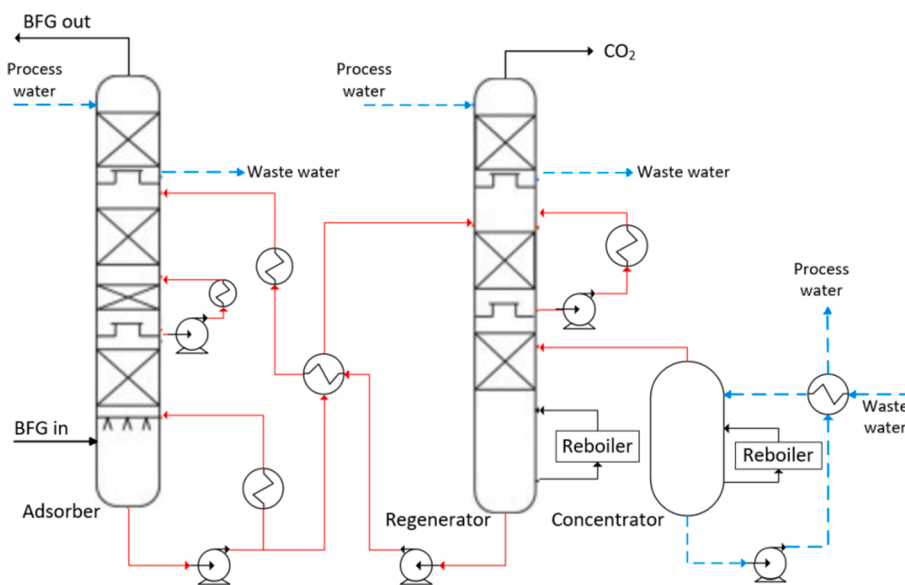
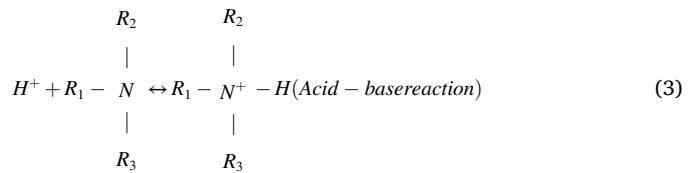
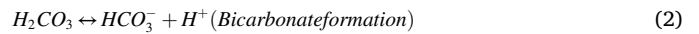
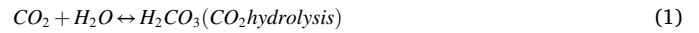
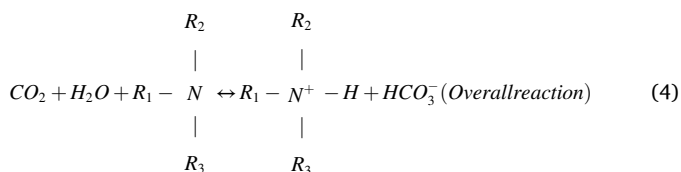
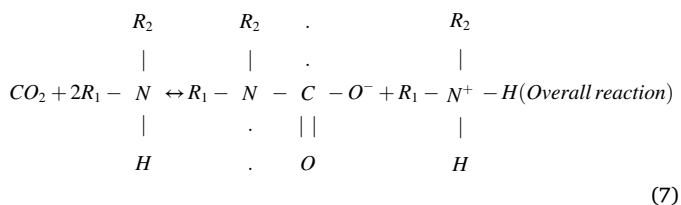
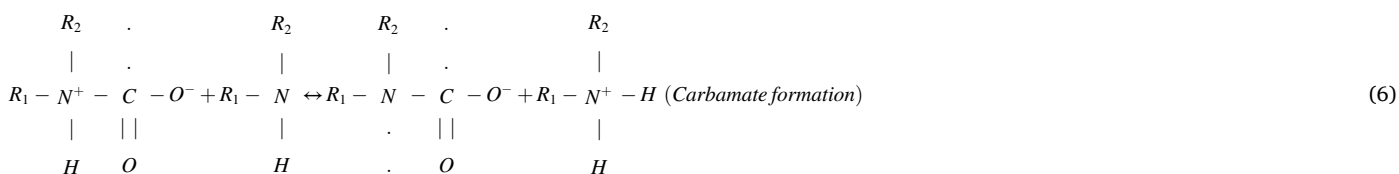
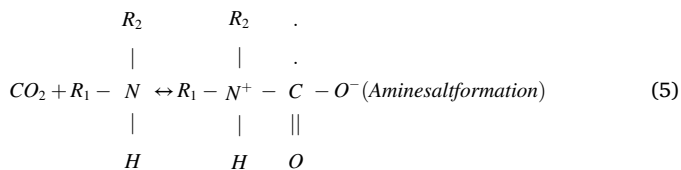


Fig. 11. Aqueous ammonia-based CO₂ capture from the blast furnace off-gas (BFG) . adapted from [155]



Carbamate mechanism:



Moreover, AMP also does not favour fast CO₂ absorption, as it hinders carbamate reaction. Since absorption process has high efficiency at low temperature and high pressure, CO₂ capture typically involves four steps: (a) gas pretreatment to reduce temperature and increase pressure through compression, (b) gas absorption with intercooling using alkaline solvent, (c) solvent regeneration by separating CO₂, and (d) CO₂ compression and drying for transportation. According to the process simulations [151], about 80–90 % of the total energy is needed for steam generation for desorber reboiler, while the remaining 10–20 % drives the gas compressors, blower, and the solvent recirculation pumps. Based on the regeneration energy requirement and the solvent recirculation rate, AMP was the most suitable solvent (reboiler steam consumption: 2.6 MJ/kg_{CO2} with AMP, 2.9 MJ/kg_{CO2} with MDEA/Pz, and 3.7 MJ/kg_{CO2} with MEA), although it would require a large absorber size due to its slow reaction with CO₂.

To enhance process efficiency and energy savings, heat recovered from BFG or from flue gas of power plant can cover the thermal needs of the amine capture system [30,153]. As it is shown in Fig. 10, an inter-cooled absorber (ICA) and rich solvent splitting (RSS) was included in the simulation model. The maximum CO₂ capture from BFG was 5 % lower than that from flue gas stream due to the lower CO₂ content in the former. The effect of feed split ratio for minimising solvent regeneration energy and the implications of such designs on the capital cost were not investigated.

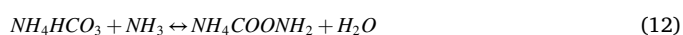
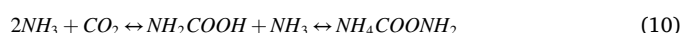
Different published works [19,21,154] present overviews of various CO₂ breakthrough programs, including ultra-low carbon dioxide steel making (ULCOS-2004) program in the EU (also discussed in [11]), CO₂

Ultimate Reduction in Steelmaking Process for Cool Earth 50 (COURSE50-2008) in Japan, Pohang Iron and Steel Company (POSCO-2020) program in South Korea, and American Iron and Steel Institute (AISI-2011) program in USA. The ULCOS program focused on the amine scrubbing using MDEA, though it suffers from high energy requirements (about 3.2 GJ/tCO₂) and further research is needed to reduce the CC energy penalties [11]. The COURSE50 program investigated CO₂ absorption with amines or amines mixed with alcohols/additives. Under this program, Onoda et al. [107] studied 21 solvents such as MEA, MDEA, 2-amino-2-methyl-1-propanol (AMP), 2-Piperidineethanol (2-PPE), diethanolamine (DEA), 2-(ethylamino) ethanol (EAE), N-isopropyl-diethanolamine (IPDEA), and 4-piperidineethanol (4PPE). The temperature of adsorbent regeneration was optimised through two pilot plant tests to reduce the CC energy penalty from about 4 to 2 GJ/tCO₂. However, further studies are needed to understand their degradation and vapourisation losses in the presence of gas contaminants and high temperature over an extended period of time.

The POSCO program recommends aqueous ammonia as solvent for

absorption process (see Eq. (8)-(12)). A pilot-scale ammonia-based absorption process for CO₂ capture was integrated in an actual I&S plant (Fig. 11) [155].

Aqueous ammonia captures CO₂ through the following reactions:



To avoid ammonia loss with the gas exiting the absorber, the treated gas is water-washed. The CO₂-loaded solution is regenerated using a reboiler (maintained near 80 °C). The CO₂-rich stream from the regenerator is also water-washed to remove ammonia escaping with it. The regenerated ammonia solution is sent to the absorber for its reuse. The regeneration of washing water takes place in the concentrator, where water is separated from ammonia. Through the comparison of CO₂ absorption by amines, ammonia, and K₂CO₃, it was argued that ammonia would require lowest regeneration energy and absorbent cost with less risk of corrosion and solvent thermal degradation. With a blast furnace off-gas flow rate of 1000 Nm³/h, the process was shown to absorb over 95 % CO₂ (about 10 t_{CO2}/day), and produced a gas stream from the regenerator with nearly 99 % CO₂. However, an economic analysis and the feasibility of a full-scale plant implementation (with high off-gas flow rate) should be evaluated since the slow rate of CO₂ absorption by ammonia may require a significantly large column size and a high solvent recirculation rate.

Ho et al. [26] studied the retrofitting of existing I&S plants in Australia with CO₂ capture using MEA for CO₂ absorption. Feed pretreatment involved selective catalytic reduction (SCR) to reduce NO_x concentration below 20 ppm, particulate filter to remove solids, and flue gas desulfurization (FGD) to reduce SO_x concentration below 20 ppm.

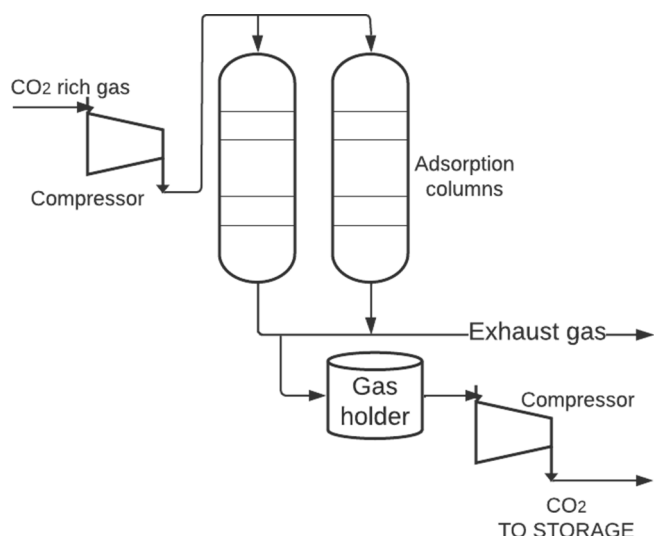


Fig. 12. Process flow diagram of a simple PSA technology, . adapted from [158]

For off-gas with low CO₂ partial pressure, water-gas shift reactor was considered to convert CO to CO₂, and thereafter, Selexol (physical solvent) was used for CO₂ absorption [26]. For MEA-based CO₂ capture, the capital and operating cost represented 20–30 % and 70–80 % of the total cost, respectively. Overall, the costs of CO₂ capture were A\$74/tCO₂ for the blast furnace and A\$56/tCO₂ for the Corex process. The CO₂ capture cost for the Corex process could be further reduced to A\$39/tCO₂, if the inlet gas was compressed to about 20 bar and the water-gas shift reactor was used to enrich the gas with CO₂ so that a physical solvent (Selexol) could be used. Ho et al. [102] extended study [26] by investigating CO₂ capture from Hismelt, Midrex, and mini mill processes. To be cost-effective, it was recommended to capture CO₂ only from main emission sources (for example, from onsite power plant with cost estimate of A\$76/tCO₂, coke ovens and stoves with a cost of A\$80/tCO₂, and sinter plant with a cost of A\$90/tCO₂) that can achieve CO₂ emission reduction by about 80 % from an I&S industry. Moreover, the use of vacuum pressure swing adsorption (VPSA) with concentrated gas was found to involve lower cost of CO₂ capture than MEA absorption due to lower regeneration energy requirements.

Some studies in the literature provide the comparison of different techniques for CO₂ capture. Quadar et al. [156] compared the efficiency and energy requirements of some CO₂ capture studies from the literature from blast furnace/stove off-gas using aqueous ammonia [155], MEA [26,38], and MDEA/MEA [157], and VPSA [106]. Since the efficiency of

CO₂ capture and energy requirements depend on the gas composition, temperature, pressure, and flow rate, a direct comparison of the results from studies utilizing different feed parameters is difficult. Quadar et al. [156] noted that both absorption and adsorption are suitable for CO₂ capture. Kim et al. [158] simulated PSA and MEA-based absorption processes for CO₂ capture from furnace off-gas. The MEA-based absorption process suffered from solvent degradation and loss and equipment corrosion. The absorption process led to a CO₂ stream with higher purity than the PSA process with a difference of 9.95 mol%, but the energy required in the absorption process was 224.78 MW higher than that for PSA. While PSA required lower regeneration energy than the absorption process, its adsorption efficiency was high only when CO₂ was present in high concentrations. Gazzani et al. [29] studied CO₂ capture from an iron & steel plant in the pre-combustion and post-combustion stages using MEA and MDEA solvents, and with sorption-enhanced-water-gas-shift (SEWGS) reactors. The results were presented in terms of specific primary energy consumption for CO₂ avoided (SPECCA, MJ/kgCO₂) versus % of CO₂ avoided. The pre-combustion CO₂ capture process involved compressing the off-gas, passing it through a water saturator, sending the resultant gas to water-gas shift reactors, and finally, using MDEA-based absorption unit to capture CO₂ and send the H₂-rich gas to the onsite power plant for combustion. This process, with heat integration, achieved a CO₂ avoidance of 89 % with a SPECCA of about 3 MJ/kgCO₂. In comparison, the processes requiring passing the off-gas from the iron & steel plant to a water saturator and water-gas shift reactors, and thereafter, capturing CO₂ using adsorbents and sending the H₂-rich gas to the onsite power plant, achieved a CO₂ avoidance of 78–89 % with a low SPECCA of about 2.2–2.95 MJ/kgCO₂.

Onarheim et al. [159] compared the processes involving CO₂ capture using a chemical (MEA) and a physical (Selexol) solvent. A process with a blast furnace using oxygen-enriched air (40 % O₂) and a basic oxygen furnace was simulated, where the off-gas was passed through a water-gas shift reactor. In comparison to the reference base process (without CCS) with CO₂ emission of 81.6 kg/s, the absorption processes with MEA and Selexol reduced the CO₂ emissions to 34.7 and 24.8 kJ/mol, respectively. The study did not present an economic analysis to understand if the cost associated with physical absorption was lower than that with MEA, as they involved different process equipment.

Chung et al. [160] studied CO₂ capture from the blast furnace off-gas through amine (piperazine) absorption, membrane separation, and hybrids of absorption and membrane separation. A hybrid process required lowest utility and capital costs with operating cost change of \$77 M/year, capital cost of \$66 M, highest rate of CO₂ avoided (69.4 %), and lowest CO₂ avoidance cost (\$30.4/tCO₂). Their calculations suggest that the membrane separation unit has a better rate of CO₂ avoided (66.4 %) and a lower CO₂ avoidance cost (\$34.4/tCO₂) as compared to the amine

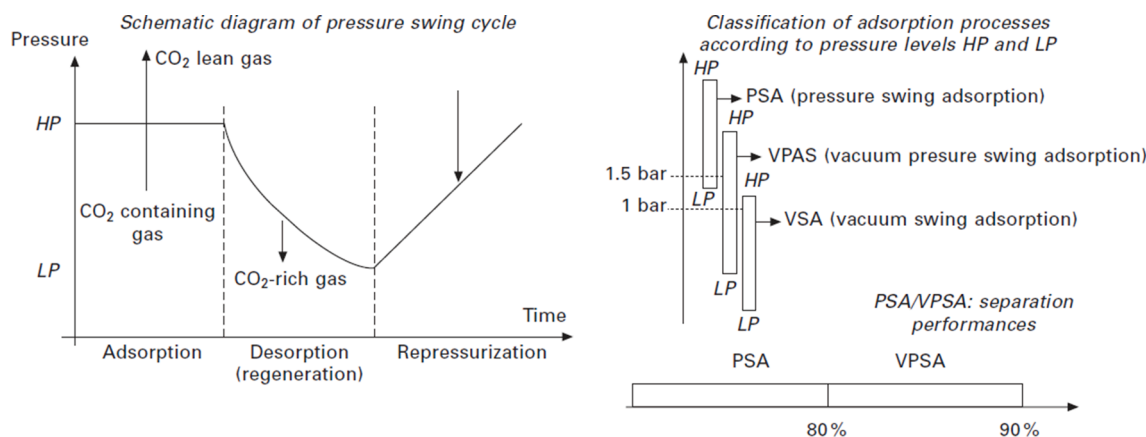


Fig. 13. Principle of the PSA CO₂-scrubbing techniques (left) and various domains of application and performances of the variant techniques, PSA, VPSA and VSA . adapted from [163]

absorption unit (with their respective values of 61.3 % and \$40.6/tCO₂). However, only one amine (piperazine) was tested to reach this conclusion.

The above studies indicate that CO₂ absorption by amines may be a well-established technique, but VPSA is more suitable for iron & steel plants with less energy demand and lower cost of CO₂ avoidance. This has motivated several adsorption studies, such as by Arasto et al. [106] and Tsupari et al. [146], on CO₂ capture from the oxygen blast furnace, and by Hisashige et al. [161] on CO₂ capture from the smelter off-gas, as discussed in Section 6.2.

The technologies for CO₂ absorption using chemical solvents (e.g., MEA, MDEA) are widely deployed in industries (e.g., for removal of H₂S and CO₂ from sour gas in natural gas processing facilities), and thus exhibits high TRL (9). However, their application and demonstration to carbon capture from BFG, BOFG, and COG streams at industrial scale has been limited to date. Thus the majority of pre-combustion-based CC technologies reviewed in Section 6.1 are found to be of low TRL (2–4), and to have been investigated solely through simulation rather than experiments, with the exception of a few pilot scale studies conducted as part of either COURSE50, ULCOS, and STEPWISE projects [87,111,144,148,157] for technologies (carbon capture using various amines) or in an independent study [134] for technology (carbon capture using ammonia).

6.2. Pre-combustion CC with adsorption processes

In physical adsorption, the main interacting force is the Van der Waals force. In this system, there is no chemical reaction between the gases and the adsorbents, as the gas is physically bonded to the surface of the solid. There are four different ways of regenerating the adsorbent: pressure swing adsorption (PSA), vacuum-pressure swing adsorption (VPSA), temperature swing adsorption (TSA) and electrical swing adsorption (ESA).

A key finding from the ULCOS project indicated that the chemisorption methods such as amine scrubbing, physisorption technologies such as VPSA or PSA, and cryogenics have different domains of optimality, with the concentration of CO₂ in the gas stream to be treated being one of the most critical determining factors [162]. The physisorption schemes are the best in terms of technical performance and cost, both operational and capital, at the concentration levels reported in the BF instance and the TGR–BF scenarios [163]. The technologies for CO₂ capture and separation in the pre-combustion stage with adsorption can be categorized in terms of the adsorbent type (such as alumina, zeolite & MOFs, and activated carbon), regeneration method (such as pressure swing, temperature swing, electric swing, and washing), and type of monolith molecular sieve (e.g., carbon coated substrate and carbon/carbon fiber monolith) [101]. Among these, pressure swing adsorption (PSA)-based techniques have been the most advanced adsorption techniques reported till date. A typical PSA technology involves two stages that could be either operated as a single stage (as shown in Fig. 12) or as dual stage, wherein CO and N₂ are first separated from CO₂ in the first unit packed with a CO₂ adsorbent, such as Zeolite 13X [158] or Zeolom F-9 [164]. Thereafter, the off-gas from the first PSA, consisting mainly of N₂ and CO, is introduced to the second unit which accomplishes the separation of CO and N₂ using a CO adsorbent [165]. Under plant conditions, the total energy required to recover about 90 % CO₂ with a purity of 91 % is ca. 36.9 MW [158]. In addition, a highly-concentrated combustible gas (with 72 mol% of CO and H₂) is obtained as an exhaust gas byproduct that could be utilized in power generation boilers [158]. Fig. 13.

Although high CO₂ recoveries (>90 %) have been reported [158], the purity of CO₂ in the recovered stream is relatively low with PSA technology compared to other CO₂ separation processes. Rao et al. [166] employed a PSA for recovering CO₂ from coke oven gas. The final CO₂ recovered stream contained as much as 24 % moisture, making it unsuitable for compression and storage without further purification. To

overcome this demerit, advances in PSA technology have resulted in a combination of other CO₂ mitigation or separation processes with PSA [167]. Another similar technique to PSA is the Vacuum Pressure Swing Adsorption (VPSA) wherein adsorption is carried out at ca. 1.5 bar and desorption is performed under vacuum conditions at 0.05 bar. As the exhaust gas from BF contains moderately high partial pressure of CO₂ (>0.2 bar), VPSA has been shown to be more economical in reducing the CO₂ capture costs [168]. A detailed economic analysis by Ho et al. [102] has shown that VPSA has the potential to reduce CO₂ capture costs by 25–40 % compared to MEA solvent absorption technology.

Two of the key concepts of a top gas recycling blast furnace (TGR-BF) are the removal of CO₂ from the top gas to enhance recycling and injection of oxygen into the furnace in the place of air [169]. In general, the oxygen blast furnace has the benefits of substantially lowering the CO₂ emissions compared to the traditional blast furnace [135]. The net CO₂ emissions from an integrated steel mill combined with a TGR-OBF is about 1.17 t/t crude steel compared to 1.75 t/t crude steel for a process without CCS. Arasto et al. [106] evaluated a VPSA-based CO₂ capture process configuration from the oxygen blast furnace, based on an existing steel facility in Finland. CO₂ emissions are estimated to be reduced between 1.2 and 1.4 Mt/year (from over 4 Mt/year) when CCS is applied to the oxygen blast furnace with and without CO₂ storage, respectively.

TGR-OBF equipped with VPSA for CO₂ capture has also been beneficial for reducing energy consumption of an integrated steel mill (ISM). This technology significantly decreases generation of metallurgy gas in the coking and ironmaking processes (by ca. 53.4 %) which might consequently decrease the electricity output of the power plant. Moreover, the electricity demand of the ISM is ca. 60 % higher than that of the conventional process due to increased consumption by VPSA and oxygen generation plants [103,119]. Thus, additional electricity may have to be generated or purchased for the ISM with TGR-OBF compared with conventional ISM. On the other hand, the energy input to the ISM with TGR-OBF process decreases substantially from 17.89 to 14.67 GJ/t owing to decreased coking coal consumption. It has been demonstrated in the literature that compared to a traditional air blast furnace equipped with TGR, the utilization of a lean grade burden with pulverized coal injection (PCI) and high blast oxygen enrichment (medium, 50 %; enriched 98 %) can curtail carbon consumptions by 14.1 % and 20.2 %, respectively [110,135]. Nevertheless, various strategies for overcoming issues related to low gas volume have been addressed in the literature [130]. Sen [142] proposed a CO₂ accounting procedure based on a carbon balance, which can facilitate evaluation of the recovery of surplus gases to reduce emissions. Carbon capture using VPSA with top exhaust gas recycling was identified as a promising emissions reduction approach. The analysis based on an oxygen blast furnace with top gas recycle configuration developed by ULCOS indicated ca. 90 % reduction in CO₂ content with a CO₂ purity of 87.7 % in the recovered stream. On a commercial scale, VPSA has been successfully implemented in combination with top gas recycling of blast furnace effluent gases [11,111]. While mitigation measures resulted in a 24 % CO₂ emission reduction, nearly 52 % reduction was accomplished with the VPSA technology [11,102]. However, an unsteady state analysis of the TGR-OBF blast furnace have indicated a slightly lower CO₂ emissions reduction (ca. 69 %), and the deviations have been attributed to the differences in CO₂ emission calculation method [128]. Nevertheless, the amount of treated gas recycled into the blast furnace determines the CO₂ reduction levels in the system. For example, at a gas injection rate of 600 Nm³/thm, the level of CO₂ emission reduction that could be attained is ca. 1000 kg/thm. However, when the feed rate is increased to 1100 Nm³/thm, the reduction levels increase to about 1200 kg/thm [131].

In addition, the coupling of a cryogenic system with the VPSA results in further purification to attain a higher level of purity (ca. 96.2 % on dry basis) in CO₂ stream, thus making it favourable for transport and storage. In comparison, the level of CO₂ purity attained with PSA and VPSA is about 79.7 % and 87.2 %, respectively. However, the energy

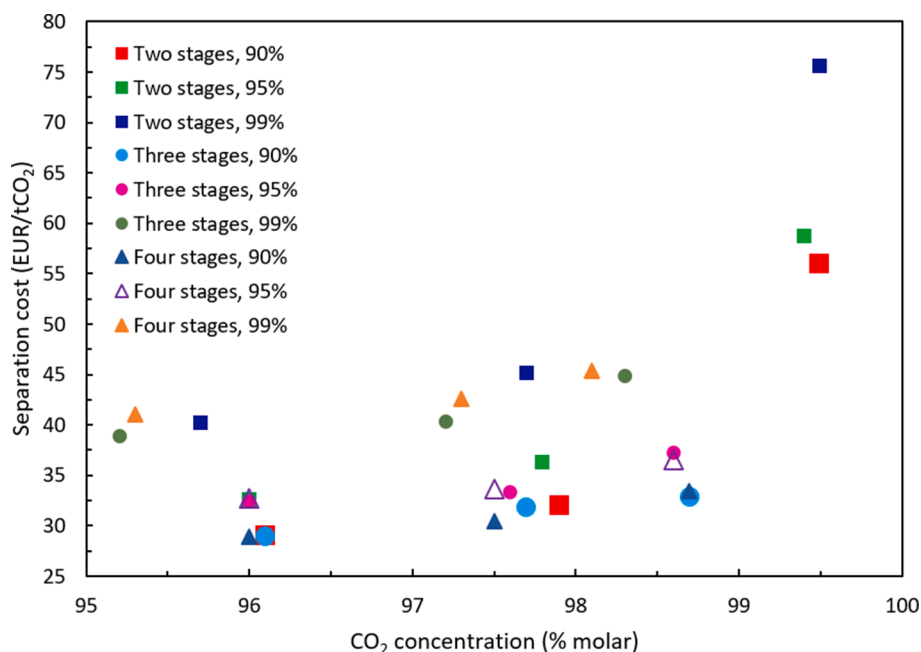


Fig. 14. Separation cost vs CO₂ product purity for the optimal configurations with two, three and four stages (% values in the legend refer to CO₂ recovery). Data from left to right on each series correspond to N₂ residual contents of 1%, 0.5% and 0.1% respectively, . adapted from [162]

consumption for the combined VPSA and cryogenic technique is ca. 2.9 times higher than those for PSA or VPSA systems that could be directly attributed to the energy requirements for CO₂ compression [11]. Another interesting option is the combination of PSA with cryogenic distillation compression that is capable of delivering a 99–100 % pure CO₂ stream [106], with a total energy consumption of 1.12 GJ/tCO₂ compared to 1.05 GJ/tCO₂ for a VPSA + compression and cryogenic system.

An in-situ method for capturing CO₂ by adsorption during the process of COG gasification was proposed by Wang et al. [170]. By integrating the sub-processes of SMR, WGS, and carbonation in an adsorption enhanced hydrogen amplification reactor (AEHAR), CO₂ was captured using calcium oxide packed in the reactor and later released during the regeneration process. In order to remove other desorbed gases from the recovered CO₂ stream, a hot potassium carbonate process was applied in the downstream section. Since this study focused on maximizing hydrogen recovery, the details on CO₂ recovery and efficiencies were not reported.

Generally, higher adsorption pressure results in a higher recovery ratio, but also in higher blower power consumption. Simultaneously, a lower desorption pressure results in a larger recovery ratio and higher vacuum pump power usage. In an attempt to minimize the recovery cost, Saima et al. [164] experimentally investigated the effects of cycle time and CO₂ concentration in the exhaust blast furnace gas on the efficiency

of pressure swing adsorption (PSA) to extract CO and CO₂. Increasing the CO₂ feed concentration from 22 % to 33 % increases CO₂ recovery by ca. 1.6 times. However, the effect of process cycle time on CO₂ recovery levels was found to be negligible and an optimum time of 225 s was recommended. These findings were further demonstrated at a bench scale setup that recovered about 6.3 tCO₂/day resulting in an overall reduction in capture cost by 63 % [164].

The primary carbon capture technology by adsorption relies on the pressure swing adsorption or vacuum pressure swing adsorption technique. The adsorption process using solid sorbents such as Selexol, alumina, zeolites, and activated carbon, has been demonstrated for several industrial applications (TRL = 9), including hydrogen production in refineries, nitrogen production, and dehydration applications [171]. Most of the studies listed in section 6.2 employ various sorbents in PSA/VPSA with top gas recycling for carbon capture from steel plant off-gases. Nevertheless, since these studies are simulation-based without actual pilot- or plant-scale demonstration, their TRL has been characterized as low (2).

6.3. Pre-combustion with membranes

Hasan et al. [172] found that membrane separation is preferable to amine scrubbing when CO₂ concentration in the flue gas is higher than 36 %, which is true for blast furnace off-gas. A wide variety of membrane

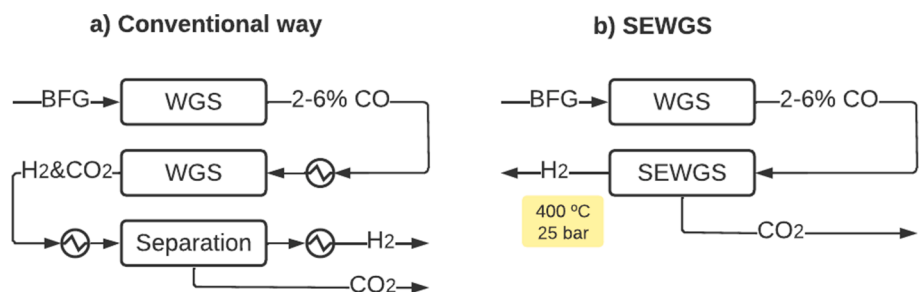


Fig. 15. Process flow diagram of a) conventional WGS and b) SEWGS technology, . adapted from [179]

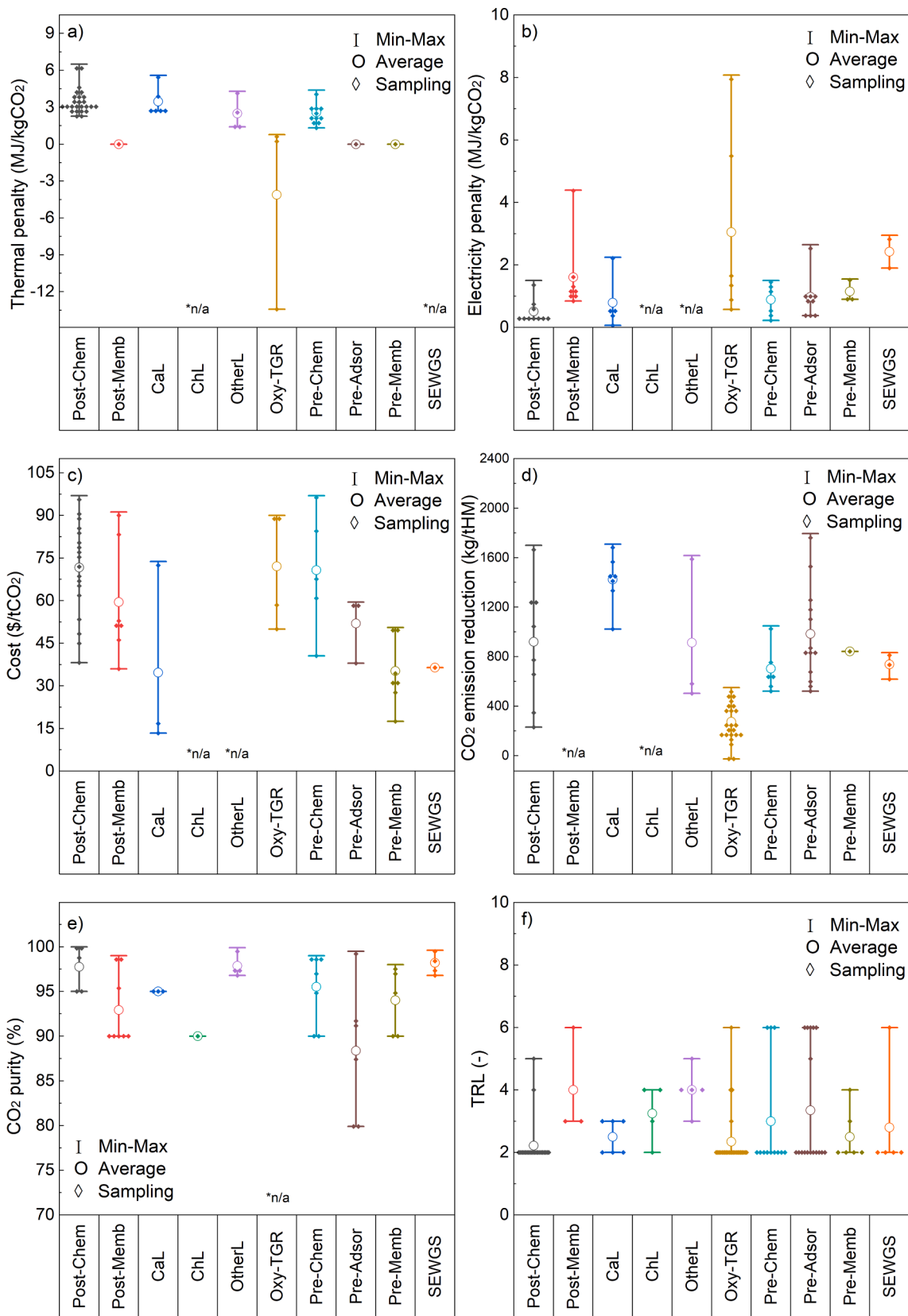


Fig. 16. KPI comparison for different carbon capture technologies: a) thermal penalty (MJ/tHM), b) electricity penalty (MJ/tHM) c) cost (\$/tCO₂), d) CO₂ emission reduction (kg/tHM), e) CO₂ purity (%) and f) TRL (-). Nomenclature: Post-Chem is postcombustion carbon capture (CC) by chemical absorption; Post-Memb is postcombustion CC by membranes; CaL is calcium looping; ChL is chemical looping; OthL is other looping processes; Oxy-TGR is oxy-blast furnace with top gas recycling (TGR); Pre-Chem is precombustion CC by chemical absorption; Pre-Adsor is precombustion CC by adsorption; Pre-Memb is precombustion CC by membranes and SEWGS is sorption enhanced water gas shift.

materials including polymers [160,162,173–175] and zeolites [176] have been employed for CO₂ recovery from various industrial waste gases, among which polymeric membranes are largely preferred for coke oven gas and blast furnace gas. Lie et al. [174] compared the performance of a fixed site carrier membrane (FSCM), comprised of amine groups in the polymeric backbone, with carbon based membranes. The results showed that aminated polymeric membranes exhibited enhanced permeability and selectivity in favour of CO₂. Moreover, while the presence of moisture in the blast furnace top gas stream is detrimental for carbon-based membranes, it is beneficial for FSCM as the active amine groups act as carriers for CO₂-water complex. Membrane Technology Research (MTR)'s Polaris© membrane has been the most widely employed one in lab and pilot-scale CO₂ capture studies as it possesses the highest CO₂ Gas permeance unit (GPU) of 1000, CO₂-N₂ selectivity of 50, and CO₂-CO selectivity of more than 67 measured at 10 °C [177] in addition to higher stability. Such improvements in membrane selectivity have enabled development of novel CO₂ capture concepts in sub-ambient conditions using membrane-hybrid cryogenic processes. However, simulation studies have demonstrated that such hybrid techniques are economically suitable only for high-CO₂ content flue gas, and not significant for lean CO₂ streams like BF and COG [178].

CCU from BF or BOF process off-gas by polymeric membranes typically requires multi-staged operation to meet the desired CO₂ recovery and purity levels, owing to the inherent limitations of the solution-diffusion separation mechanism [162]. An advanced superstructure design with unrestricted stream connections and interlooping proposed by Santos et al (2018) [162] demonstrates the feasibility to minimise overall CO₂ recovery cost by optimising numerous membrane architectures parameters, including number of stages, downstream pressure, membrane surface area, and stream connections.

The effect of CO₂ recovery and residual N₂ constraints on separation cost has been reported in a few studies [160,162,173–175]. The desired CO₂ product purity is significantly higher for MTR's Polaris membrane due to its high CO₂-N₂ selectivity. CO₂ concentration and recovery are also influenced sharply by the number of membrane stages considered in the study. In fact, even with a high CO₂/CO selectivity, a high level of CO₂ purity (>95 %) cannot be achieved by a single membrane unit. A three-staged membrane reactor setup connected in series could reduce CO concentration in the exit gas below 0.2 % [160]. Further, it is possible to achieve product streams with very low N₂ content (0.1–1 %) using a superstructure design as demonstrated by Santos et al 2018 [162]. Nevertheless, it has been shown that at fixed recoveries, decreasing N₂ residual content, say from 1 % to 0.1 % in the CO₂-rich product stream would increase separation cost, although the reverse (increasing recovery at constant purity) has a less pronounced effect [162].

It could be expected that increasing the number of membrane stages may increase the overall CO₂ separation cost. Based on a global optimisation procedure employing a non-linear programming formulation of the multistage superstructure design, the influence of CO₂ purity and number of membrane stages on separation costs was investigated and has been illustrated in Fig. 14 [162]. It can be noted that when high recovery and/or purity are required, three-stage setups were optimum in reducing separation costs and projected to be in the range of 32–50 \$/tCO₂. The third stage is utilised to collect CO₂ from the retentate of the first stage for CO₂ recoveries of 95–99 % and 0.5 % N₂ concentration. In the 90–99 % recovery range, a three-stage cascade was the optimum arrangement for a 0.1 % N₂ content. Separation costs were not significantly reduced in process designs with four membrane stages. Fig. 15.

The effect of effluent gas composition on CO₂ recovery and associated costs could be analyzed from the works of Lie et al. [174] and Chung et al. [160]. The variations in the furnace off-gas composition could result from the type of reducing gas (air or oxygen) used for combustion reactions. Generally, a decreased nitrogen content in the feed stream (oxygen as RG) improves the CO₂ flux through the membrane that results in increased recovery percentages. However, the

resulting increased CO₂ handling volume increases the membrane section duty by about 66 % relative to the conventional air-driven blast furnace [174]. On the other hand, considering the expander energy and other associated costs involved in the two processes, the difference in overall CO₂ recovery costs remains quite similar. The relatively higher recovery cost (ca. 34.4 \$/tCO₂) reported by Chung et al. in comparison to Lie et al. is due to the higher CO₂ purity levels considered in the product stream. In addition, the presence of impurities (NO_x, SO_x and H₂S) in the effluent stream tends to deteriorate the membrane performance with time. Sensitivity studies on the membrane permeability values and CO₂ selectivity ratio indicated that a 50 % reduction in overall membrane permeability resulted in a 10 % increase in separation cost, whereas a 50 % reduction in CO₂ selectivity resulted in a 34 % increase [160].

Among all membranes including polymers and zeolites, Membrane Technology Research (MTR)'s Polaris© membrane has been widely tested for carbon capture in lab and pilot-scale studies. Technology development for pre-combustion CC with membranes has mainly focused on the optimization of the number of stages and process configuration in a superstructure-staged membrane process. While the single-staged process has been validated at lab-scale (TRL = 4) [175], most of the multistage processes have been investigated using modelling and simulation (TRL = 2) [160,162,173,174]. The technology for other concepts including membrane process coupled with cryogenics [178] has also been demonstrated only in simulations.

6.4. Pre-combustion with SEWGS

Sorption enhanced water gas shift (SEWGS) technology combines pre-combustion carbon capture via adsorption and hydrogen production from syngas via WGS reaction in a single process capsule [179]. The catalyst used for the WGS reaction also serves to adsorb the CO₂ generated by the WGS reaction. Among various catalytic materials, potassium carbonate hydrotalcite-based materials are the most common [29,45,179] and technically advanced in terms of dual efficiency. While other commercial iron-based high-temperature catalysts have been tested for WGS reaction, CO₂ adsorptive performance has not been assessed [180]. However, regeneration of the sorbents is highly energy-intensive and thus necessitates efficient thermal integration techniques, like Ca-Cu chemical looping [93] and pressure release followed by steam purging using a PSA approach [29,32]. The demand for steam by the WGS reaction and SEWGS rinse/purge is responsible for the CO₂ capture energy penalty. Hence different configurations, such as expander layout, and saturator layout, focusing on heat integration between steam cycle and SEWGS have been investigated in the literature [32,45].

A common method to provide the SEWGS reactor with the precise wet gas composition and reduce steam purge during regeneration is incorporating a pre-shift WGS reactor prior to the SEWGS unit [45,179]. For the pre-shift reactor, a Johnson Matthey catalyst has been proven to reduce CO concentration in the BFG effluent stream from 20 % to 5 % [179]. It has been demonstrated that integrating SEWGS for both exhaust gas recycling and power generation section reduces overall CO₂ plant emissions by nearly 85 % compared to only 50 % for power generation section of steel industry [32]. In particular, the treatment of effluent gases (BFG, BOFG, and COG) using SEWGS could curtail CO₂ emissions by 80–90 % [29,32,93]. Furthermore, sorbent reusability studies estimated a 12 % decrease in CO₂ avoidance efficiency between fresh and thousands of cycles-old material [29], which is negligible considering the lifetime and cost of the material. In comparison to other amine-based technologies, SEWGS exhibits the highest energy efficiency (ca. 37.7 %) and lowest specific primary energy consumption ca. 2.2 MJ/kgCO₂) [29].

The technology for pre-combustion CC with SEWGS has been demonstrated in a pilot-scale facility (14 t/d CO₂ removal rate) using BFG as part of the STEPWISE project (TRL = 6) [157, 181]. However, investigations focused on either process modification to SEWGS or their

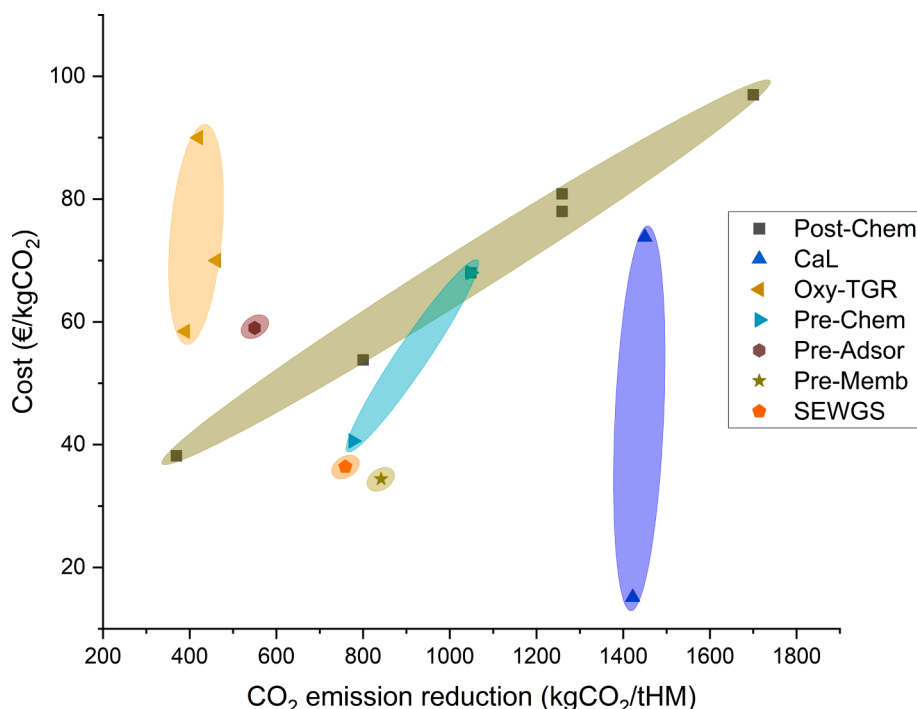


Fig. 17. Cost of CC as a function of CO₂ emission reduction, by carbon capture technology. Nomenclature identical to Fig. 16.

applications to other steel plant off-gases (i.e., other than BFG) [30,43,27,74] is currently at low TRL (2).

7. Carbon capture technologies comparison

A comparison of all carbon capture technologies is presented here with respect to 6 different KPIs: thermal penalty (MJ/kgCO₂), electrical penalty (MJ/kgCO₂), economic cost (\$/tCO₂), CO₂ emission reduction (kgCO₂/tHM), CO₂ purity (%) and TRL (-). The results of this comparison are compiled in Fig. 16. In order to provide a holistic insight, the

minimum and maximum values, as well as the average and all the sampling found from all the reviewed scientific works, are included in Fig. 16. Such a representation highlights differences in the presence of each technology in the bibliography.

For oxy blast furnaces with top gas recycling (oxy-TGR) technology, only the decrease of CO₂ because of injecting TGR is accounted for, and not the decrease that would be achieved by storing underground the pure CO₂ from the CC stage. However, pre-combustion with adsorption (pre-adsor) technology is usually coupled with oxy-BF and TGR, and the combined benefit is taken into account for this technology.

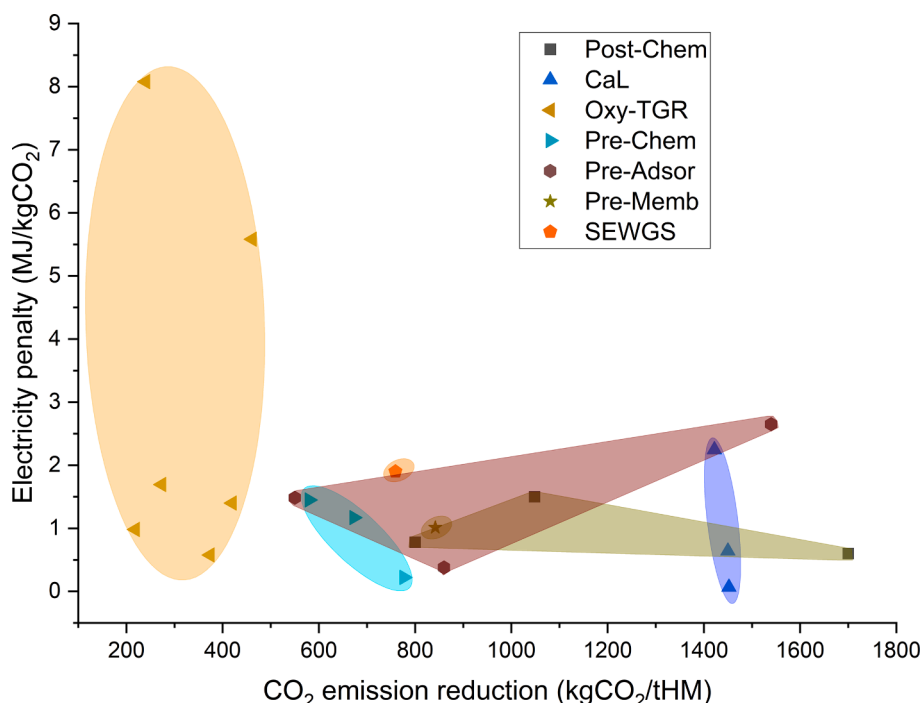


Fig. 18. Electricity penalty as a function of CO₂ emission reduction, by carbon capture technology. Nomenclature identical to Fig. 16.

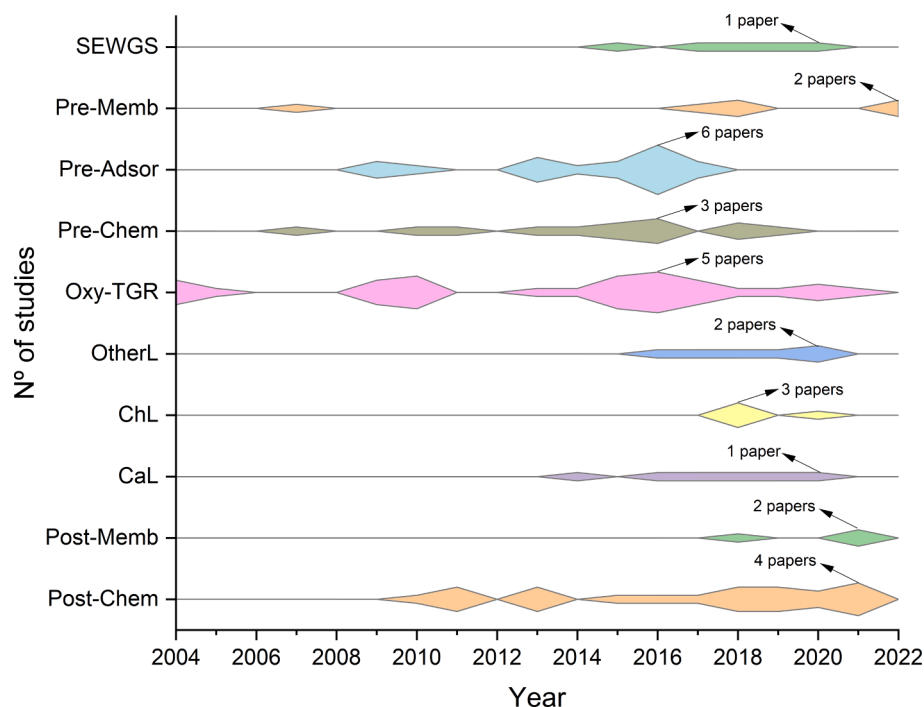


Fig. 19. Number of scientific papers published, organised by year and technology. Nomenclature identical to Fig. 16.

The thermal penalty ranges between 1.3 and 6.2 MJ/kgCO₂ for post-combustion chemical absorption (post-chem), calcium looping (CaL), other looping processes (OtherL) and pre-combustion chemical absorption (pre-chem); in contrast oxy-TGR stands out with 0.3–0.8 MJ/kgCO₂ of thermal energy consumption. Even more, Jin et al. [119] calculates a negative penalty (i.e. there is an excess of available energy) of –13.4 MJ/kgCO₂. Other technologies such as post- and pre-combustion with membranes (post-memb and pre-memb), chemical looping (ChL), pre-adsor and sorption enhanced water gas shift (SEWGS) do not have thermal consumption or the data was not available (n/a).

Regarding the electricity penalty, the ranges are wider. Post-chem, CaL and pre-chem are the technologies with the lower electricity consumption, with less than 1 MJ/kgCO₂ on an average. This is due to the process energy consumption being mainly thermal, with a marginal electricity consumption related to the CO₂ compression (up to 100–150 bar). Post-memb, pre-memb, SEWGS and pre-adsor are technologies that compress the gases to capture CO₂, and therefore have a higher electricity penalty between 1 and 3 MJ/kgCO₂ on an average. Post-memb has an outlier data much higher than the rest (4.4 MJ/kgCO₂, when the average is 1.8 MJ/kgCO₂). This study corresponds to sinter flue gas capture, which has a very low CO₂ content that provokes higher compression consumptions. Oxy-TGR has the wider range of electricity penalty of all technologies due to the great variety of configurations available to implement this technology, from less than 1 MJ/kgCO₂ to more than 8 MJ/kgCO₂.

The bulk of published economic analyses have focused on post-chem technology, with more than 19 samplings compared to other technologies that combine less than 7 samplings. This technology as well as oxy-TGR and pre-chem have the higher average economic cost, with approximately 70 \$/tCO₂, while other technologies range between 30 and 60 \$/tCO₂. It is important to notice that only three studies have performed economic analyses for CaL technology, only one study for SEWGS technology, and none for ChL and OtherL. This highlights a lack of economic analyses in this area. The potential CO₂ mitigation is very high, especially if it is considered that typical emissions from the BF-BOF route range between 2000 and 2200 kgCO₂/tHM. Nevertheless, most of the published works are simulations with a low TRL, and therefore they do not take into account the physical or economical constraints of actual

facilities. So far, the maximum CO₂ amount captured has been 1270 kgCO₂/tHM, reached under TRL 6 in a pilot plant in Lulea (Sweden), within ULCOS project [131]. They focused on an experimental oxy-blast furnace with top gas recycling and VPSA to capture the CO₂.

Only one technology achieves a CO₂ purity lower than 90 %, which is the pre-adsor, with an average of 88 % and a minimum of 80 %. From the remaining technologies, only membranes have an average value lower than 95 % for economic reasons. Many of the technologies analysed have a TRL 6, such as post-memb, oxy-TGR, pre-chem, pre-adsor and SEWGS; nevertheless, most studies are simulations with TRL 2. Thus, only 11 published works belong to TRL 6, 5 belong to TRL 5, and 101 belong to TRL 2–4.

In order to clarify the interrelations between specific economic costs and CO₂ mitigation potential, both KPIs are represented in Fig. 17. Additionally, the space defining the range for each KPI in the reviewed works is highlighted. Firstly, the shape of the coloured regions allows distinguishing the most investigated technologies from those with scarce data. Specifically, the natural trend should be the lower CO₂ reduction, the lower specific cost. Such behaviour is only exhibited by the pre- and post-chem technologies. This is an expected result as it is the only commercial technology of carbon capture, at least in the power sector. Secondly, the two works related to CaL present very different specific costs. This difference is due to the consideration or not of different issues, i.e. Tian et al. [66] considers retrofitting of an existing lime kiln, cutting costs and directly capturing calcination CO₂, while Cormos et al. [39] considers natural gas as a fuel in the calciner instead of coal. Therefore, it can be concluded that it could be a very promising and competitive technology, but further research is needed to corroborate the results. According to the available data, oxy-TGR is expensive and the CO₂ mitigation potential is limited, which is because the CO₂ decrease accounted only takes into account oxy and TGR, and not the CO₂ from the carbon capture stage. Finally, further research is required for pre-adsor, pre-memb and SEWGS to generate sufficient and reliable data, as only one published data is available for each one. Also, post-memb, ChL and OtherL need further research, because no study was found that carried out a technical and economic analysis in the same paper.

The interrelation between electricity penalty and CO₂ mitigation

Table 8
Main KPIs for reviewed CC technologies.

Method/ Technology	Source	Thermal Penalty	Electrical Penalty	Cost	CO ₂ emission reduction	Total CO ₂ produced ^{e)}	CO ₂ purity	TRL	Year	Country	Ref.
–	–	MJt/ kgCO ₂	MJe/ kgCO ₂	\$/tCO ₂	kgCO ₂ /tHM	kgCO ₂ /tHM	%	–	–	–	–
3.1. Post combustion CC with chemical absorption											
MEA	BF flue gas	n/a	n/a	78.5 ^{a)}	n/a	n/a	n/a	2	2011	Australia	[24]
MEA	Coke oven	n/a	n/a	85.6 ^{a)}	n/a	n/a	n/a	2	2011	Australia	[24]
MEA	Hot Stoves, Power plant, Coke oven, Lime kiln	3.0–6.2	n/a	n/a	n/a	n/a	greater than95	2	2019	Romania	[34]
MEA. AEEA and PZ	Hot stoves	3.7	n/a	n/a	n/a	n/a	n/a	4	2010	Taiwan	[25]
MEA	Power plant	3.8–4.1	n/a	n/a	n/a	n/a	n/a	2	2015	Switzerland	[29]
MEA	BF flue gas	n/a	1.5 ^{a)}	68 ^{a)}	2.7 Mt/year	n/a	n/a	2	2011	Australia	[26]
MEA	BF flue gas	4.5–6.5	n/a	n/a	n/a	n/a	n/a	2	2017	Taiwan	[27]
MEA	BF flue gas	4.2	n/a	n/a	1 t/day	n/a	n/a	5	2011	Japan	[28]
RITE-A	BF flue gas	3.3	n/a	n/a	1 t/day	n/a	n/a	5	2011	Japan	[28]
RITE-B	BF flue gas	3.1	n/a	n/a	1 t/day	n/a	n/a	5	2011	Japan	[28]
MEA	Power plant	3.1	0.4 ^{a)}	n/a	n/a	n/a	n/a	2	2018	Sweden	[30]
MDEA	Hot Stoves, Power plant, Coke oven, Lime kiln	2.9	n/a	76–80 ^{a)}	1259	2093	greater than95	2	2016	Romania	[35]
MEA	Power plant, Hot stoves, BFG	3.2	n/a	48.8 ^{a)}	n/a	1700	n/a	2	2019	Sweden	[36]
MEA	Power plant	3.0	n/a	n/a	688	1625	n/a	2	2018	Spain	[31]
MEA	Power plant	n/a	n/a	n/a	230	2093	n/a	2	2019	Romania	[45]
MEA	Hot Stoves, Power plant	3.4	0.4 ^{a)}	92–204 ^{a)}	n/a	n/a	n/a	2	2013	Finland	[37 38]
MEA	Coke Oven	4.1	n/a	n/a	n/a	n/a	n/a	2	2018	UK	[33]
MDEA	Hot Stoves, Power plant, Coke oven, Lime kiln	3.1	n/a	80.9 ^{a)}	1259	2093	n/a	2	2020	Romania	[39]
MEA	Power plant	4.8	n/a	38.2 ^{a)}	369	2016	100	2	2020	Italy	[32]
MEA	Hot Stoves, Power plant	n/a	n/a	46–90 ^{a)}	n/a	n/a	n/a	2	2013	Finland	[40]
MDEA	Coke oven, Lime kiln, BOF	n/a	n/a	n/a	n/a	n/a	98.8	2	2021	Romania	[41]
MEA	Sintering	3.5	0.28 ^{a)}	84.9 ^{a)}	n/a	n/a	n/a	2	2021	S. Korea	[42]
MEA	Coke oven	3.3	0.28 ^{a)}	69.9 ^{a)}	n/a	n/a	n/a	2	2021	S. Korea	[42]
MEA	Lime kiln	3.3	0.28 ^{a)}	72.2 ^{a)}	n/a	n/a	n/a	2	2021	S. Korea	[42]
MEA	Power plant	2.9	0.28 ^{a)}	63 ^{a)}	n/a	n/a	n/a	2	2021	S. Korea	[42]
MEA	Hot stoves	2.9	0.28 ^{a)}	65.6 ^{a)}	n/a	n/a	n/a	2	2021	S. Korea	[42]
MEA	–	4.4	0.78 ^{a)}	53.8 ^{a)}	800	1800	n/a	2	2021	The Nether.	[43]
MDEA/PZ	Hot Stoves, Power plant, Coke oven, Lime kiln	2.3	0.6 ^{a)}	97 ^{a)}	1700	2200	n/a	2	2021	The Nether.	[44]
3.2. Post combustion CC with membranes											
Polaris	Power plant, Hot stoves	n/a	0.9–1.0 ^{a)}	36–47 ^{a)}	n/a	n/a	99	6	2018	USA	[59]
PVAm	Power plant, Hot stoves	n/a	1.2 ^{a)}	n/a	n/a	n/a	95.5	3	2021	Romania	[41]
Polyimide	Sintering	n/a	4.4 ^{a)}	252.7 ^{a)}	n/a	n/a	90	3	2021	S. Korea	[42]
Polyimide	Coke oven	n/a	1.7 ^{a)}	91.2 ^{a)}	n/a	n/a	90	3	2021	S. Korea	[42]
Polyimide	Lime kiln	n/a	1.4 ^{a)}	83.3 ^{a)}	n/a	n/a	90	3	2021	S. Korea	[42]
Polyimide	Power plant	n/a	1.2 ^{a)}	54 ^{a)}	n/a	n/a	90	3	2021	S. Korea	[42]
Polyimide	Hot stoves	n/a	1.2 ^{a)}	52.5 ^{a)}	n/a	n/a	90	3	2021	S. Korea	[42]
4.1. Looping CC with calcium looping											
CaO-CaCO ₃	BFG, BOFG, hot stoves, lime kilns, COG	2.9	0.5	60.2 €/tHRC	1023–1709	2090	95	2	2016	Romania	[35]
CaO-CaCO ₃	Hot stoves, BFG, BOFG, COG, lime kiln	2.7	0.1 ^{a)}	N.A.	1337–1567	2090	95	2	2019	Romania	[34]
CaO-CaCO ₃	Hot stoves, BFG, BOFG, COG, lime kiln	n.a.	0.6	73.8 ^{a)}	1450	2090	95	2	2020	Romania	[39]

(continued on next page)

Table 8 (continued)

Method/ Technology	Source	Thermal Penalty	Electrical Penalty	Cost	CO ₂ emission reduction	Total CO ₂ produced ^{c)}	CO ₂ purity	TRL	Year	Country	Ref.
–	–	MJt/ kgCO ₂	MJe/ kgCO ₂	\$/tCO ₂	kgCO ₂ /tHM	kgCO ₂ /tHM	%	–	–	–	–
3.1. Post combustion CC with chemical absorption											
CaO-CaCO ₃	COG	4.2–5.6	n.a.	n.a.	68071 t/year	68400 t/ year	high- purity	3	2017	China	[62]
CaO-CaCO ₃	BFG	n.a.	n.a.	n.a.	n.a.	n.a.	high- purity	3	2014	Israel	[63]
CaO-CaCO ₃	BFG, COG	2.7	2.5 ^{a)}	13–17 ^{a)}	1422	1800	high- purity	3	2018	China – UK	[66]
4.2. Looping CC with chemical looping											
Fe ₂ O ₃ -FeO	COG	n/a	n/a	n/a	n/a	n/a	n/a	2	2018	China	[81]
Fe ₂ O ₃ -FeO	Sinter plant	n/a	n/a	n/a	n/a	n/a	n/a	3	2020	UK	[80]
Fe ₂ O ₃ -FeO	COG	n/a	n/a	n/a	1771–2280 kmol/h	n/a	90	4	2018	China	[79]
NiO-Ni	COG	n/a	n/a	n/a	n/a	n/a	n/a	4	2018	China	[76]
4.3. Looping CC with other looping processes											
Ca-Cu	BFG	1.4–1.7	n/a	n/a	n/a	n/a	57–99.9	4	2020	Spain	[91]
Ca-Cu		n/a	n/a	n/a	n/a	n/a	n/a		2019	Spain	[90]
Ca-Cu. (COG)	BFG, COG	2.6	n/a	n/a	502	1619	97.4	4	2018	Spain	[31]
Ca-Cu. (COG + NG)	BFG, COG	4.3	n/a	n/a	1618	1619	97.4	4	2018	Spain	[31]
CaCO ₃ /FeO- CaO/Fe ₂ O ₃	BFG	n/a	n/a	n/a	n/a	n/a	n/a	4	2016	China	[92]
MgO/MoO ₃ - MgCO ₃ / MoS ₂	BFG	n/a	n/a	n/a	n/a	n/a	n/a	3	2020	China	[94]
Ca-Cu + SEWGS	BFG, BOFG, COG	n/a	n/a	n/a	617	1922	96.8	5	2017	Spain	[93]
5. CC with Oxy-blast furnace with TGR											
Oxy-BF + TGR	BFG	0.3	5.6	50–90	462 ^{b)}	1244	n/a	2	2014	Finland	[106 146]
Oxy-BF + TGR	BFG	n/a	n/a	n/a	550 ^{b)}	1320	n/a	2	2004	Japan	[108]
Oxy-BF + TGR	BFG	0.8	1.7	n/a	273 ^{b)}	1367	n/a	4	2015–2016	Japan	[100109 111]
Oxy-BF + TGR + PtG	BFG	n/a	n/a	n/a	176 ^{b)}	1330	n/a	2	2021	Spain	[121]
Oxy-BF + TGR	BFG	n/a	0.6	n/a	374 ^{b)}	1338	n/a	2	2004	Korea	[182]
Oxy-BF + TGR	BFG	n/a	1.4	90	420 ^{b)}	2420	n/a	2	2013	Australia	[102]
Oxy-BF + TGR	BFG	n/a	0.98	n/a	218 ^{b)}	1377	n/a	2	2016	China	[112]
Oxy-BF + TGR	BFG	n/a	n/a	n/a	489 ^{b)}	1429	n/a	2	2016	China	[103]
Oxy-BF + TGR	BFG	–13.4	8.1	n/a	240 ^{b)}	2066	n/a	2	2015	China	[119]
Oxy-BF + TGR	BFG	n/a	n/a	n/a	n/a	n/a	n/a	2	2019	Japan	[161]
Oxy-BF + TGR	BFG	n/a	n/a	n/a	181 ^{b)}	1267	n/a	2	2010	China	[127]
Oxy-BF + TGR	BFG	n/a	n/a	n/a	n/a	n/a	n/a	4	2018	China	[139]
Oxy-fuel coke oven	COG	n/a	n/a	n/a	n/a	n/a	n/a	3	2020	China	[183]
Oxy-BF + TGR	BFG	n/a	n/a	n/a	184	1397	n/a	2	2020	China	[184]
Oxy-BF + TGR	BFG	n/a	n/a	n/a	167 ^{b)}	1450	n/a	2	2010	Finland	[104]
Oxy-BF + TGR	BFG	n/a	n/a	58.5	390	1600	n/a	2	2010	Finland	[105]
Oxy-BF + TGR	BFG	n/a	n/a	n/a	100	n/a	n/a	2	2015	Finland	[114]
Oxy-BF + TGR	BFG	n/a	n/a	n/a	367 ^{b)}	1547	n/a	2	2015	India	[110]
Oxy-BF + TGR	BFG	n/a	n/a	n/a	401 ^{b)}	1670	n/a	6	2009	Sweden	[169131 115]
Oxy-BF + TGR	BFG	n/a	n/a	n/a	506 ^{b)}	1701	n/a	2	2017	China	[129]

(continued on next page)

Table 8 (continued)

Method/ Technology	Source	Thermal Penalty	Electrical Penalty	Cost	CO ₂ emission reduction	Total CO ₂ produced ^{c)}	CO ₂ purity	TRL	Year	Country	Ref.
–	–	MJt/ kgCO ₂	MJe/ kgCO ₂	\$/tCO ₂	kgCO ₂ /tHM	kgCO ₂ /tHM	%	–	–	–	–
3.1. Post combustion CC with chemical absorption											
Oxy-BF + TGR	BFG	n/a	n/a	n/a	166 ^{b)}	1305	n/a	2	2016	China	[128]
Oxy-BF + TGR	BFG	n/a	n/a	n/a	159 ^{b)}	1330	n/a	2	2017	China	[130]
Oxy-BF + TGR	BFG	n/a	n/a	n/a	279 ^{b)}	1305	n/a	2	2017	China	[185]
Oxy-BF + TGR	BFG	n/a	n/a	n/a	−7.5 ^{b)}	1436	n/a	2	2004	Japan	[137]
Oxy-BF + TGR	BFG	n/a	n/a	n/a	−27 ^{b)}	1408	n/a	2	2005	Japan	[116]
Oxy-BF + TGR	BFG	n/a	n/a	n/a	247 ^{b)}	1379	n/a	2	2010	Japan	[186]
6.1. Pre combustion CC with chemical absorption											
TGR-BF + PZ	BFG	2.4	0.2 ^{a)}	40.6 ^{a)}	777	1268	95	2	2018	Korea	[160]
unspecified	BFG	2	n/a	n/a	n/a	n/a	90	6	2016	Japan	[165]
MEA	BFG, BOFG, COG	3.1	0.4 ^{a)}	62 ^{a)}	130 t/h	141 t/h	99	2	2015	Korea	[158]
NH ₃	BFG	2.5	n/a	n/a	9.3 t/day	10.3 t/day	99	6	2014	Korea	[155]
TGR-BF + MEA	BFG	n/a	1.4–1.5 ^{a)}	86–97 ^{a)}	520–640	n/a	n/a	2	2013	Australia	[102]
ULCOS-MDEA	BFG	3.2	0.61 ^{a)}	n/a	n/a	n/a	n/a	–	2010	France	[163]
MEA	BFG	n/a	n/a	n/a	564	n/a	n/a	2	2019	Sweden	[153]
MEA	BFG	3	0.4 ^{a)}	n/a	n/a	n/a	n/a	2	2018	Sweden	[30]
unspecified	BFG	2	n/a	n/a	n/a	n/a	n/a	6	2016	Japan	[107]
MEA	BFG, BOFG	1.3	1.2 ^{a)}	n/a	673	961	n/a	2	2016	Finland	[159]
MDEA	BFG, BOFG, COG	2.9	n/a	n/a	1192 g/kW _{el}	1339 g/ kW _{he}	98.6	2	2015	Italy	[29]
MEA	BFG	4.4	n/a	68 ^{a)}	1048	1188	n/a	2	2011	Australia	[26]
MEA, MDEA/ PZ and AMP	BFG	2.2–2.4	n/a	n/a	n/a	n/a	90–97	2	2007	Norway	[151]
6.2. Pre combustion CC with adsorption											
TGR + VPSA	OBFG	n/a	0.4 ^{a)}	n/a	860	1650	n/a	2	2016	Malaysia	[101]
PSA	BFG	n/a	n/a	n/a	3 t/day	3.8 t/day	99.5	6	2016	Japan	[165]
PSA	BFG, BOFG, COG	n/a	1.0 ^{a)}	38 ^{a)}	130 t/h	141 t/h	91.2	2	2015	S. Korea	[158]
TGR-VPSA	BFG	n/a	0.9–1.0 ^{a)}	59–60 ^{a)}	520–580	n/a	n/a	2	2013	Australia	[102]
VPSA	BFG	n/a	1.1 ^{a)}	n/a	n/a	n/a	n/a	2	2010	France	[163]
TGR + VPSA	OBFG	n/a	0.9	n/a	2.7 Mt/a	3.2 Mt/a	92	2	2014	Finland	[106]
TGR + VPSA	BFG	n/a	n/a	n/a	1794	1900	87.7	2	2013	India	[142]
PSA	BFG	n/a	0.4 ^{a)}	n/a	6.3 t/day	7.8 t/day	n/a	5	2013	Japan	[187]
TGR + VPSA	BFG	n/a	n/a	n/a	1270	1671	n/a	6	2009	Sweden	[131]
TGR + VPSA	OBFG	n/a	n/a	n/a	n/a	n/a	n/a	2	2017	China	[130]
TGR + VPSA	OBFG	n/a	2.7 ^{a)}	n/a	1540	2080	n/a	2	2017	China	[119]
TGR + VPSA	OBFG	n/a	10.4 kJ/ thm	n/a	701	1485	n/a	2	2016	China	[103]
TGR + VPSA	OBFG	n/a	0.4 ^{a)}	n/a	860	1650		6	2016	Malaysia	[11]
TGR + VPSA	BFG	n/a	n/a	n/a	1200	n/a	79.9	2	2016	Japan	[111]
TGR + VPSA	OBFG	n/a	n/a	n/a	906–1117	1313	n/a	2	2016	China	[128]
TGR + VPSA	OBFG	n/a	n/a	n/a	607	n/a	79.9	2	2015	India	[110]
TGR + VPSA	BFG	n/a	n/a	n/a	n/a	n/a	n/a	6	2009	Sweden	[169]
6.3. Pre combustion CC with membranes											
2-stage memb.	BFG	n/a	n/a	32.5–50.6	n/a	n/a	90–98	2	2018	France	[162]
TGR + 2-stage memb.	BFG	n/a	1.0 ^{a)}	34.4 ^{a)}	842	1268	95	2	2018	Korea	[160]
2-stage memb.	BFG	n/a	n/a	32.5–50.6	n/a	n/a	n/a	2	2017	France	[173]
2-stage memb.	BFG	n/a	0.9 ^{a)}	17.5 ^{a)}	407 t/h	420 t/h	90	3	2007	Norway	[174]
1-stage memb.	BOFG	n/a	n/a	n/a	n/a	n/a	55	4	2022	Korea	[175]
membrane + cryogenic	BFG	n/a	1.6 ^{a)}	28.8 ^{a)}	n/a	n/a	97.1	2	2022	China	[178]
6.4. Pre combustion CC with SEWGS											

(continued on next page)

Table 8 (continued)

Method/ Technology	Source	Thermal Penalty	Electrical Penalty	Cost	CO ₂ emission reduction	Total CO ₂ produced ^{c)}	CO ₂ purity	TRL	Year	Country	Ref.
–	–	MJ/ kgCO ₂	MJe/ kgCO ₂	\$/tCO ₂	kgCO ₂ /tHM	kgCO ₂ /tHM	%	–	–	–	–
3.1. Post combustion CC with chemical absorption											
SEWGS	BFG	n/a	n/a	n/a	14 t/day	n/a	n/a	6	2018	The Nether.	[179]
WGS + SEWGS	BFG, BOFG, COG	n/a	1.9 ^{a)}	36.4 ^{a)}	759	2016	97.6–99.6	2	2020	Italy	[32]
SEWGS w/ saturator	BFG, BOFG, COG	n/a	n/a	n/a	833	1422	n/a	2	2019	Romania	[45]
SEWGS w/ saturator	BFG, BOFG, COG	n/a	2.9 ^{a)}	n/a	149 g/kWhe	1339 g/ kWhe	98.85	2	2015	Italy	[29]
SEWGS + Ca- Cu	BFG, BOFG, COG	n/a	n/a	n/a	617	1922	96.8	2	2017	Spain	[93]

^{a)} Electricity penalty and cost includes CO₂ compression to 100–150 bar.

^{b)} It is only accounted the decrease of CO₂ because of injecting TGR and making oxy-blast furnace, and not the decrease that would be achieved by storing underground the pure CO₂ from the CC stage.

^{c)} Typical CO₂ emissions in an integrated steel plant are 1900–2200 kgCO₂/tHM. When lower than this value, the total CO₂ produced in the industry refers to the steel plant without the power plant, or refers only to the blast furnace.

potential is represented in Fig. 18, where the space representing the range of each KPI values in the reviewed works is highlighted for each category of technologies. The highest electricity penalty with lower CO₂ emission reduction is obtained with oxy-TGR, but as noted above, the CO₂ decrease depends on how it is measured, and the high electricity penalty is due to the high demand of pure oxygen (high ASU consumption). For post-chem, CaL, pre-chem and pre-adsor, three studies per technology were identified that analysed both KPIs. Post-chem and pre-adsor have a wide range of CO₂ reduction while maintaining the electricity penalty at reasonable levels. CaL and pre-chem have less sparse range of values obtaining similar electricity penalty, with the former achieving higher CO₂ emission reduction. Finally, the technologies with scarce data are pre-memb and SEWGS, while for post-memb, ChL and OtherL no scientific studies analysing both KPIs were identified, which highlights the necessity for further research in this field.

With the aim of highlighting the most studied or novel technologies, Fig. 19 shows the number of scientific papers organised by year and technology. Oxy-TGR is the most studied technology, with 30 published papers, with the first one from 2004. Other well studied and known technologies are post-chem, pre-chem and pre-adsor, with 22, 13 and 17 published papers, respectively, starting between 2007 and 2010. Among the most novel technologies, pre- and post-memb, SEWGS, CaL, ChL and OtherL can be found. Investigation of these novel technologies in the I&S industry began between 2014 and 2016, and therefore, none of them have more than 6 published papers.

For a rapid development of all CC technologies, the high costs and the availability of underground storage are the main bottlenecks. For new promising technologies, such as membranes, CaL or otherL, the low TRL and the lack of pilot plants can also be considered as a bottleneck, which slow down the development of these technologies. Besides, another technical limitation appears for pre-combustion CC technologies. These technologies capture CO₂ from energetic streams such as BFG, COG and BOFG, which also have significant amounts of CO and CH₄. The two latter components will be combusted, converted to CO₂ and emitted to the atmosphere. This means that this pre-combustion technologies will never be able to capture all the CO₂, and a significant amount of it will be released to the atmosphere. However, this problem can be partially solved if the CO is injected to the BF (i.e., TGR or hot reducing gases) as a reducing agent, as it will be oxidised to CO₂ and captured from the BFG in the pre-combustion CC stage. This partial solution is limited by the amount of CO that can be injected in the BF. Other possible solution to this problem is implementing a water gas shift (WGS) prior the pre-combustion CC stage, shifting all the CO to CO₂ and generating additional H₂. However, this solution arises other economic and technical limitations regarding the use of pure H₂, such as different

combustion behaviour, high volatility, and higher compression energy.

8. Conclusions

In this article, we have presented the first systematic review of the state of the art related to carbon capture (CC) technologies investigated to date for the blast furnace-basic oxygen furnace route of the I&S industry, in terms of energy penalisation, economic cost, efficiency, TRL and volume of CO₂ captured. These technologies can capture CO₂ from different streams, producing a concentrated stream of CO₂ that can subsequently be stored or used.

Based on a systematic review procedure, 120 research articles were identified and reviewed. The reported carbon capture technologies were classified into four different groups: post-combustion, looping cycles, oxy-combustion and pre-combustion. Each group of technologies was further divided according to the specific CC technology applied: chemical absorption, physical adsorption, membranes, calcium looping, chemical looping and sorption enhanced water gas shift (SEWGS). For oxy-combustion CC technology, only oxy-blast furnace technology was found to be reported (i.e. oxy-combustion of BOFG or other gases were not found in literature, in relation to CC technology). When the capture is allocated in streams with combustible species such as BFG, BOFG or COG, it is considered a pre-combustion system, while when applied to flue gases, it is a post-combustion system. No study related to pre-combustion CC systems was identified for the solid fuels employed in the I&S industry (i.e., coal and coke).

An in-depth review of each carbon capture (CC) technology was carried out individually, as well as a detailed comparison of the technologies considering 6 KPIs: thermal penalty (MJ/kgCO₂), electrical penalty (MJ/kgCO₂), economic cost (\$/tCO₂), CO₂ emission reduction (kg/tHM), CO₂ purity (%) and TRL.

Calcium looping (CaL), precombustion CC by adsorption (pre-adsor) and postcombustion CC by chemical absorption (post-chem) are the three technologies that have lower electricity penalty when capturing high amounts of CO₂. CaL technology achieves the highest CO₂ emission reduction at a lower cost, but other technologies such as post-chem, other looping processes (otherL) and pre-adsor achieve similar carbon capture rates. Precombustion CC by membranes (pre-memb) has the lowest economic costs, but with lower potential of carbon capture to date.

None of the technologies investigated for carbon capture are yet commercially available in the I&S industry, but some have achieved TRL 6, including postcombustion CC by membranes (post-memb), oxy-blast furnace with top gas recycling (oxy-TGR), precombustion CC by chemical absorption (pre-chem), pre-adsor and sorption enhanced water gas

shift (SEWGS). The latter stands out as a promising alternative because, despite its high penalty, it has a low cost and very high CO₂ purity.

In the I&S industry, as the main CO₂ streams have a similar CO₂ content (i.e., 15–28 %vol) and a similar pressure (atmospheric to 2 bar), it is not of significant importance where the CC stage is implemented, the capture will behave similarly. Taking into account the 6 KPIs analysed, pre-adsor is found to be the most promising technology for capturing CO₂ in the I&S industry, according to its high TRL and CO₂ emission reduction at reasonable cost and penalty. The only drawback of this technology is the level of CO₂ purity (88 % on an average), that is lower than for other technologies.

There are substantial opportunities for further investigation to characterise the actual potential of those technologies for decarbonisation of the I&S industry. Future research must reaffirm technologies with demonstrated positive outcomes such as pre-adsor, and also focus on new promising technologies such as membranes, CaL or other L. Furthermore, the currently high costs of this kind of systems must be tackled, and there is a necessity to optimize the thermal and electrical penalties associated with carbon capture.

Declaration of Competing Interest

The authors declare that they have no known competing financial interests or personal relationships that could have appeared to influence the work reported in this paper.

Data availability

Data will be made available on request.

Acknowledgements

The work presented in this paper has been supported by both the University of Zaragoza under the project UZ2020-TEC-06 and Khalifa University project CIRA-2020-080. This work has also received funding from the European Union's Horizon 2020 research and innovation program under the Marie Skłodowska-Curie grant agreement no. 887077. This work is part of the R&D project PID2021-126164OB-I00, funded by MCIN/AEI/10.13039/501100011033/ and by the "ERDF A way of making Europe".

Annex A. Main KPIs for reviewed CC technologies

The KPI data (i.e., thermal and electrical penalties (MJ/kgCO₂), economic cost (\$/tCO₂), CO₂ emission reduction (kgCO₂/tHM), CO₂ purity (%) and TRL) extracted from the 120 studies reviewed and analysed in Section 7, is compiled in Table A for each technology, namely post combustion CC with either chemical absorption or membranes, Looping CC with calcium, chemical or other looping process, CC with oxy-blast furnace and TGR, and pre combustion CC with either chemical absorption, adsorption, membranes or SEWGS.

References

- [1] "European Commission." https://energy.ec.europa.eu/index_es (accessed Apr. 26, 2022).
- [2] "International Energy Agency, Iron and Steel Technology Roadmap, 2020." <https://www.iea.org/reports/iron-and-steel-technology-roadmap> (accessed Apr. 26, 2022).
- [3] Huitu K, Helle H, Helle M, Kekkonen M, Saxén H. Optimization of steelmaking using fastmet direct reduced iron in the blast furnace. *ISIJ Int* 2013;53(12): 2038–46. <https://doi.org/10.2355/isijinternational.53.2038>.
- [4] Larrea and García-Berezo. *Economía Industrial* 406 (2017) 139–150.
- [5] "Midrex Technologies, 2022. The Midrex (R) Process." <https://www.midrex.com/technology/midrex-process/> (accessed May 10, 2022).
- [6] "Energiiron, DRI technology by Tenova and Danieli." <https://www.energiiron.com/> (accessed Jun. 08, 2022).
- [7] Ariyama T, Takahashi K, Kawahiri Y, Nouchi T. Diversification of the Ironmaking Process Toward the Long-Term Global Goal for Carbon Dioxide Mitigation. *J Sustain Metall* 2019;5(3):276–94. <https://doi.org/10.1007/s40831-019-00219-9>.
- [8] He H, Guan H, Zhu X, Lee H. Assessment on the energy flow and carbon emissions of integrated steelmaking plants. *Energy Rep* 2017;3:29–36. <https://doi.org/10.1016/j.egypr.2017.01.001>.
- [9] J. von Scheele, "Decarbonization of Ironmaking," 2021.
- [10] Deerberg G, Oles M, Schlögl R. The Project Carbon2Chem®. *Chem-Ing-Technol* 2018;90(10):1365–8. <https://doi.org/10.1002/cite.201800060>.
- [11] Abdul Quader M, Ahmed S, Dawal SZ, Nukman Y. Present needs, recent progress and future trends of energy-efficient Ultra-Low Carbon Dioxide (CO₂) Steelmaking (ULCOS) program. *Renew Sustain Energy Rev Mar.* 2016;55:537–49. <https://doi.org/10.1016/j.rser.2015.10.101>.
- [12] X. L. Zhou and Z. N. Du, "The introduction of COREX process development," *Adv. Mater. Res.*, vol. 774–776, no. November 1995, pp. 1430–1433, 2013, doi: 10.4028/www.scientific.net/AMR.774-776.1430.
- [13] N. Goodman, "Operation of the First Hismelt Plant in China," 2017. [Online]. Available: https://www.researchgate.net/publication/298065259_Hismelt_Plant_Ramp-Up.
- [14] Yi SH, Choi ME, Kim DH, Ko CK, Il Park W, Kim SY. FINEX® as an environmentally sustainable ironmaking process. *Ironmak Steelmak* 2019;46(7): 625–31. <https://doi.org/10.1080/03019233.2019.1641682>.
- [15] Tonomura S. Outline of course 50. *Energy Procedia* 2013;37:7160–7. <https://doi.org/10.1016/j.egypro.2013.06.653>.
- [16] J. Perpiñán, M. Bailera, B. Peña, L. M. Romeo, and V. Evely, "Technical and economic assessment of iron and steelmaking decarbonisation via power to gas and amine scrubbing," 2022, [Online]. Available: Under Revision.
- [17] K. He and L. Wang, "A review of energy use and energy-efficient technologies for the iron and steel industry," *Renew. Sustain. Energy Rev.*, vol. 70, no. September 2016, pp. 1022–1039, 2017, doi: 10.1016/j.rser.2016.12.007.
- [18] Leeson D, Mac Dowell N, Shah N, Petit C, Fennell PS. A Techno-economic analysis and systematic review of carbon capture and storage (CCS) applied to the iron and steel, cement, oil refining and pulp and paper industries, as well as other high purity sources. *Int J Greenh Gas Control* 2017;61:71–84. <https://doi.org/10.1016/j.ijggc.2017.03.020>.
- [19] Zhao J, Zuo H, Wang Y, Wang J, Xue Q. Review of green and low-carbon ironmaking technology. *Ironmak Steelmak* 2020;47(3):296–306. <https://doi.org/10.1080/03019233.2019.1639029>.
- [20] Fan Z, Friedmann SJ. Review Low-carbon production of iron and steel: Technology options, economic assessment, and policy. *Joule* 2021;1–34. <https://doi.org/10.1016/j.joule.2021.02.018>.
- [21] Quader MA, Ahmed S, Ghazilla RAR, Ahmed S, Dahari M. A comprehensive review on energy efficient CO₂ breakthrough technologies for sustainable green iron and steel manufacturing. *Renew Sustain Energy Rev* 2015;50:594–614. <https://doi.org/10.1016/j.rser.2015.05.026>.
- [22] "European Commission. Research and innovation. Horizon 2020. Work programme 2014–2015, general annexes." https://ec.europa.eu/research/participants/data/ref/h2020/wp/2014_2015/annexes/h2020-wp1415-annex-gr-trl_en.pdf (accessed Nov. 15, 2022).
- [23] W. Y. Hong, "A techno-economic review on carbon capture, utilisation and storage systems for achieving a net-zero CO₂ emissions future," *Carbon Capture Sci. Technol.*, vol. 3, no. December 2021, p. 100044, 2022, doi: 10.1016/j.ccs.2022.100044.
- [24] Wiley DE, Ho MT, Bustamante A. Assessment of opportunities for CO₂ capture at iron and steel mills: An Australian perspective. *Energy Procedia* 2011;4:2654–61. <https://doi.org/10.1016/j.egypro.2011.02.165>.
- [25] Cheng H, Shen J, Tan C. CO₂ capture from hot stove gas in steel making process. *Int J Greenh Gas Control* 2010;4(3):525–31. <https://doi.org/10.1016/j.ijggc.2009.12.006>.
- [26] Ho MT, Allinson GW, Wiley DE. Comparison of MEA capture cost for low CO₂ emissions sources in Australia. *Int J Greenh Gas Control* 2011;5(1):49–60. <https://doi.org/10.1016/j.ijggc.2010.06.004>.
- [27] Chamchan N, et al. Comparison of rotating packed bed and packed bed absorber in pilot plant and model simulation for CO₂ capture. *J Taiwan Inst Chem Eng* 2017;73:20–6. <https://doi.org/10.1016/j.jtice.2016.08.046>.
- [28] Goto K, Okabe H, Chowdhury FA, Shimizu S, Fujioka Y, Onoda M. Development of novel absorbents for CO₂ capture from blast furnace gas. *Int J Greenh Gas Control* 2011;5(5):1214–9. <https://doi.org/10.1016/j.ijggc.2011.06.006>.
- [29] Gazzani M, Romano MC, Manzolini G. CO₂ capture in integrated steelworks by commercial-ready technologies and SEWGS process. *Int J Greenh Gas Control* 2015;41:249–67. <https://doi.org/10.1016/j.ijggc.2015.07.012>.
- [30] M. Sundqvist, M. Biermann, F. Normann, M. Larsson, and L. Nilsson, "Evaluation of low and high level integration options for carbon capture at an integrated iron and steel mill," *Int. J. Greenh. Gas Control*, vol. 77, no. July 2017, pp. 27–36, 2018, doi: 10.1016/j.ijggc.2018.07.008.
- [31] Martínez I, Fernández JR, Abanades JC, Romano MC. Integration of a fluidised bed Ca–Cu chemical looping process in a steel mill. *Energy* 2018;163:570–84. <https://doi.org/10.1016/j.energy.2018.08.123>.
- [32] G. Manzolini, A. Giuffrida, P. D. Cobden, H. A. J. van Dijk, F. Ruggeri, and F. Consonni, "Techno-economic assessment of SEWGS technology when applied to integrated steel-plant for CO₂ emission mitigation," *Int. J. Greenh. Gas Control*, vol. 94, no. February 2019, p. 102935, 2020, doi: 10.1016/j.ijggc.2019.102935.
- [33] Oko E, Zaccchello B, Wang M, Fethi A. Process analysis and economic evaluation of mixed aqueous ionic liquid and monoethanolamine (MEA) solvent for CO₂ capture from a coke oven plant. *Greenh Gases Sci Technol* 2018;8(4):686–700. <https://doi.org/10.1002/ghg.1772>.

- [34] Chisalita DA, Petrescu L, Cobden P, H. A. J. (Eric. van Dijk, A. M. Cormos, and C. C. Cormos). Assessing the environmental impact of an integrated steel mill with post-combustion CO₂ capture and storage using the LCA methodology. *J Clean Prod* 2019;211:1015–25. <https://doi.org/10.1016/j.jclepro.2018.11.256>.
- [35] Cormos CC. Evaluation of reactive absorption and adsorption systems for post-combustion CO₂ capture applied to iron and steel industry. *Appl Therm Eng* 2016;105:56–64. <https://doi.org/10.1016/j.applthermaleng.2016.05.149>.
- [36] Biermann M, Ali H, Sundqvist M, Larsson M, Normann F, Johnsson F. Excess heat-driven carbon capture at an integrated steel mill – Considerations for capture cost optimization. *Int J Greenh Gas Control* 2019;vol. 91, no. September:102833. <https://doi.org/10.1016/j.ijggc.2019.102833>.
- [37] Arasto A, Tsupari E, Kärki J, Pislä E, Sorsamäki L. Post-combustion capture of CO₂ at an integrated steel mill - Part I: Technical concept analysis. *Int J Greenh Gas Control* 2013;16:271–7. <https://doi.org/10.1016/j.ijggc.2012.08.018>.
- [38] Tsupari E, Kärki J, Arasto A, Pislä E. Post-combustion capture of CO₂ at an integrated steel mill - Part II: Economic feasibility. *Int J Greenh Gas Control* 2013; 16:278–86. <https://doi.org/10.1016/j.ijggc.2012.08.017>.
- [39] Cormos A-M, Dragan S, Petrescu L, Sandu V, Cormos C-C. Techno-Economic and Environmental Evaluations of Decarbonized Fossil-Intensive Industrial Processes by Reactive Absorption & Adsorption CO₂ Capture Systems. *Energies* 2020;no. 13. <https://doi.org/10.3390/en13051268>.
- [40] Arasto A, Tsupari E, Kärki J, Sihvonen M, Lilja J. Costs and potential of carbon capture and storage at an integrated steel mill. *Energy Procedia* 2013;37: 7117–24. <https://doi.org/10.1016/j.egypro.2013.06.648>.
- [41] Luca AV, Petrescu L. Membrane technology applied to steel production: Investigation based on process modelling and environmental tools. *J Clean Prod* 2021;294:126256. <https://doi.org/10.1016/j.jclepro.2021.126256>.
- [42] Yun S, Jang MG, Kim JK. Techno-economic assessment and comparison of absorption and membrane CO₂ capture processes for iron and steel industry. *Energy* 2021;229:120778. <https://doi.org/10.1016/j.energy.2021.120778>.
- [43] F. Yang, J. C. Meerman, and A. P. C. Faaij, "Carbon capture and biomass in industry: A techno-economic analysis and comparison of negative emission options," *Renew. Sustain. Energy Rev.*, vol. 144, no. June 2020, p. 111028, 2021, doi: 10.1016/j.rser.2021.111028.
- [44] Yang F, Meerman H, Faaij A. Harmonized comparison of virgin steel production using biomass with carbon capture and storage for negative emissions. *Int J Greenh Gas Control* 2021;112(April). <https://doi.org/10.1016/j.ijggc.2021.103519>.
- [45] Petrescu L, Chisalita DA, Cormos CC, Manzolini G, Cobden P, van Dijk HAJ. Life cycle assessment of SEWGS technology applied to integrated steel plants. *Sustain* 2019;11(7):pp. <https://doi.org/10.3390/su11071825>.
- [46] D. Kearns, H. Liu, and C. Consoli, "Technology Readiness and Costs of CCS - Global CCS Institute," *Glob. CCS Institute*, no. March, p. 50, 2021, [Online]. Available: <https://www.globalccsinstitute.com/wp-content/uploads/2021/03/Technology-Readiness-and-Costs-for-CCS-2021-1.pdf>.
- [47] Sendeco2, "sendeco2." <https://www.sendeco2.com/es/precios-co2%0A> (accessed Mar. 02, 2022).
- [48] Han Y, Ho WSW. Recent advances in polymeric membranes for CO₂ capture. *Chinese J Chem Eng* 2018;26(11):2238–54. <https://doi.org/10.1016/j.cjche.2018.07.010>.
- [49] Lei L, Lindbråthen A, Hillestad M, Sandru M, Favvas EP, He X. Screening Cellulose Spinning Parameters for Fabrication of Novel Carbon Hollow Fiber Membranes for Gas Separation. *Ind Eng Chem Res* 2019;58(29):13330–9. <https://doi.org/10.1021/acs.iecr.9b02480>.
- [50] Cao Y, Zhang K, Sanyal O, Koros WJ. Carbon Molecular Sieve Membrane Preparation by Economic Coating and Pyrolysis of Porous Polymer Hollow Fibers. *Angew Chemie - Int Ed* 2019;58(35):12149–53. <https://doi.org/10.1002/anie.201906653>.
- [51] Elsaidi SK, et al. Custom Formulation of Multicomponent Mixed-Matrix Membranes for Efficient Post-combustion Carbon Capture. *Cell Reports Phys Sci* 2020;1(7):100113. <https://doi.org/10.1016/j.xcrp.2020.100113>.
- [52] Cacho-Bailo F, et al. On the molecular mechanisms for the H₂/CO₂ separation performance of zeolite imidazolate framework two-layered membranes. *Chem Sci* 2017;8(1):325–33. <https://doi.org/10.1039/c6sc02411d>.
- [53] Huang A, Dou W, Caro J. Steam-stable zeolitic imidazolate framework ZIF-90 membrane with hydrogen selectivity through covalent functionalization. *J Am Chem Soc* 2010;132(44):15562–4. <https://doi.org/10.1021/ja108774v>.
- [54] Kim HW, et al. Selective Gas Transport Through Few-Layered Graphene and Graphene Oxide Membranes. *Science* 2013;380(-), 342(October):91–5.
- [55] Lei L, Bai L, Lindbråthen A, Pan F, Zhang X, He X. Carbon membranes for CO₂ removal: Status and perspectives from materials to processes. *Chem Eng J* 2020; vol. 401, no. June:126084. <https://doi.org/10.1016/j.cej.2020.126084>.
- [56] Freeman BD. Basis of permeability/selectivity tradeoff relations in polymeric gas separation membranes. *Macromolecules* 1999;32(2):375–80. <https://doi.org/10.1021/ma9814548>.
- [57] N. McQueen, C. M. Woodall, P. Psarras, and J. Wilcox, "Chapter 11: CCS in the iron and steel industry," in *RSC Energy and Environment Series*, vol. 2020-Janua, no. 26, 2020, pp. 353–391. doi: 10.1039/9781788012744-00353.
- [58] Hägg MB, Lindbråthen A, He X, Nodeland SG, Cantero T. Pilot Demonstration-reporting on CO₂ Capture from a Cement Plant Using Hollow Fiber Process. *Energy Procedia* 2017;114(1876):6150–65. <https://doi.org/10.1016/j.egypro.2017.03.1752>.
- [59] Baker RW, Freeman B, Kniep J, Huang YI, Merkel TC. CO₂ Capture from Cement Plants and Steel Mills Using Membranes. *Ind Eng Chem Res* 2018;57(47): 15963–70. <https://doi.org/10.1021/acs.iecr.8b02574>.
- [60] White LS, Amo KD, Wu T, Merkel TC. Extended field trials of Polaris sweep modules for carbon capture. *J Memb Sci* 2017;542(August):217–25. <https://doi.org/10.1016/j.memsci.2017.08.017>.
- [61] Global CCS Institute Toshiba, "Applying carbon capture and storage to a Chinese steel plant." 2015.
- [62] Xie H, Yu Q, Zhang Y, Zhang J, Liu J, Qin Q. New process for hydrogen production from raw coke oven gas via sorption-enhanced steam reforming: Thermodynamic analysis. *Int J Hydrogen Energy* 2017;42(5):2914–23. <https://doi.org/10.1016/j.ijhydene.2016.12.046>.
- [63] Halmann M, Steinfeld A. Reforming of blast furnace gas with methane, steam, and lime for syngas production and CO₂ capture: A thermodynamic study. *Miner Process Extr Metall Rev* 2015;36(1):7–12. <https://doi.org/10.1080/08827508.2013.793682>.
- [64] Liu Y, Li Z, Xu L, Cai N. Effect of sorbent type on the sorption enhanced water gas shift process in a fluidized bed reactor. *Ind Eng Chem Res* 2012;51(37):11989–97. <https://doi.org/10.1021/ie301100y>.
- [65] Tian S, Jiang J, Yan F, Li K, Chen X, Manovic V. Highly efficient CO₂ capture with simultaneous iron and CaO recycling for the iron and steel industry. *Green Chem* 2016;18(14):4022–31. <https://doi.org/10.1039/c6gc00400h>.
- [66] Tian S, Jiang J, Zhang Z, Manovic V. Inherent potential of steelmaking to contribute to decarbonization targets via industrial carbon capture and storage. *Nat Commun* 2018;9(1):1–8. <https://doi.org/10.1038/s41467-018-06886-8>.
- [67] Arias B, et al. Demonstration of steady state CO₂ capture in a 1.7 MWth calcium looping pilot. *Int J Greenh Gas Control* Oct. 2013;18:237–45. <https://doi.org/10.1016/j.IJGGC.2013.07.014>.
- [68] Kremer J, Galloy A, Ströhle J, Epple B. Continuous CO₂ Capture in a 1-MWth Carbonate Looping Pilot Plant. *Chem Eng Technol* 2013. <https://doi.org/10.1002/ceat.201300084>.
- [69] M. Hornberger, R. Spörl, and G. Scheffknecht, "Calcium Looping for CO₂ Capture in Cement Plants - Pilot Scale Test," *Energy Procedia*, vol. 114, no. November 2016, pp. 6171–6174, 2017, doi: 10.1016/j.egypro.2017.03.1754.
- [70] Hanak DP, Anthony EJ, Manovic V. A review of developments in pilot-plant testing and modelling of calcium looping process for CO₂ capture from power generation systems. *Energy Environ Sci* 2015;8(8):2199–249. <https://doi.org/10.1039/c5ee01228g>.
- [71] Arias B, et al. Operating Experience in la Pereda 1.7 MWth Calcium Looping Pilot. *Energy Procedia* 2017;114:149–57. <https://doi.org/10.1016/j.egypro.2017.03.1157>.
- [72] Ströhle J, Junk M, Kremer J, Galloy A, Epple B. Carbonate looping experiments in a 1 MWth pilot plant and model validation. *Fuel Jul.* 2014;127:13–22. <https://doi.org/10.1016/J.FUEL.2013.12.043>.
- [73] M. Helbig, J. Hiltz, M. Haaf, A. Daikeler, J. Ströhle, and B. Epple, "Long-term Carbonate Looping Testing in a 1 MWth Pilot Plant with Hard Coal and Lignite," *Energy Procedia*, vol. 114, no. November 2016, pp. 179–190, 2017, doi: 10.1016/j.egypro.2017.03.1160.
- [74] "CLEANKER, CLEAN clinKER production by calcium looping process (2022)." <http://www.cleanker.eu/the-project/project-contents> (accessed Nov. 08, 2022).
- [75] "C4U, Advanced Carbon Capture for steel industries integrated in CCUS Clusters (2022)." <https://c4u-project.eu/work-packages/> (accessed Nov. 08, 2022).
- [76] Luo M, Wang C, Yi Y, Liu K, Cai J, Wang Q. Power Generation from Coke Oven Gas Using Chemical Looping Combustion: Thermodynamic Simulation. *Chem Eng Technol* 2018;41(3):524–31. <https://doi.org/10.1002/ceat.201700322>.
- [77] Wang S, Wang G, Jiang F, Luo M, Li H. Chemical looping combustion of coke oven gas by using Fe₂O₃/CuO with MgAl₂O₄ as oxygen carrier. *Energy Environ Sci* 2010;3(9):1353–60. <https://doi.org/10.1039/b926193a>.
- [78] Bahzad H, Katayama K, Boot-Handford ME, Mac Dowell N, Shah N, Fennell PS. Iron-based chemical-looping technology for decarbonising iron and steel production. *Int J Greenh Gas Control* 2019;vol. 91, no. June:102766. <https://doi.org/10.1016/j.ijggc.2019.06.017>.
- [79] Xiang D, Zhao S. Parameter optimization and thermodynamic analysis of COG direct chemical looping hydrogen processes. *Energy Convers Manag* 2018;172 (July):1–8. <https://doi.org/10.1016/j.enconman.2018.07.007>.
- [80] Katayama K, Bahzad H, Boot-Handford M, Patzschke CF, Fennell PS. Process Integration of Chemical Looping Water Splitting with a Sintering Plant for Iron Making. *ACS Appl Mater Interfaces* 2020. <https://doi.org/10.1021/acs.iecr.9b05945>.
- [81] Xiang D, Zhou Y. Concept design and techno-economic performance of hydrogen and ammonia co-generation by coke-oven gas-pressure swing adsorption integrated with chemical looping hydrogen process. *Appl Energy* 2018;229 (April):1024–34. <https://doi.org/10.1016/j.apenergy.2018.08.081>.
- [82] Kevin J Whitty JoAnn S Lighty, "Development of Enabling Technologies for Chemical Looping Combustion and Chemical Looping with Oxygen Uncoupling,," *NETL Transform. Power Gener. Proj. Rev. Meet. 28 Sept. 2020.*, [Online]. Available: https://netl.doe.gov/sites/default/files/netl-file/20TPGVPR_Whitty.pdf.
- [83] I. Gogolev, A. H. Soleimanisalam, C. Linderholm, and A. Lyngfelt, "Commissioning, performance benchmarking, and investigation of alkali emissions in a 10 kWth solid fuel chemical looping combustion pilot," *Fuel*, doi: 10.1016/j.fuel.2020.119530.
- [84] Adánez J, Abad A, Mendiara T, Gayán P, de Diego LF, García-Labiano F. Chemical looping combustion of solid fuels. *Prog Energy Combust Sci* 2018;65:6–66. <https://doi.org/10.1016/j.pecs.2017.07.005>.
- [85] Lyngfelt A, Brink A, Langørgen Ø, Mattisson T, Rydén M, Linderholm C. 11,000 h of chemical-looping combustion operation—Where are we and where do we want to go? *Int J Greenh Gas Control* 2019;88(May):38–56. <https://doi.org/10.1016/j.ijggc.2019.05.023>.

- [86] Di Giuliano A, Capone S, Anatone M, Gallucci K. Chemical Looping Combustion and Gasification: A Review and a Focus on European Research Projects. *Ind Eng Chem Res* 2022. <https://doi.org/10.1021/acs.iecr.2c02677>.
- [87] "IFP. 2021. BUILDING OF THE LARGEST CHEMICAL LOOPING COMBUSTION (CLC) FACILITY TO DRIVE DOWN THE COST OF CARBON CAPTURE." <https://www.ifpenergiesnouvelles.com/article/building-largest-chemical-looping-combustion-clc-facility-drive-down-cost-carbon-capture> (accessed Nov. 08, 2022).
- [88] "INEA. INNOVATION AND NETWORKS EXECUTIVE AGENCY. CHEERS project. 2021." <https://ec.europa.eu/inea/en/horizon-2020/projects/h2020-energy/carbon-capture-storage-power-plants/cheers> (accessed Nov. 08, 2022).
- [89] "SINTEF. 2021. LOUISE – Low-cost CO2 capture by chemical looping combustion of waste-derived fuels." <https://www.sintef.no/en/projects/2021/louise-low-cost-co2-capture-by-chemical-looping-combustion-of-waste-derived-fuels/> (accessed Nov. 08, 2022).
- [90] Martínez I, et al. Recent progress of the Ca-Cu technology for decarbonisation of power plants and carbon intensive industries. *Int J Greenh Gas Control* Jun. 2019; 85:71–85. <https://doi.org/10.1016/j.jggc.2019.03.026>.
- [91] Fernández JR, Spallina V, Abanades JC. Advanced Packed-Bed Ca-Cu Looping Process for the CO2 Capture From Steel Mill Off-Gases. *Front Energy Res* 2020;8 (July):1–13. <https://doi.org/10.3389/fenrg.2020.00146>.
- [92] Tian S, Li K, Jiang J, Chen X, Yan F. CO2 abatement from the iron and steel industry using a combined Ca-Fe chemical loop. *Appl Energy* 2016;170:345–52. <https://doi.org/10.1016/j.apenergy.2016.02.120>.
- [93] Fernández JR, Martínez I, Abanades JC, Romano MC. Conceptual design of a Ca-Cu chemical looping process for hydrogen production in integrated steelworks. *Int J Hydrogen Energy* 2017;42(16):11023–37. <https://doi.org/10.1016/j.ijhydene.2017.02.141>.
- [94] Sun Z, Liu J, Sun Z. Synergistic decarbonization and desulfurization of blast furnace gas via a novel magnesium-molybdenum looping process. *Fuel* 2020;vol. 279, no. May:118418. <https://doi.org/10.1016/j.fuel.2020.118418>.
- [95] M. Bailera, P. Lisbona, B. Peña, and L. M. Romeo, "A review on CO2 mitigation in the Iron and Steel industry through Power to X processes," *J. CO2 Util.*, vol. 46, no. November 2020, p. 101456, 2021, doi: 10.1016/j.jcou.2021.101456.
- [96] Lisbona P, Bailera M, Peña B, Romeo LM. Integration of CO2 capture and conversion. In: Rahimpour MR, Farsi M, Makarem MA, editors. *Advances in Carbon Capture*. Woodhead Publishing; 2020. p. 503–22. <https://doi.org/10.1016/B978-0-12-819657-1.00022-0>.
- [97] Geerdes M, Chaigneau R, Lingardi O, R. van O. Molenaar, S. Y. R., and J. Warren., *Modern Blast Furnace Ironmaking An Introduction*. IOS Press; 2020.
- [98] Bailera M, Hanak DP, Lisbona P, Romeo LM. Techno-economic feasibility of power to gas-oxy-fuel boiler hybrid system under uncertainty. *Int J Hydrogen Energy* 2019;4:9505–16. <https://doi.org/10.1016/j.ijhydene.2018.09.131>.
- [99] Díez LI, Lupiáñez C, Guedea I, Bolea I, Romeo LM. Anthracite oxy-combustion characteristics in a 90 kWth fluidized bed reactor. *Fuel Process Technol* 2015; 139:196–203. <https://doi.org/10.1016/j.fuproc.2015.07.021>.
- [100] Sato M, Takahashi K, Nouchi T, Ariyama T. Prediction of next-generation ironmaking process based on oxygen blast furnace suitable for CO2 mitigation and energy flexibility. *ISIJ Int* 2015;55(10):2105–14. <https://doi.org/10.2355/isijinternational.ISIJINT-2015-264>.
- [101] Quader MA, Ahmed S, Raja Ghazilla RA, Ahmed S, Dahari M. Evaluation of criteria for CO2 capture and storage in the iron and steel industry using the 2-tuple DEMATEL technique. *J Clean Prod* May 2016;120:207–20. <https://doi.org/10.1016/J.JCLEPRO.2015.10.056>.
- [102] Ho MT, Bustamante A, Wiley DE. Comparison of CO2 capture economics for iron and steel mills. *Int J Greenh Gas Control* Nov. 2013;19:145–59. <https://doi.org/10.1016/J.IJGGC.2013.08.003>.
- [103] Jin P, Jiang Z, Bao C, Lu Y, Zhang J, Zhang X. Mathematical Modeling of the Energy Consumption and Carbon Emission for the Oxygen Blast Furnace with Top Gas Recycling. *Steel Res Int* 2016;87(3):320–9. <https://doi.org/10.1002/srin.201500054>.
- [104] Helle H, Helle M, Pettersson F, Saxén H. Multi-objective Optimization of Ironmaking in the Blast Furnace with Top Gas Recycling. *ISIJ Int* 2010;50(10): 1380–7. <https://doi.org/10.2355/isijinternational.50.1380>.
- [105] Helle H, Helle M, Saxén H, Pettersson F. Optimization of Top Gas Recycling Conditions under High Oxygen Enrichment in the Blast Furnace. *ISIJ Int* 2010;50 (7):931–8. <https://doi.org/10.2355/isijinternational.50.931>.
- [106] Arasto A, Tsupari E, Kärki J, Lilja J, Sihvonon M. Oxygen blast furnace with CO2 capture and storage at an integrated steel mill-Part I: Technical concept analysis. *Int J Greenh Gas Control* 2014;30:140–7. <https://doi.org/10.1016/j.jggc.2014.09.004>.
- [107] Onoda M, Matsuzaki Y, Chowdhury FA, Yamada H, Goto K, Tonomura S. Sustainable Aspects of Ultimate Reduction of CO2 in the Steelmaking Process (COURSE50 Project), Part 2: CO2 Capture. *J Sustain Metall* 2016;2(3):209–15. <https://doi.org/10.1007/s40831-016-0067-3>.
- [108] Murai R, Sato M, Ariyama T. Design of Innovative Blast Furnace for Minimizing CO2 Emission Based on Optimization of Solid Fuel Injection and Top Gas Recycling. *ISIJ Int* 2004;44(12):2168–77. <https://doi.org/10.2355/isijinternational.44.2168>.
- [109] Takahashi K, Nouchi T, Sato M, Ariyama T. Perspective on Progressive Development of Oxygen Blast Furnace for Energy Saving. *ISIJ Int* 2015;55(9): 1866–75. <https://doi.org/10.2355/isijinternational.ISIJINT-2015-196>.
- [110] Sahu RK, Roy SK, P. K. Sen A. "Applicability of Top Gas Recycle Blast Furnace with Downstream Integration and Sequestration in an Integrated Steel Plant" 2015;1(5):502–16. <https://doi.org/10.1002/srin.201400196>.
- [111] Ariyama T, Sato M, Nouchi T, Takahashi K. Evolution of blast furnace process toward reductant flexibility and carbon dioxide mitigation in steel works. *ISIJ Int* 2016;56(10):1681–96. <https://doi.org/10.2355/isijinternational.ISIJINT-2016-210>.
- [112] Wang H, et al. Mathematical Simulation on Blast Furnace Operation of Coke Oven Gas Injection in Combination with Top Gas Recycling. *Steel Res Int* 2016;87(5): 539–49. <https://doi.org/10.1002/srin.201500372>.
- [113] Tseitlin MA, Lazutkin SE, Styopin GM. A Flow-chart for Iron Making on the Basis of 100% Usage of Process Oxygen and Hot Reducing Gases Injection. *ISIJ Int* 1994;34(7):570–3. <https://doi.org/10.2355/isijinternational.34.570>.
- [114] Helle M, Saxén H. Operation windows of the oxygen blast furnace with top gas recycling. *ISIJ Int* 2015;55(10):2047–55. <https://doi.org/10.2355/isijinternational.ISIJINT-2015-083>.
- [115] van der Stel J, Louwerse G, Sert D, Hirsch A, Eklund N, Pettersson M. Top gas recycling blast furnace developments for 'green' and sustainable ironmaking. *Ironmak Steelmak* Oct. 2013;40(7):483–9. <https://doi.org/10.1179/0301923313Z.000000000221>.
- [116] Nogami H, Chu M, Yagi J-I. Numerical analysis of blast furnace operations with top gas recycling. *Rev Metallurgie Mar.* 2005;102(3):189–97. <https://doi.org/10.1051/metal:2005132>.
- [117] Zhou Z, Yi Q, Wang R, Wang G, Ma C. Numerical Investigation on Coal Combustion in Ultralow CO2 Blast Furnace: Effect of Oxygen Temperature. *Processes* Jul. 2020;8(7):877. <https://doi.org/10.3390/pr8070877>.
- [118] Helle H, Helle M, Saxén H. Nonlinear optimization of steel production using traditional and novel blast furnace operation strategies. *Chem Eng Sci* Dec. 2011; 66(24):6470–81. <https://doi.org/10.1016/j.ces.2011.09.006>.
- [119] Jin P, Jiang Z, Bao C, Hao S, Zhang X. The energy consumption and carbon emission of the integrated steel mill with oxygen blast furnace. *Resour Conserv Recycl* 2017;117:58–65. <https://doi.org/10.1016/j.resconrec.2015.07.008>.
- [120] Perpiñán J, Bailera M, Romeo LM, Peña B, Eveloy V. CO2 Recycling in the Iron and Steel Industry via Power-to-Gas and Oxy-Fuel Combustion. *Energies* 2021;14 (21):7090. <https://doi.org/10.3390/en14217090>.
- [121] M. Bailera, T. Nakagaki, and R. Kataoka, "Revisiting the Rist diagram for predicting operating conditions in blast furnaces with multiple injections," 2021, doi: 10.12688/openreseurope.14275.1.
- [122] Oda J, Akimoto K, Sano F, Tomoda T. Diffusion of energy efficient technologies and CO2 emission reductions in iron and steel sector. *Energy Econ* 2007;29(4): 868–88. <https://doi.org/10.1016/j.eneco.2007.01.003>.
- [123] M. A. A.-M. Rainer Remus and L. D. S. Serge Roudier, *Best Available Techniques (BAT) Reference Document for Iron and Steel Production*. 2013. doi: doi:10.2791/98516.
- [124] Wu J, Wang R, Pu G, Qi H. Integrated assessment of exergy, energy and carbon dioxide emissions in an iron and steel industrial network. *Appl Energy* 2016;183: 430–44. <https://doi.org/10.1016/j.apenergy.2016.08.192>.
- [125] Suzuki K, Hayashi K, Kuribara K, Nakagaki T, Kasahara S. Quantitative evaluation of CO2 emission reduction of active carbon recycling energy system for ironmaking by modeling with aspen plus. *ISIJ Int* 2015;55(2):340–7. <https://doi.org/10.2355/isijinternational.55.340>.
- [126] A. F. Prego et al., "Operación de hornos altos con inyección de carbón pulverizado en diferentes condiciones tecnológicas." *Rev. Metal.*, vol. 37, pp. 423–436, 2001, doi: 10.3989/revmetal.2001.v37.i3.50810.3989/revmetal.2001.v37.i3.508.
- [127] Zhang H, Li H, Tang Q, Bao W. Conceptual design and simulation analysis of thermal behaviors of TGR blast furnace and oxygen blast furnace. *Sci China Technol Sci* 2010;53(1):85–92. <https://doi.org/10.1007/s11431-010-0029-0>.
- [128] Zhang W, Zhang J, Xue Z, Zou Z, Qi Y. Unsteady Analyses of the Top Gas Recycling Oxygen Blast Furnace. *ISIJ Int* 2016;56(8):1358–67. <https://doi.org/10.2355/isijinternational.ISIJINT-2016-090>.
- [129] She X, An X, Wang J, Xue Q, Kong L. Numerical analysis of carbon saving potential in a top gas recycling oxygen blast furnace. *J Iron Steel Res Int Jun.* 2017;24(6):608–16. [https://doi.org/10.1016/S1006-706X\(17\)30092-4](https://doi.org/10.1016/S1006-706X(17)30092-4).
- [130] Zhang W, Xue Z, Zhang J, Wang W, Cheng C, Zou Z. Medium oxygen enriched blast furnace with top gas recycling strategy. *J Iron Steel Res Int Aug.* 2017;24(8): 778–86. [https://doi.org/10.1016/S1006-706X\(17\)30117-6](https://doi.org/10.1016/S1006-706X(17)30117-6).
- [131] Danloy G, et al. ULCOS - Pilot testing of the low-CO2 Blast Furnace process at the experimental BF in Lulea. *Rev Metall Cah D'Informations Tech* 2009;106(1):1–8. <https://doi.org/10.1051/metal/2009008>.
- [132] Bargiacchi E, Candelaresi D, Valente A, Spazzafumo G, Frigo S. Life Cycle Assessment of Substitute Natural Gas production from biomass and electrolytic hydrogen. *Int J Hydrogen Energy* 2021;46(72):35974–84. <https://doi.org/10.1016/j.ijhydene.2021.01.033>.
- [133] Faria DG, Carvalho MMO, Neto MRV, de Paula EC, Cardoso M, Vakkilainen EK. Integrating oxy-fuel combustion and power-to-gas in the cement industry: A process modeling and simulation study. *Int J Greenh Gas Control* Feb. 2022;114: 103602. <https://doi.org/10.1016/j.jggc.2022.103602>.
- [134] M. Bailera, T. Nakagaki, and R. Kataoka, "Limits on the integration of power to gas with blast furnace ironmaking," *J. Clean. Prod.*, p. 134038, Sep. 2022, doi: 10.1016/J.JCLEPRO.2022.134038.
- [135] W. Zhang, J. Dai, C. Li, X. Yu, Z. Xue, and H. Saxén, "A Review on Explorations of the Oxygen Blast Furnace Process," *Steel Res. Int.*, vol. 92, no. 1, 2021, doi: 10.1002/srin.202000326.
- [136] Lan R, Wang J, Han Y, She X, Wang L, Qing-guo X. Reduction Behavior of Sinter Based on Top Gas Recycling-Oxygen Blast Furnace. *J Iron Steel Res Int Sep.* 2012; 19(9):13–9. [https://doi.org/10.1016/S1006-706X\(13\)60003-5](https://doi.org/10.1016/S1006-706X(13)60003-5).
- [137] Chu M, Nogami H, Yagi J. Numerical Analysis on Blast Furnace Performance under Operation with Top Gas Recycling and Carbon Composite Agglomerates

- Charging. ISIJ Int 2004;44(12):2159–67. <https://doi.org/10.2355/isijinternational.44.2159>.
- [138] Babich A. Blast furnace injection for minimizing the coke rate and CO₂ emissions. *Ironmak Steelmak Jul.* 2021;48(6):728–41. <https://doi.org/10.1080/03019233.2021.1900037>.
- [139] Li Z, et al. Numerical Investigation of Novel Oxygen Blast Furnace Ironmaking Processes. *Metall Mater Trans B Aug.* 2018;49(4):1995–2010. <https://doi.org/10.1007/s11663-018-1259-y>.
- [140] R. Kumar Sahu, C. Halder, and P. K. Sen, "Optimization of Top Gas Recycle Blast Furnace Emissions with Implications of Downstream Energy," *steel Res. Int.*, vol. 87, no. 9, pp. 1190–1202, Sep. 2016, doi: 10.1002/srin.201500312.
- [141] Ohno Y, Matsuura M, Mitsufuji H, Furukawa T. Process Characteristics of a Commercial-scale Oxygen Blast Furnace Process with Shaft Gas Injection. *ISIJ Int* 1992;32(7):838–47. <https://doi.org/10.2355/isijinternational.32.838>.
- [142] Sen PK. Co₂ Accounting and Abatement: An Approach for Iron and Steel Industry. *Trans Indian Inst Met Dec.* 2013;66(5–6):711–21. <https://doi.org/10.1007/s12666-013-0285-3>.
- [143] Liu L, et al. Effects of top gas recycling on in-furnace status, productivity, and energy consumption of oxygen blast furnace. *Energy Nov.* 2018;163:144–50. <https://doi.org/10.1016/j.energy.2018.08.114>.
- [144] Qi Y, Yan D, Gao J, ZHANG J, LI M. Study on Industrial test of the Oxygen Blast Furnace. *Iron Steel* 2011;36–8. <https://doi.org/10.13228/j.boyuan.issn0449-749x.2011.03.013>.
- [145] Zuo G, Hirsch A. The trial of the Top Gas Recycling Blast Furnace at LKAB's EBF and Scale-Up. *La Rev Métallurgie* 2009;2(5):255.
- [146] Tsupari E, Kärki J, Arasto A, Lilja J, Kinnunen K, Sihvonen M. Oxygen blast furnace with CO₂ capture and storage at an integrated steel mill - Part II: Economic feasibility in comparison with conventional blast furnace highlighting sensitivities. *Int J Greenh Gas Control* 2015;32:189–96. <https://doi.org/10.1016/j.jggc.2014.11.007>.
- [147] Austin PR, Nogami H, Yagi J. Prediction of Blast Furnace Performance with Top Gas Recycling. *ISIJ Int* 1998;38(3):239–45. <https://doi.org/10.2355/isijinternational.38.239>.
- [148] D. H. B. Andrade, R. P. Tavares, A. C. B. Quintas, V. E. de S. Moreira, A. O. Viana, and V. M. Gasparini, "Evaluation of the permeability of the dripping zone and of flooding phenomena in a blast furnace," *J. Mater. Res. Technol.*, vol. 8, no. 1, pp. 134–139, Jan. 2019, doi: 10.1016/j.jmrt.2017.10.001.
- [149] Marketwatch. *NYSE American, Steel Index 2022*.
- [150] Kuramochi T, Ramírez A, Turkenburg W, Faaij A. Comparative assessment of CO₂ capture technologies for carbon-intensive industrial processes. *Prog Energy Combust Sci Feb.* 2012;38(1):87–112. <https://doi.org/10.1016/j.pecs.2011.05.001>.
- [151] Tobiesen FA, Svendsen HF, Mejdell T. Modeling of blast furnace CO₂ capture using amine absorbents. *Ind Eng Chem Res* 2007;46(23):7811–9. <https://doi.org/10.1021/ie061556j>.
- [152] Kidnay AJ, Paarish WR, Mccartney DG. *Fundamentals of Natural Gas Processing*. 3rd ed. Boca Raton: CRC Press; 2019.
- [153] Martínez Castilla G, Biermann M, Montañés RM, Normann F, Johnsson F. Integrating carbon capture into an industrial combined-heat-and-power plant: performance with hourly and seasonal load changes. *Int J Greenh Gas Control Mar.* 2019;82:192–203. <https://doi.org/10.1016/j.jggc.2019.01.015>.
- [154] xun Fu J, hao Tang G, jie Zhao R, sing Hwang W. Carbon Reduction Programs and Key Technologies in Global Steel Industry. *J Iron Steel Res Int Mar.* 2014;21(3): 275–81. [https://doi.org/10.1016/S1006-706X\(14\)60042-X](https://doi.org/10.1016/S1006-706X(14)60042-X).
- [155] Han K, Ahn CK, Lee MS. Performance of an ammonia-based CO₂ capture pilot facility in iron and steel industry. *Int J Greenh Gas Control* 2014;27:239–46. <https://doi.org/10.1016/j.jggc.2014.05.014>.
- [156] Quader MA, Ahmed S, Ghazilla RAR, Ahmed S. CO₂ Capture and Storage for the iron and steel manufacturing industry: Challenges and Opportunities. *J Appl Sci Agric* 2014;9(21):60–7.
- [157] Parla JCM, Hendriks CA, Blok K. Carbon dioxide recovery from industrial processes. *Clim Change* 1995;29(4):439–61. <https://doi.org/10.1007/BF01092428>.
- [158] Kim H, Lee J, Lee S, Lee IB, J. hyoung Park, and J. Han, "Economic process design for separation of CO₂ from the off-gas in ironmaking and steelmaking plants. *Energy Aug.* 2015;88:756–64. <https://doi.org/10.1016/J.ENERGY.2015.05.093>.
- [159] Onarheim K, Arasto A. Staged implementation of alternative processes in an existing integrated steel mill for improved performance and reduced CO₂ emissions – Part I: Technical concept analysis. *Int J Greenh Gas Control Feb.* 2016;45:163–71. <https://doi.org/10.1016/J.JGGC.2015.12.008>.
- [160] Chung W, Roh K, Lee JH. Design and evaluation of CO₂ capture plants for the steelmaking industry by means of amine scrubbing and membrane separation. *Int J Greenh Gas Control Jul.* 2018;74:259–70. <https://doi.org/10.1016/J.JGGC.2018.05.009>.
- [161] Hisashige S, Nakagaki T, Yamamoto T. CO₂ emission reduction and exergy analysis of smart steelmaking system adaptive for flexible operating conditions. *ISIJ Int* 2019;59(4):598–606. <https://doi.org/10.2355/isijinternational.ISIJINT-2018-355>.
- [162] Ramírez-Santos AA, Bozorg M, Addis B, Piccialli V, Castel C, Favre E. Optimization of multistage membrane gas separation processes. Example of application to CO₂ capture from blast furnace gas. *J Memb Sci* 2018;566. <https://doi.org/10.1016/j.memsci.2018.08.024>.
- [163] J.-P. Birat, "Carbon dioxide (CO₂) capture and storage technology in the iron and steel industry," *Dev. Innov. Carbon Dioxide Capture Storage Technol.*, pp. 492–521, 2010, doi: 10.1533/9781845699574.5.492.
- [164] Saima WH, Mogi Y, Haraoka T. Development of PSA System for the Recovery of Carbon Dioxide and Carbon Monoxide from Blast Furnace Gas in Steel Works. *Energy Procedia* 2013;37:7152–9. <https://doi.org/10.1016/j.egypro.2013.06.652>.
- [165] Y. Ujisawa, S. Tonomura, N. Ishiwata, Y. Nabeshima, and K. Saito, "Topic: CO₂ breakthrough program by COURSE50 in Japanese steel industry sector," in *Energy Technology Roadmaps of Japan: Future Energy Systems Based on Feasible Technologies Beyond 2030*, 2016. doi: 10.1007/978-4-431-55951-1_29.
- [166] Rao M, Fernandes A, Pronk P, Aravind PV. Design, modelling and techno-economic analysis of a solid oxide fuel cell-gas turbine system with CO₂ capture fueled by gases from steel industry. *Appl Therm Eng* 2019;148. <https://doi.org/10.1016/j.applthermaleng.2018.11.108>.
- [167] A. B. Rao, E. S. Rubin, and M. B. Berkenpas, "AN INTEGRATED MODELING FRAMEWORK FOR CARBON MANAGEMENT TECHNOLOGIES," Pittsburgh, PA, 2004.
- [168] Ho MT, Allinson GW, Wiley DE. Reducing the Cost of CO₂ Capture from Flue Gases Using Pressure Swing Adsorption. *Ind Eng Chem Res Jul.* 2008;47(14): 4883–90. <https://doi.org/10.1021/ie070831e>.
- [169] Zuo G, Hirsch A. The trial of the top gas recycling blast furnace at LKABsEBF and scale-up. *Rev Metall Cah D'Informations Tech* 2009;106(9):pp. <https://doi.org/10.1051/metal/2009067>.
- [170] Wang X, Wang T. Hydrogen amplification from coke oven gas using a CO₂ adsorption enhanced hydrogen amplification reactor. *Int J Hydrogen Energy* 2012;37(6):pp. <https://doi.org/10.1016/j.ijhydene.2011.12.018>.
- [171] Siqueira RM, et al. Carbon Dioxide Capture by Pressure Swing Adsorption. *Energy Procedia Jul.* 2017;114:2182–92. <https://doi.org/10.1016/J.EGYPRO.2017.03.1355>.
- [172] Hasan MMF, Baliban RC, Elia JA, Floudas CA. Modeling, Simulation, and Optimization of Postcombustion CO₂ Capture for Variable Feed Concentration and Flow Rate. 1. Chemical Absorption and Membrane Processes. *Ind Eng Chem Res Dec.* 2012;51(48):15642–64. <https://doi.org/10.1021/ie301571d>.
- [173] Ramírez-Santos AA, Castel C, Favre E. Utilization of blast furnace flue gas: Opportunities and challenges for polymeric membrane gas separation processes. *J Memb Sci* 2017;526. <https://doi.org/10.1016/j.memsci.2016.12.033>.
- [174] Lie JA, Vassbotn T, Hägg MB, Grainger D, Kim TJ, Mejdell T. Optimization of a membrane process for CO₂ capture in the steelmaking industry. *Int J Greenh Gas Control* 2007;1(3):pp. [https://doi.org/10.1016/S11750-5836\(07\)00069-2](https://doi.org/10.1016/S11750-5836(07)00069-2).
- [175] Jeon J-Y, Park B-R, Kim J-H. Numerical Simulation and Optimization of 4-Component LDG Separation in the Steelmaking Industry Using Polysulfone Hollow Fiber Membranes. *Membranes* 2022;12(1):pp. <https://doi.org/10.3390/membranes12010097>.
- [176] Roy P, Das N. A lithium–aluminosilicate zeolite membrane for separation of CO₂ from simulated blast furnace gas. *J Porous Mater* 2020;27(3):pp. <https://doi.org/10.1007/s10934-019-00849-0>.
- [177] Baker RW, Freeman B, Knief J, Wei X, Merkel T. CO₂ capture from natural gas power plants using selective exhaust gas recycle membrane designs. *Int J Greenh Gas Control* 2017;66:35–47. <https://doi.org/10.1016/j.jggc.2017.08.016>.
- [178] Li R, Lian S, Zhang Z, Song C, Han R, Liu Q. Techno-economic evaluation of a novel membrane-cryogenic hybrid process for carbon capture. *Appl Therm Eng* 2022;200. <https://doi.org/10.1016/j.applthermaleng.2021.117688>.
- [179] Van Dijk HAJ, et al. "Stepwise project: Sorption-enhanced water-gas shift technology to reduce carbon footprint in the iron and steel industry", *Johnson Matthey Technol Rev* 2018;62(4):pp. <https://doi.org/10.1595/205651318X15268923666410>.
- [180] Lukashuk L, et al. A new application of the commercial high temperature water gas shift catalyst for reduction of CO₂ emissions in the iron and steel industry: Lab-scale catalyst evaluation. *Int J Hydrogen Energy* 2021;46(79):39023–35. <https://doi.org/10.1016/j.ijhydene.2021.09.203>.
- [181] "Stepwise Project." <https://www.stepwise.eu/project/>.
- [182] Choi JO, Kim JG. CO₂ reduction by blast furnace top gas recycling combined with waste hydrocarbon gasification. *Stud Surf Sci Catal* 2004;153:125–30. [https://doi.org/10.1016/S0167-2991\(04\)80230-1](https://doi.org/10.1016/S0167-2991(04)80230-1).
- [183] Li Z, et al. Numerical study and design strategy for a low emission coke oven system using oxy-fuel combustion of coke oven gas. *J Clean Prod* 2020;vol. 252, no. x:119656. <https://doi.org/10.1016/j.jclepro.2019.119656>.
- [184] Qie Y, Lyu Q, Lan C, Zhang S. Energy Conservation and CO₂ Abatement Potential of a Gas-injection Blast Furnace. *High Temp Mater Process* 2020;39(1):96–106. <https://doi.org/10.1515/htmp-2020-0031>.
- [185] Zhang W, Zhang J, Xue Z. Exergy analyses of the oxygen blast furnace with top gas recycling process. *Energy* 2017;121:135–46. <https://doi.org/10.1016/j.energy.2016.12.125>.
- [186] Chu M, Yagi JI. Numerical evaluation of blast furnace performance under top gas recycling and lower temperature operation. *Steel Res Int* 2010;81(12):1043–50. <https://doi.org/10.1002/srin.201000103>.
- [187] H. Takashi, Y. Mogi, and H. Saima, "PSA System for the Recovery of Carbon Dioxide from Blast Furnace Gas in Steel Works The Influence of Operation Conditions on CO₂ Separation," 2013, doi: 10.1252/kakoronbunshu.39.439.