



Technical and economic assessment of iron and steelmaking decarbonization via power to gas and amine scrubbing

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ABSTRACT

The iron and steel industry is one of the most energy-intensive industries, emitting 5% of the total anthropogenic carbon dioxide (CO₂). The control of CO₂ emissions has become increasingly stringent in the European Union (EU), resulting in EU allowance above 90 €/tCO₂. Carbon capture will be required to achieve CO₂ emissions control, and carbon utilization via power-to-gas could significantly increase interest in carbon capture in the iron and steel sector. This paper presents a new concept that combines amine scrubbing with power-to-gas to reduce emissions in blast furnace-basic oxygen furnace steelmaking plants. Synthetic natural gas (SNG) is produced using green hydrogen from water electrolysis and CO₂ from steelmaking. The synthetic natural gas is later used as a reducing agent in the blast furnace, constantly recycling carbon in a closed loop and avoiding geological storage. The oxygen by-produced via electrolysis eliminates the necessity of an air separation unit. By applying these innovations to steelmaking, a reduction in CO₂ emissions of 9.4% is obtained with an energy penalty of 16.2 MJ/kgCO₂, and economic costs of 52 €/t_{HM} or 283 €/tCO₂. A sensitivity analysis with respect to electricity and the CO₂ allowances prices is also performed.

1. Introduction

The current European targets on global warming include the reduction of greenhouse gas (GHG) emissions by 80–95%, compared to 1990 levels, by the year 2050 [1]. One of the largest industrial emitters of GHG emissions, in particular CO₂, is the steel sector. Iron and steel making processes are still highly dependent on fossil fuels, thereby releasing a substantial amount of CO₂ [2]. These emissions accounted for 223 MtCO₂-eq in the European Union in 2010 [3], i.e. 5% of the total GHG emissions of the EU-27 [4]. Steel is mainly produced by two routes: (i) blast furnace–basic oxygen furnace (BF–BOF) with 70% market share, and (ii) electric arc furnace (EAFs) with 30% market share [4].

The BF–BOF route, which uses coal as the main primary energy resource, mainly consists of the following processes: sinter strand, coke oven, blast furnace (BF), basic oxygen furnace (BOF), and the final stage of casting and rolling. The sintering is used to agglomerate iron ore. The coke oven allows obtaining coke from coal. In the BF, the iron ore is reduced by coke, obtaining hot metal (HM). Then, the BOF lowers the carbon content of the molten iron to produce crude steel (CS) which

finally passes through casting and rolling stages to obtain the desired product. Along the involved processes, different exhaust gases are obtained, which can be used as fuel in the steel plant (coke oven gas - COG, blast furnace gas - BFG, and basic oxygen furnace gas - BOFG). BF–BOF is highly energy- and carbon-intense, resulting in a net energy consumption of 13–14 GJ per ton of crude steel (GJ/t_{CS}) and specific emissions of 2200 kgCO₂/t_{CS} [5].

The EAF route produces steel by melting recycled scrap using electricity. When the availability of scrap is limited, direct reduced iron (DRI) can also be used as feedstock. The DRI contains >90% metallic iron, and it is produced in combustion-free reactors by using natural gas or coal-based syngas as reducing agents. The dominant commercially available technologies, Energiron and Midrex [6], use iron pellets in shaft furnaces with countercurrent moving beds. Depending on the scrap-metallic iron share in the DRI–EAF, the energy consumption and CO₂ emissions are significantly lower, 4–10 GJ/t_{CS} and 400–1300 kgCO₂/t_{CS}, respectively, than for the BF–BOF route [5]. Substitution of natural gas or coal-based syngas with hydrogen can further abate DRI–EAF emissions significantly [7].

However, since the global steel demand cannot be covered through

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Nomenclature*Abbreviation Meaning*

AFT	Adiabatic Flame Temperature
ASU	Air Separation Unit
AHF	Air Heating Furnace
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
COG	Coke Oven Gas
DRI	Direct Reduced Iron
EAF	Electric Arc Furnace

GHG	Green House Gas
HM	Hot Metal
I&S	Iron and Steel
IEA	International Energy Agency
IRR	Internal Rate of Return
MDEA	Methyldiethanolamine
PtG	Power to Gas
PCI	Pulverised Coal Injection
PFD	Process Flow Diagram
PPAs	Power purchase agreements
SNG	Synthetic Natural Gas
tHM	Ton of Hot Metal
tCS	Ton of Crude Steel

recycled scrap, the BF-BOF route will maintain its dominance in the market, despite its high emissions. Besides, blast furnaces will only phased-out at relining, which typically takes place every 20–35 years, or up to 40 years for newly commissioned plants. Thus, it is expected that at least 20% of today's blast furnaces will still be in operation by year 2050 [8].

The need to reduce greenhouse gas emissions from steel production has gained significant attention globally, leading to widespread research into the development of alternative ultra-low-CO₂ steel production technologies. To this end, numerous projects and programs have been launched, such as Carbon2Chem [9], which explores potential uses for steel gases, and ULCORED [10], ULCOWIN [10], ULCOLYSIS [10], COREX [11], Hismelt [12], and FINEX® [13], which focus on the creation of new pathways for steel production. In addition to these projects, MIDREX® [5] and ENERGIRON® [6] processes are also being developed as alternative methods for steel production that emit fewer greenhouse gases. These processes utilize direct reduction technology to reduce iron ore (DRI) and eliminate the need for coke in the production process. Despite the challenges associated with developing these new technologies, the potential benefits to the environment and the steel industry are considerable, and these initiatives hold great promise for the future of sustainable steel production. However, out of all these projects, only ULCOS-BF [10] and COURSE50 [14] are dedicated to capturing CO₂ in the BF-BOF route through physical adsorption and chemical absorption methods from the BFG.

Carbon capture with amines is a well-established technology used in many industries to reduce greenhouse gas emissions. In the iron and steel industry, amine scrubbing is widely considered one of the most effective and practical approach for capturing CO₂ emissions from blast furnace gases. Despite its effectiveness, the use of amine scrubbing for carbon capture in the iron and steel industry still faces several challenges. One of the main challenges is the high energy consumption associated with the regeneration of the solvent, ranging between 2.9 MJ/kgCO₂ and 6.5 MJ/kgCO₂ [10], which can significantly increase the overall energy consumption of the process. Although the CO₂ capture process using amine solvents is well-established (TRL 9) and commercially available for use in natural gas and fertilizer processing plants [11], it has not been fully deployed in the iron and steel industry, resulting in a TRL2-TRL6 level [10].

Power to Gas (PtG) stands out as a promising solution to substantially reduce CO₂ emissions while providing additional benefits [12]. The Power to X concept involves processes that convert renewable electricity into valuable products, using an electrolysis stage to obtain renewable H₂ as an intermediate or final energy carrier. The hydrogen produced can be combined with captured CO₂ in a methanation reactor to obtain methane (Power to Methane, PtM). This synthetic natural gas can substitute a fraction of the currently consumed coke as a reducing agent in the blast furnace [13,14]. The application of Power to Gas to the Iron

and Steel industry can result in benefits accompanying the energy penalty of carbon capture, contrarily to other carbon reduction measures such as chemical absorption or physical adsorption. When comparing with CCS alone, underground geological storage would be required, along with transportation and compression. Furthermore, geological storage requires available site near the CO₂ source, but the suitable locations are very scarce. In addition, this is a controversial issue and the social acceptance is quite low.

Only a few studies have been found in the literature on PtG and carbon capture integration to reduce steelmaking emissions. Hisashige et al. [15] studied a shaft furnace coupled with a pressure swing adsorption stage (for carbon capture and CH₄ separation) and a CH₄ synthesis reactor (at 300 °C and 5.3 bar), injecting 33 kg_{SNG}/t_{HM} in the BF. Hisashige et al. found a 13% of CO₂ emission reduction in the blast furnace. Rosenfeld et al. [16] analysed the integration of Power to Methane, together with biomass gasification (mainly for further H₂ supply), concluding with a 14–19% of CO₂ emission reduction in the blast furnace. Perpiñán et al. [13] investigated the combination of power to gas, top gas recycling and oxy-blast furnace, achieving an 8% decrease in CO₂ emissions for the whole BF-BOF plant, accounting for 65 kg_{SNG}/t_{CS} injected in the BF. Bailera et al. [14,17,18] developed a detailed model of the blast furnace according to the Rist Diagram and studied a new concept proposal that combines oxygen blast furnaces with Power to Gas technology. The reduction of CO₂ emission obtained in that study ranged between 6% and 7% in the blast furnace, with an injection of 22 kg_{SNG}/t_{HM}.

The objective of this paper is to assess from the technical and economic point of view a novel concept that integrates PtM with amine scrubbing in the iron and steel industry to reduce the emission intensity of the conventional BF-BOF process. This is achieved using detailed process models of a complete conventional BF-BOF process plant and of the proposed PtG-based concept developed using Aspen Plus and validated using reference data. Both the conventional steelmaking process and the novel process that integrates PtM and amine scrubbing are compared in terms of demand for coal, overall energy consumption and CO₂ emissions. The influence of the amount of synthetic natural gas (SNG) injected to the BF on the adiabatic flame temperature (AFT) is analysed. Sankey diagrams for the energy and carbon flows are also presented and discussed to quantify the contribution of each process to energy consumption and carbon emission/recycling. Lastly, the economic feasibility of the proposed PtM and carbon capture integration is evaluated, to identify technical and economic conditions to reach reasonable pay-back periods and internal rate of returns (IRR) for the proposed steelmaking emission abatement process. The process models developed in this study can also be used to evaluate alternative emission reduction strategies for BF-BOF steelmaking processes.

Both the novel and conventional steelmaking processes are described in Section 2, and their modelling methodology in Section 3.

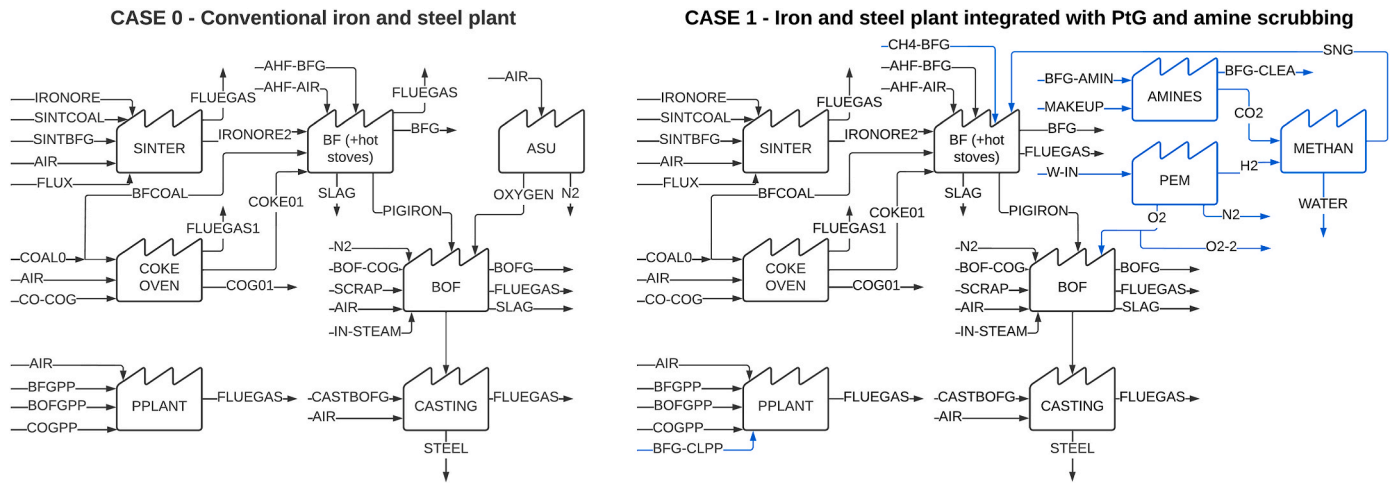


Fig. 1. Process flow diagram in Aspen Plus: (left) conventional iron and steel plant (Case 0), (right) its integration with power to gas and carbon capture (Case 1). Sub-systems drawn using blue coloured lines are the main innovations of the modified process (Case 1).

Table 1

Input and Output data of the Aspen Plus models for the proposed low-carbon steelmaking process concept and conventional BF-BOF process. Stream ID corresponding to Fig. 1.

Hierarchy	Stream ID	Composition	Flow	Temperature
SINTER	Iron ore	Input	Output	Input
SINTER	Coal	Input	Output	Input
SINTER	Flux	Input	Output	Input
SINTER	Iron ore (out)	Input	Output	Input
COKE OVEN	Coal	Input	Output	Input
COKE OVEN	Coke	Output	Output	Input ^{a)}
COKE OVEN	COG (out)	Output	Output	Output
BF	Coal	Input	Input	Input
BF	Air	Input	Output	Input
BF	BFG	Output	Output	Output
BF	Slag	Output	Output	Output
BF	Pig iron	Input	Input	Input
BOF	Scrap	Input	Input	Input
BOF	Oxygen	Output	Output	Input
BOF	N2	Input	Input	Input
BOF	Slag	Output	Output	Input
BOF	BOFG	Input	Output	Input
CASTING	Steel	Input	Output	Input
AMINES	BFG	Output	Output	Input
AMINES	BFG-CLEAN	Output	Output	Input
AMINES	CO2	Output	Output	Input
PEM	Water	Input	Output	Input
PEM	H2	Output	Output	Input
ASU/PEM	O2	Output	Output	Input
METHAN	SNG	Output	Output	Input

^{a)} O/I: Output/Input. Coke is calculated as an output in Case 0 and is maintained as an input in Case 1.

2. Process description

A conventional iron and steel plant producing 320 t_{HM}/h [19] is assumed as base case (Case 0) according to the process flow diagram (PFD) of Fig. 1. This reference plant consists of a sinter strand, coke oven, blast furnace (BF, including hot stoves), air separation unit (ASU), basic oxygen furnace (BOF), the casting-rolling stage and a combined cycle power plant. The mass flow inputs to the system are iron ore, coal, air, scrap, and flux, Fig. 1. Some processes are supplied also with COG, BFG, or BOFG for producing heat at high temperatures (streams BFG in the sinter strand, COG in the coke oven, BFG in the hot stoves, and COG in the basic oxygen furnace). The power plant generates electricity to be self-sufficient, consuming part of the available COG, BFG and BOFG (100 MW_e net power [20]). The remaining gases are supposed to be sold to nearby industries.

The second analysed concept is a PtG-steelmaking plant (Case 1) which adds an amine scrubbing to capture CO₂ from the BFG, a PEM electrolyser to produce H₂, and a methanation plant to produce synthetic natural gas. The SNG is injected into the BF to continuously recycle the CO₂ and to reduce coal consumption. The O₂ from the electrolyser is used in part to cover the demand in the BOF, avoiding the expensive ASU, while the surplus is sold to interested industries. The resulting clean gas from the amine scrubbing (stream BFG-CLEAN) is derived to the power plant for power production. In this paper, Case 1 is assessed in comparison with Case 0 in terms of energy requirements, CO₂ emission reduction and economic feasibility.

The use of PtG (SNG) as opposed to PtH₂ in the proposed process concept (Case 1) offers the following advantages. Although the addition of the methanation step to water electrolysis in general process incurs additional energy losses, the injection of methane is a demonstrated technology in the iron and steel sector [21,22], whereas the injection of H₂ in the BF is relatively new. Additionally, H₂ injection has been found to reduce the adiabatic flame temperature (AFT) of the BF at a much faster rate than CH₄ injection does [18], which means that for the same AFT, more CH₄ than H₂ can be injected, thereby achieving higher coke consumption reductions. Furthermore, H₂ has also a lower coke replacement ratio than CH₄, which means that less coke can be saved by injecting H₂. Moreover, when adopting the SNG route, part of the CO₂ is converted to SNG, instead of being emitted into the atmosphere, achieving also higher CO₂ reductions. Furthermore, thanks to the heat integration applied in this study, the energy rejected by the methanation process can provide the heat requirement of the carbon capture system (amine scrubbing), with no extra fuel requirement.

3. Modelling methodology

Models of Case 0 and Case 1 were implemented in Aspen Plus v11 using the PENG-ROB property method for iron and steel processes and the ASU, the ELECNRTL property method for amines and electrolyser, the PR-BM property method for the combined cycle power plant, and the SRK property method for methanation [23]. In the case of non-conventional solids (i.e., coal), HCOALGEN and DCOALIGT property models were used with the following correlations [24]: (i) Revised IGT correlation for the heat of combustion [25], (ii) Direct correlation for the heat of formation, (iii) Kirov correlation for the specific heat capacity, and (iv) IGT correlations for density on a dry basis. In addition, user-defined heat capacities from NIST database were used for different chemical components (instead of the data provided by the Aspen Plus database) to increase the valid range of application up to 2700 K. Simulations were performed under steady-state conditions, chemical

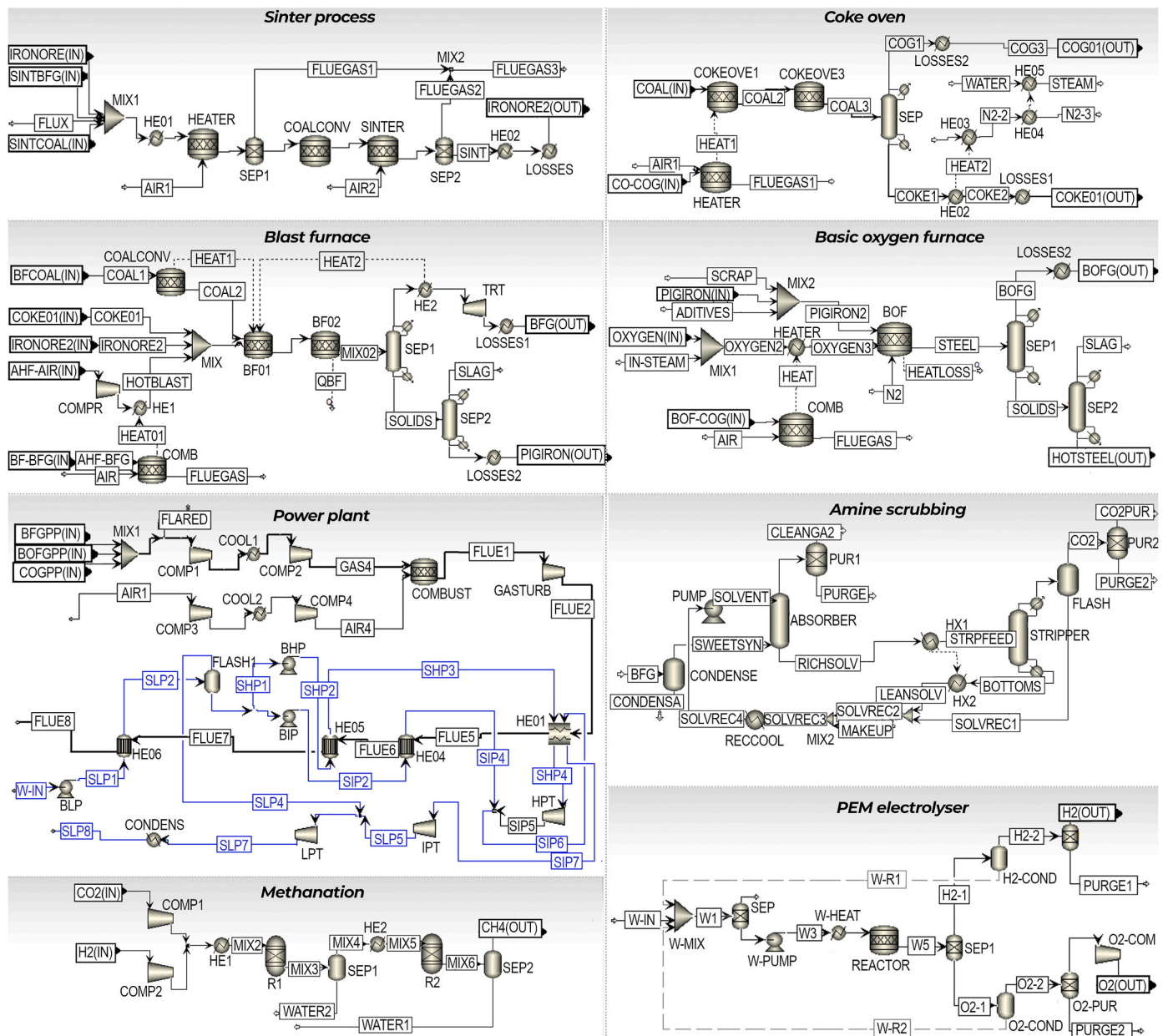


Fig. 2. Process flow diagrams of the Hierarchy blocks. Blue lines in Power Plant section represent the steam cycle. The nomenclature of each block can be found at the bottom of the figure. Further details about each block can be found in Annex A.

Table 2
Proximate and ultimate analysis of the coal [26].

Proximate Analysis	%	Ultimate Analysis	%
MOISTURE	1.20	ASH	10.89
FC	71.59	CARBON	77.73
VM	17.42	HYDROGEN	4.20
ASH	10.89	NITROGEN	1.59
		SULFUR	0.43
		OXYGEN	5.16

equilibrium, and complete combustion assumption. Table 1 details modelling input and outputs for stream compositions and temperature, for main streams of both the proposed low-carbon steelmaking process concept and conventional BF-BOF process.

The process modelling methodology for the iron and steel plant, amine scrubbing and power-to-gas sub-systems is described in Sections

3.1 to 3.3, respectively. The approaches used to evaluate the energy and economic penalties associated with carbon emission abatement are described in Sections 3.4 and 3.5, respectively.

3.1. Iron and steel making processes for both conventional and PtG-amine scrubbing based BF-BOF processes

The PFD of Fig. 1 shows the Hierarchy blocks that model the different stages of Case 0 and Case 1. Each Hierarchy block gathers the necessary sub-models of the involved processes, as shown in Fig. 2 (further details in Appendix A). The SINTER Hierarchy represents the agglomeration of fine ore particles, which takes place through an incipient fusion caused by the combustion of coal (see Fig. 2. Process flow diagrams of the Hierarchy blocks. Blue lines in Power Plant section represent the steam cycle. The nomenclature of each block can be found at the bottom of the figure. Further details about each block can be found in Annex A.

Table 2) [2]. First, a mixture of iron ore, coal (5%wt) and limestone

Table 3

Mass flows (kg/t_{HM}) of the main streams for the conventional I&S plant (Case 0) and the PtG-amine scrubbing bases I&S integration (Case 1). Stream ID corresponding to Fig. 1.

Hierarchy	Stream ID	Typical values [12,20,29,44, 45]	Case 0	Case 1
SINTER	Iron ore	1315–1510	1464	1464
SINTER	Coal	45–75	75	75
SINTER	Iron ore (out)	1315–1510	1555	1555
COKE OVEN	Coal	380–534	443	443
COKE OVEN	Coke	300–400	332	332
COKE OVEN	COG (out)	100–140	111	111
BF	Coal	150–200	150	89
BF	Air	1100–1600	1449	1520
BF	BFG	1800–2420	2216	2242
BF	Slag	260–298	261	261
BF	Pig iron	1000	1000	1000
BOF	Scrap	150	150	150
BOF	Oxygen	60–90	87	87
BOF	N2	29	9	9
BOF	BOFG	90–130	128	128
CASTING	Steel	1000–1100	1079	1079
AMINES	BFG	–	–	604
AMINES	BFG-CLEAN	–	–	461
AMINES	CO2	–	–	145
PEM	Water	–	–	245
PEM	H2	–	–	26
ASU/PEM	O2	–	87	207
METHAN	SNG	–	–	54

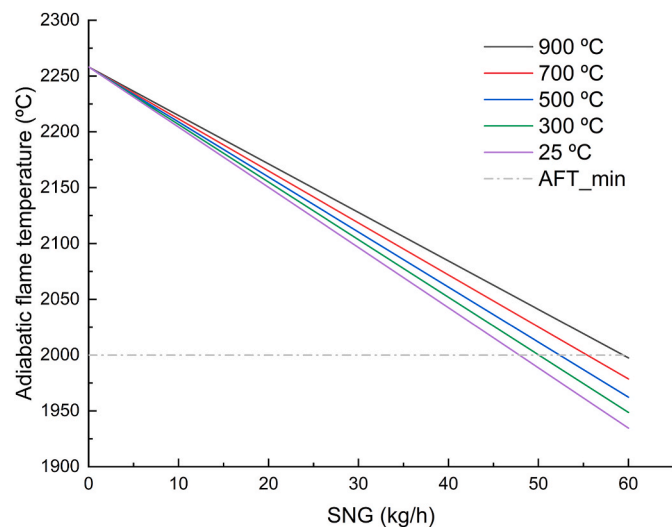


Fig. 3. Adiabatic flame temperature of the BF as a function of the SNG flow rate and temperature, when replacing coal by SNG for the proposed plant with PtG and amine scrubbing (Case 1).

is heated up to 800 °C using BFG as fuel, and then the coal of the mixture is burnt to form the sinter [20,21]. The COKEOVEN Hierarchy simulates the coke-making process from coal under anaerobic conditions. Temperature increases up to 1100 °C by burning COG (indirect heat exchange), producing coke as the target product and coke oven gas as a by-product. Then, the coke is cooled to 150 °C in the coke dry quenching (CDQ).

The BF Hierarchy comprises hot stoves and a simplified version of a blast furnace. In the hot stoves, the air is compressed up to 5 bar and indirectly heated up to 1200 °C. In the blast furnace, a first reactor reduces iron oxides to metallic iron, and a second reactor adjusts the BFG composition as a function of the CO and H₂ utilization (Eq. (1) and Eq.

(2), respectively). The pulverised coal rate in the BF is initially fixed at 150 kg/t_{HM}, the H₂ utilization at 0.47, and the carbon content in hot metal at 4.5 wt% [21,22]. The rest of the necessary parameters (BFG temperature and composition, air and coke flow rates, and flame temperature) were externally calculated through a revised Rist diagram [14, 17,27]. These data were introduced into the Aspen Plus model as inputs. When injecting SNG to the BF (Case 1), the coal rate is decreased to keep the same total coke input.

$$\mu_{CO} = CO_2 / (CO_2 + CO) \tag{1}$$

$$\mu_{H_2} = H_2O / (H_2O + H_2) \tag{2}$$

The BOF Hierarchy converts the hot metal (4.5 wt% C) to steel (0.267 wt% C) by blowing pure O₂ to react with carbon, producing CO and CO₂. This reaction is highly exothermal, so 150 kg/t_{HM} of scrap (0.267 wt% C) are added to take advantage of this energy [28,29]. The scrap melts and dissolves with the rest of the hot metal, increasing the total steel production and controlling the temperature of the process (1650 °C [19,30]). Also, some nitrogen is added from the bottom of the BOF to stir the mixture [19]. Part of the BOFG is used to preheat the oxygen inlet stream up to 1650 °C. This O₂ is produced in the ASU Hierarchy, which is based on the Hampson-Linde cycle. The ASU model was directly taken from the Aspen Plus repository (example block Power, Coal gasification, IGCC [31]), so it is not included in Fig. 2. This process separates the oxygen at –189 °C by liquefaction with 94% purity [32–34].

The PPLANT Hierarchy models a combined cycle power plant fuelled by unused BFG, COG and BOFG (100 MWe net power, [20]) [29,35]. The gas power cycle (stream in black in Fig. 2) consists of two stage air compression with intercooling before the combustion chamber and one expansion turbine for the flue gas. The turbine inlet pressure and temperature are 27 bar and 1389 °C, respectively [36]. The HRSG block is formed by four heat exchangers in counter-current mode that transfer the energy from the flue gases to the steam (stream in blue in Fig. 2), being the highest steam temperature 560 °C [36]. The steam power cycle consists of three expansion stages, corresponding to high, intermediate and low pressure turbines (blocks HPT, IPT and LPT).

3.2. Amine scrubbing processes

The amine scrubbing model was elaborated from a pre-defined Example File (Reactor and process design in Ref. [37]). It has been modified to capture 90% of the inlet CO₂, using methyldiethanolamine (MDEA) as solvent (50 wt% MDEA and 50 wt% water). Before diverting BFG to the capture plant, water is condensed. The absorption column has been modelled with 15 stages, the gas enters at the bottom (stage 15) while the lean solvent enters at the top (stage 1). The clean gas exits the absorber at atmospheric pressure and 42 °C through the top, while the rich solvent exits through the bottom at 43 °C and 1.1 bar (load of 0.2 kmol_{CO2}/kmol_{MDEA}). Then, the rich solvent is heated up to 100 °C in a heat recovery exchanger that cools the lean solvent exiting the stripper from 112 °C to 56 °C before entering again to the absorber [37]. The stripper column has 19 stages, including condenser and reboiler stages. The reflux ratio of the condenser is 0.01, and the boil-up ratio of the reboiler is 0.1 [37]. In the stripper, the rich solvent enters at the second stage, and it descends along the reactor as the CO₂ is released (the lean solvent leaves the stripper through the last stage, and the CO₂ through the first stage). A condenser separates the remaining solvent by cooling the CO₂-rich stream from 85 °C to 35 °C, achieving 95 vol% CO₂ purity. The condensed solvent is recirculated back to the absorber with the lean solvent [37]. Finally, a makeup stream, also mixed with the lean solvent before the absorber, is necessary to compensate for the water and MDEA lost in the clean gas and the CO₂ streams.

Table 4

Molar composition and lower heating value for gas streams in both the conventional process plant (Case 0) and proposed plant with PtG and amine scrubbing (Case 1).

Composition (vol%)	Case 0 – Conventional I&S					Case 1 – I&S plant integrated with PtG and amine scrubbing								
	COG	BOFG	BFG	O ₂	Sold gases	COG	BOFG	BFG	Clean BFG	O ₂	H ₂	CO ₂	SNG	Sold gases
O ₂	0.07	1.93	–	95.00	0.15	0.07	3.31	–	–	99.98	–	–	–	0.23
N ₂	2.26	8.00	55.10	0.8	39.95	2.26	8.00	54.52	64.67	–	–	0.02	0.02	44.94
CO	9.37	66.00	21.98	–	22.10	9.37	66.00	21.21	25.01	–	–	0.79	–	23.41
CO ₂	2.34	19.35	19.37	–	15.51	2.34	20.69	17.30	2.05	–	–	95.82	0.99	7.51
H ₂	57.31	2.00	2.25	–	14.70	57.31	2.00	5.67	6.72	0.02	100.0	0.00	3.86	17.51
H ₂ O	–	–	1.3	–	0.92	–	–	1.30	1.55	–	–	3.37	0.50	–
CH ₄	28.65	–	–	–	6.49	28.65	–	–	–	–	–	–	94.65	6.41
Ar	–	2.72	–	4.2	0.18	–	–	–	–	–	–	–	–	–
LHV (MJ/kg)	39.27	6.18	2.23	–	5.81	39.27	6.18	2.53	3.29	–	119.9	–	48.73	6.69

3.3. Power to gas processes

The PEM Hierarchy stands for the water electrolysis process. The input stream is water at ambient conditions (25 °C and 1 atm), and the energy consumption is set to 3.8 kWh/Nm³_{H₂} [38]. The outlet hydrogen stream drags 2 mol of H₂O per mole of H₂ (which are condensed and recirculated), and 0.1% of O₂ (which is purified). The outlet oxygen flow has only traces of hydrogen (0.01%) [38,39].

The METHAN Hierarchy contains the methanation plant where the CO₂ from the amine scrubbing reacts with the H₂ from the electrolyser to produce SNG, under stoichiometric conditions [40]. The plant uses two isothermal fixed-bed reactors working at 5 bar [41,42]. The temperature of the gas entering to the first reactor is 250 °C, while the reactor is kept isothermal at 350 °C. The outlet stream is cooled up to 100 °C to condensate the water, and then heated up to 250 °C before being diverted to the second stage. The second reactor operates at 300 °C. After this reactor, water is condensed at 25 °C to reach a SNG of 95 vol% CH₄.

3.4. Energy penalty assessment

The energy penalty associated with carbon emission reduction relative to the conventional BF-BOF process is defined as the net energy consumed per kilogram of CO₂ avoided through Power to Gas and amines carbon capture integration (Eq. (3)).

$$E_{penalty} = \frac{(E_{cons}^{Case1} - E_{prod}^{Case1}) - (E_{cons}^{Case0} - E_{prod}^{Case0}) - (\dot{m}_{coal}^{Case0} - \dot{m}_{coal}^{Case1}) * Q_{coal} * \mu_{elec} - (\dot{m}_{gases}^{Case1} * Q_{gases}^{Case1} - \dot{m}_{gases}^{Case0} * Q_{gases}^{Case0}) * \mu_{elec}}{CO_2^{Case0} - CO_2^{Case1}} \quad (3)$$

Where E_{cons} is the electricity consumed in the industry (MJ/t_{HM}), E_{prod} is the electricity produced by the owned power plant (MJ/t_{HM}), \dot{m}_{coal} is the coal consumption (kg/t_{HM}), Q_{coal} is the coal heating value (MJ/kg), \dot{m}_{gases} is the steel gases sold (kg/t_{HM}), Q_{gases} is the steel gases heating value (MJ/kg) and μ_{elec} is the energy conversion factor from coal to electricity (0.33).

3.5. Economic analysis

The cash flow statement was analysed for the investment required to integrate power to gas and amine scrubbing in the I&S plant (i.e., passing from Case 0 to Case 1), assuming 4% interest (i) for discount coefficient (D_{coeff}, defined in Eq. (4), where n is the year). The power to gas system is assumed to operate 8000 h per year [43], with electricity purchased from the grid if needed with renewable electricity assumed. The annual OPEX with regards to the reference plant are the amine renovation, the catalyst renovation, the purchased electricity, the water

consumed by the electrolyser and the operation and maintenance (O&M). The annual incomes are the saved coal, the saved CO₂ taxes, the sold oxygen and the sold fuel gases to other industries.

$$D_{coeff} = \frac{1}{(1 + i)^n} \quad (4)$$

4. Results and discussion

The comparative energy analysis of the proposed integrated PtG and amine scrubbing based steelmaking process (Case 1), and conventional BF-BOF steelmaking process (Case 0), is presented in Section 4.1 based on the simulation results. Coal and emissions savings associated with the novel PtG-amine scrubbing based process are presented and discussed in Section 4.2. The heat integration analysis performed to reduce the energy penalty of the proposed low-carbon steelmaking process concept is presented in Section 4.2. The economic viability of the proposed PtG and amine scrubbing based process concept is analysed in Section 4.3.

4.1. Simulation of conventional I&S plant (case 0) and PtG integration (case 1)

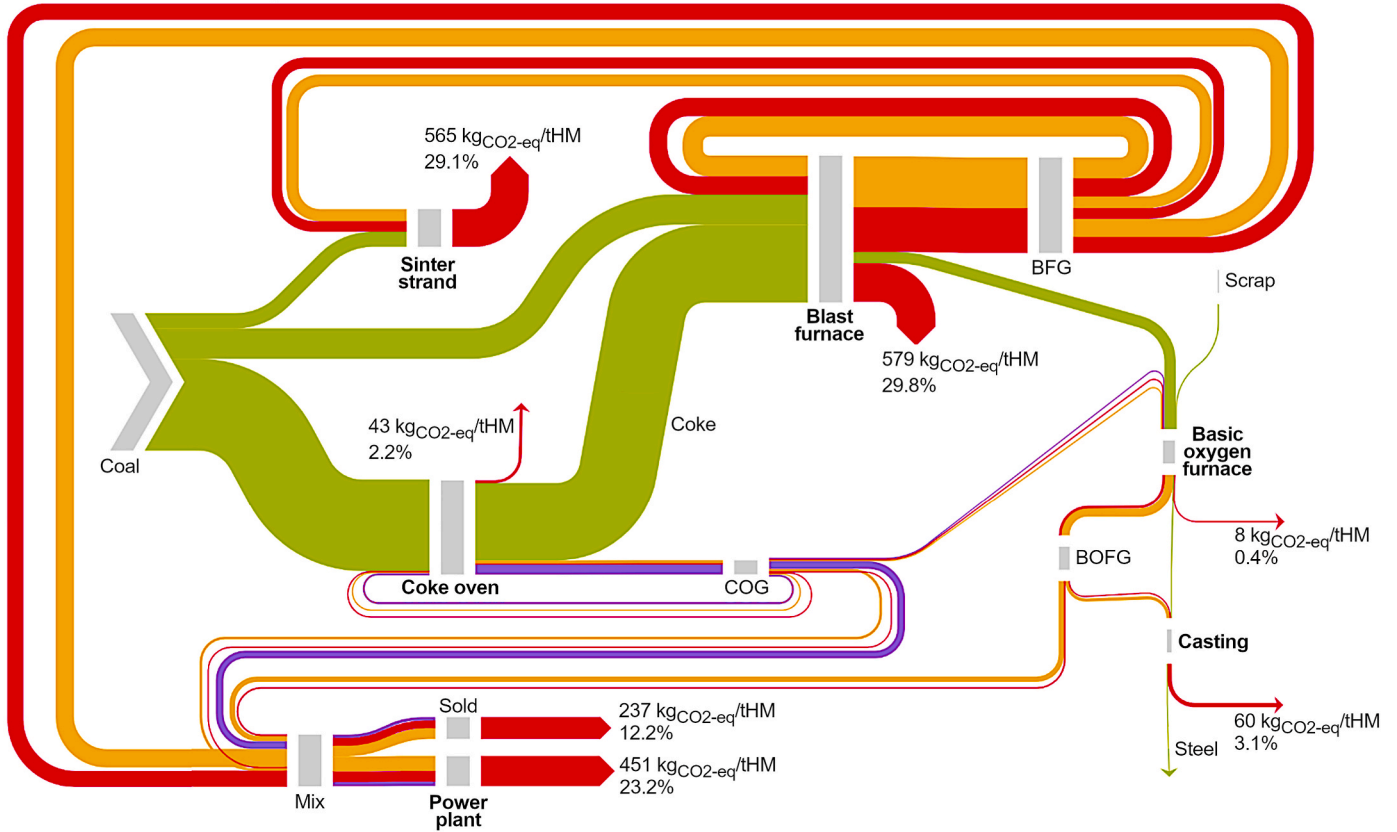
The conventional I&S plant (Case 0) was simulated considering the literature data summarized in Table 3. In the base case, 1079 kg/t_{HM} of steel are produced by consuming 1464 kg/t_{HM} of iron ore, 668 kg/t_{HM} of coal, and 150 kg/t_{HM} of scrap. After fulfilling the internal energy consumption, the remaining gases are mixed and used for electricity pro-

duction in the power plant or sold to other industries (490 kg/t_{HM} and 438 kg/t_{HM} respectively).

When integrating power to gas (Case 1), the injection of SNG allows for replacing a fraction of fossil fuel, but it leads to a decrease of the adiabatic flame temperature (AFT) at the raceways. Since the AFT should not be dropped below 2000 °C for technical reasons [46–49], the amount of SNG that can be injected is limited, as well as the size of the power to gas plant. Since replacing coke with SNG leads to a greater reduction in AFT, the substitution of coal (PCI) instead is considered to allow for greater PtG plant size. Fig. 3 shows the dependence of AFT on the SNG flow rate and on its inlet temperature when replacing coal. The preheat temperature is fixed at 700 °C (red line) according to the literature [47], so the maximum SNG flow rate to be injected is 53.6 kg_{SNG}/t_{HM}. It corresponds to a coal replacement ratio of 1.14 kg_{Coal}/kg_{SNG}, and a reduction in AFT of 4.3 °C/kg_{SNG}.

In order to obtain 53.6 kg_{SNG}/t_{HM}, a production of 26.1 kg_{H₂}/t_{HM} in the electrolyser is needed. This corresponds to a PEM electrolyzer of 335

Case 0 - Conventional I&S plant



Case 1 - I&S plant integrated with PtG and amine scrubbing

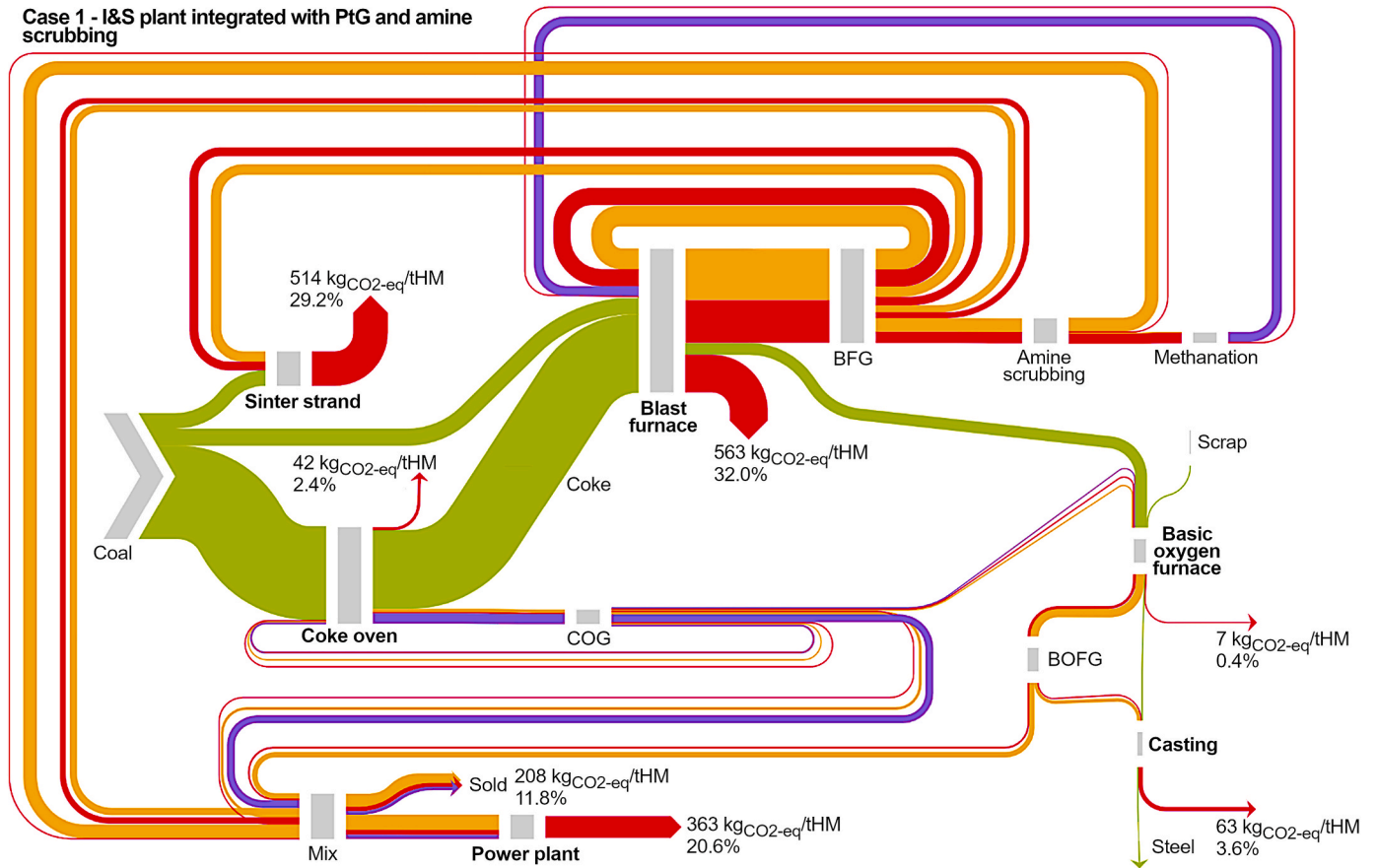


Fig. 4. Sankey diagram of the carbon mole flows (kgCO₂-eq/tHM) in both the conventional process plant (Case 0) and proposed plant with PtG and amine scrubbing (Case 1). Green: C; Orange: CO; Red: CO₂; Purple: CH₄.

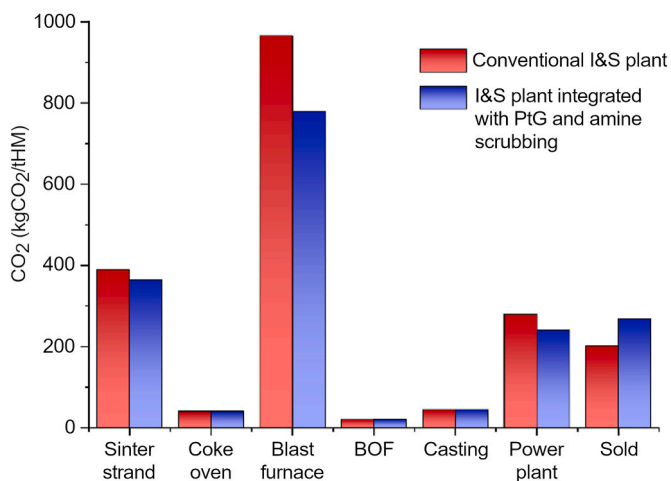


Fig. 5. CO₂ production by process (kgCO₂/t_{HM}) for the conventional I&S plant (Case 0) and its integration with PtG and amine scrubbing (Case 1).

Table 5

Process temperature and heat released and absorbed by each block of the amine and methanation system for the PtG and amine scrubbing based process (Case 1).

Block ID	Type	Load (MJ/t _{HM})	T in (°C)	T out (°C)
R1	✓ Cooling	-500.6	350	350
R2	✓ Cooling	-12.0	300	300
SEP1	✓ Cooling	-324.1	350	100
SEP2	✓ Cooling	-103.9	300	25
Stripper	✓ Heating	+542.3	110	130
HE1	✓ Heating	+28.5	193	250
HE2	✓ Heating	+28.8	100	250

MW, assuming a hot metal production of 320 t_{HM}/h. According to the simulation, the total electricity consumption of the electrolysis plant is 4021 MJ/t_{HM} (3981 MJ/t_{HM} in the stack, 39.8 MJ/t_{HM} in the compressors and 0.06 MJ/t_{HM} in the pump). The O₂ by-produced in the electrolyser amounts to 207 kgO₂/t_{HM}, which is enough to replace the ASU, and sell the excess to other industries. The water consumption accounts for 245 kgH₂O/t_{HM}, but the most part can be supplied internally by recycling water streams from the I&S plant and the PtG plant.

The carbon capture plant is also sized according to the SNG requirements (to obtain 53.6 kgSNG/t_{HM}, 145 kgCO₂/t_{HM} are needed). Since amine scrubbing captures 90% of the inlet CO₂, and the BFG has 17.3 vol % of CO₂ (Table 4), the total mass flow of BFG that has to be treated in the amine plant is 604 kgBFG/t_{HM}, in order to obtain a CO₂ stream of 145

kgCO₂/t_{HM} for the methanation plant. Results show a specific heat consumption of 3.74 MJ/kgCO₂ (i.e., 542.3 MJ/t_{HM}) for the amine scrubbing process (which is in agreement with literature, 3.1–3.64 MJ/kgCO₂ [50]). The electricity consumption in the amine plant is 29 kJ/kgCO₂ (i.e., 4.2 MJ/t_{HM}) [50].

In the methanation plant, there is only electricity consumption, which amounts to 2.09 MJ/kgSNG (i.e., 112 MJ/t_{HM}). The heat released by the exothermic reaction was 16.3 MJ/kgSNG (i.e. 874 MJ/t_{HM}), which is enough to satisfy the methanation preheaters as well as the thermal demand of the amine scrubbing process.

4.2. CO₂ emissions and coal saving

The total operational CO₂-equivalent emissions of the conventional I&S plant are 1943 kgCO₂-eq/t_{HM} which is in agreement with the literature, 1800–2500 kgCO₂-eq/t_{HM} [30,51–53]. These emissions can be diminished to 1760 kgCO₂-eq/t_{HM} by integrating power to gas, representing a 9.4% reduction in CO₂-equivalent emissions. The cut in CO₂-equivalent emissions is directly related to the saving in coal consumption. The amount of pulverised coal substituted by SNG in the BF is 65 kg/t_{HM} (56.7% replacement), which represents 9.6% of the total coal consumed in the whole I&S plant accounting sinter strand, coke oven and BF (680 kg/t_{HM} in the base case).

The Sankey diagrams (Fig. 4) show how these equivalent emissions are distributed through the BF-BOF plant by type of component (green: C, orange: CO, red: CO₂, and purple: CH₄). The equivalent CO₂ released in the sinter strand is reduced by 9% when integrating PtG because the BFG burnt in the preheating of the sinter has a higher energy content and lower CO₂ volume fraction. Therefore, less BFG mass flow rate is needed

Table 6

Heat exchanger network of the pinch analysis for the PtG and amine scrubbing based process (Case 1).

HEX	Block ID	Load (MJ/t _{HM})	Area (cm ² /t _{HM} /h)	Area (m ² ^a)	ΔT min Hot (°C)	ΔT min Cold (°C)
HE-A	R1 to HE1	28.5	20319	650	99.5	153.8
HE-B	R1 to HE2	28.8	16970	543	100.0	249.5
HE-C	SEP1 to STRIPPER	84.2	6104	195	220.0	175.1
HE-D	R1 to STRIPPER	443.3	23778	761	220.0	239.5
HE-E	SEP1 to WATER	239.9	22315	714	265.1	85.0

^a Assuming a BF of 320 t_{HM}/h.

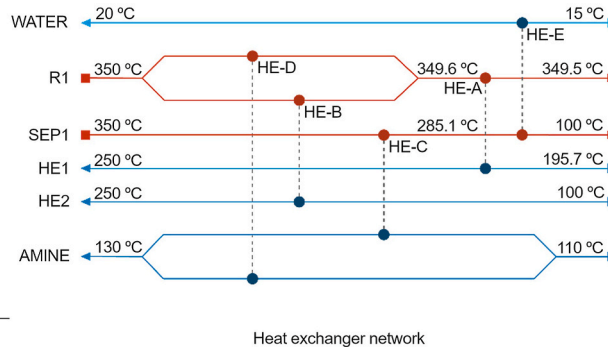
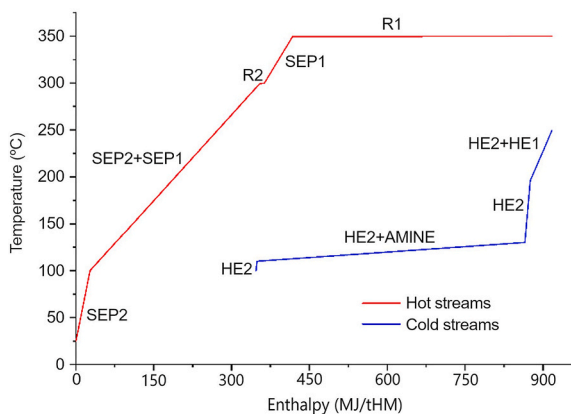
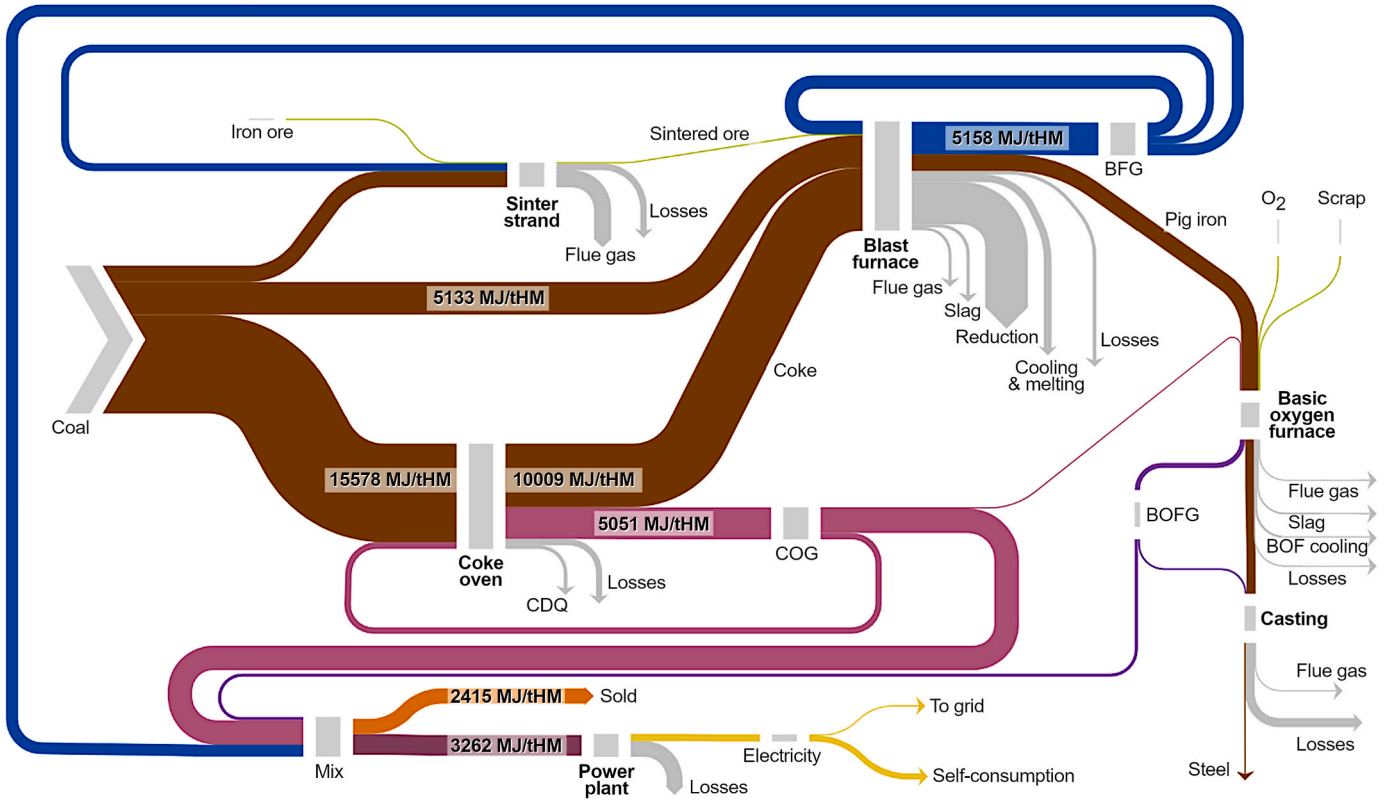


Fig. 6. Heat integration through pinch analysis of the amine plant and the methanation plant for the PtG and amine scrubbing based process (Case 1). Left: Composite curve; Right: Heat exchanger network.

Case 0 - Conventional I&S plant



Case 1 - I&S plant integrated with PtG and amine scrubbing

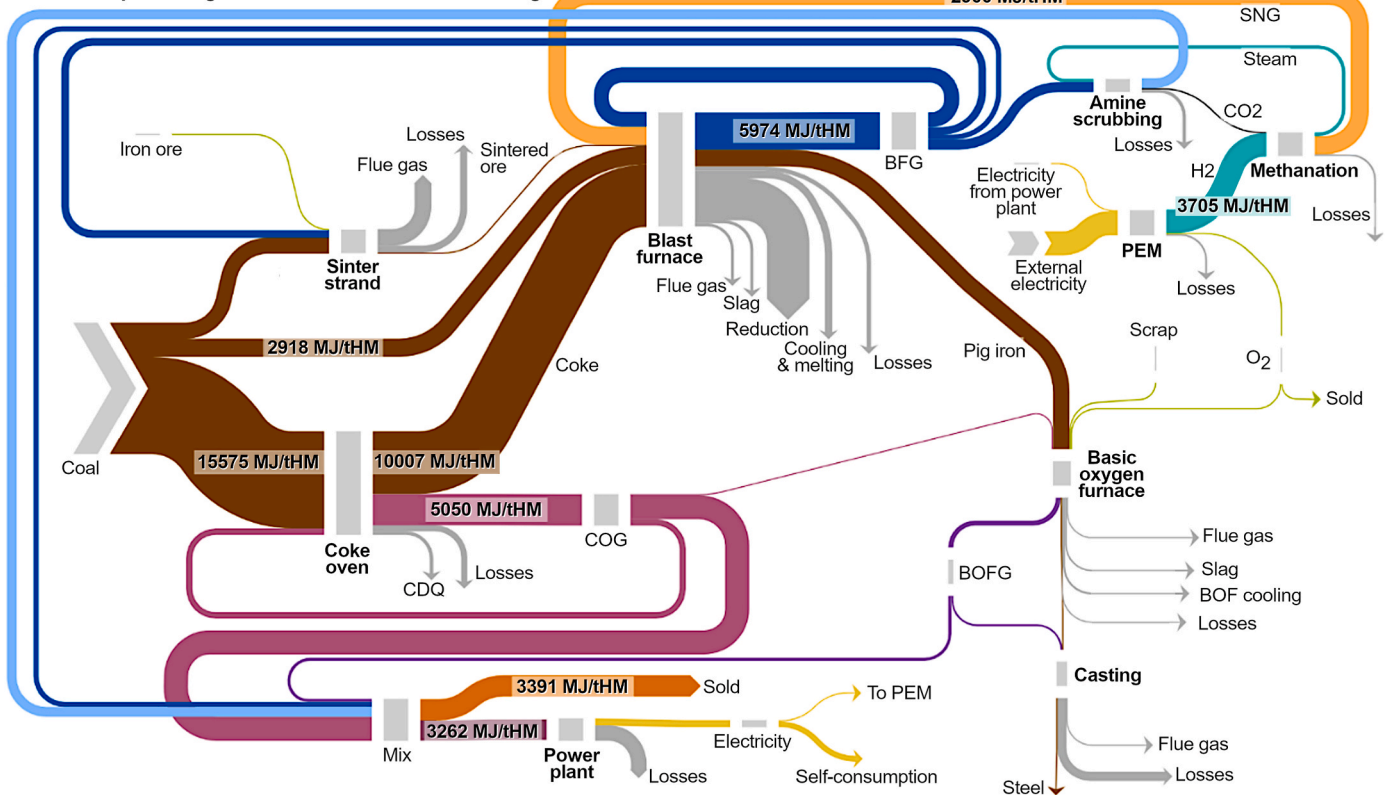


Fig. 7. Sankey diagram of the energy flows in both the conventional process plant (Case 0) and proposed plant with PtG and amine scrubbing (Case 1).

Table 7

Main electrical consumptions and productions (MJ/t_{HM}) calculated for the conventional I&S plant (Case 0) and its integration with PtG and amine scrubbing (Case 1).

	Case 0	Case 1
Electricity consumption		
Sintering [29]	180	180
Coke oven [29]	42	42
Blast Furnace [29]	376	376
ASU	119	–
PEM	–	4021
Amines	–	5
Methanation	–	108
BOF [29]	128	128
Other [29]	105	105
Total	950	4965
Electricity production		
Power plant	1122	1122
Power plant efficiency	34.4%	34.4%

to provide the same energy. This effect also occurs in the BF block, whose direct emissions are related to the BFG consumed to preheat the hot blast. However, in this case the reduction in CO₂-equivalent emissions is lower (3%) because the effect is counterbalanced by the higher amount of hot blast required and by the necessity of preheating the SNG to 700 °C (i.e., more BFG is consumed). The major reduction in CO₂-equivalent emissions takes place in the power plant (19.5% reduction), since most of the CO₂ that was being diverted (because of the BFG) is now recycled into methane. This can be seen in the red line which is now captured by the amine scrubbing and converted to methane (purple line). Other sub-processes such as the coke oven, the BOF, and the casting step, are barely affected by the PtG integration.

Despite the CO₂-equivalent emissions are similar in the hot stoves, the sintering process and the downstream gases (power plant and sold gases), most of the CO₂ released was originally generated in the blast furnace. Fig. 5 depicts the CO₂ actually generated in each process of the I&S plant (not the CO₂-equivalent emissions), including the conversion shown in the Sankey diagram for C (green colour), CO (yellow colour) or CH₄ (purple colour). The CO₂ pre-existing in the input stream is accounted into the stage where is generated (e.g., in the sintering process it is not accounted for the CO₂ coming from the BFG, since it was actually generated and considered in the BF). Under this framework, it can be seen in Figs. 4 and 5 that the BF actually produces 966 kgCO₂/t_{HM} in Case 0, but these emissions are later diverted in the form of BFG and emitted in other processes (sintering and power plant mainly). When integrating power to gas, the production of CO₂ in the BF is decreased by

19% (to 780 kgCO₂/t_{HM}), but it is still the major producer of CO₂ in the I&S plant. The reduction is due to the recycling of CO₂ (carbon closed loop) and the utilization of fuel with greater hydrogen content (CH₄ instead of coal). It can be seen that the CO₂ generated in the sinter strand and in the power plant barely changes, since both processes keep producing the same amount of energy (heat or electricity), which means the same conversion of CO into CO₂. The little decrease is due to the greater H₂ content and lower CO₂ content in the BFG which in turn reduces the CO consumption as less inert gas has to be heated. The processes in coke oven, BOF and casting produce small amounts of CO₂, since their outlet gases are mainly composed of CO and CH₄.

4.3. Heat integration and energy penalty

When integrating Power to Gas and amine scrubbing (Case 1), new thermal streams arise in the process flow diagram. The methanation process provides heat at 300–350 °C, while the CO₂ desorption requires heat at 110–130 °C. Moreover, additional preheating and condensation stages come into operation. Therefore, the proposed concept was thermally integrated using Pinch analysis to reduce the energy penalty. The overall cooling and heating needs of the PtG plant are 940.6 MJ/t_{HM} and 599.6 MJ/t_{HM}, respectively (Table 5). The main heating requirement (solvent regeneration in Stripper) can be satisfied by the cooling needs of methanation process. The analysis with Aspen Energy Analyser tool provided 11 different possible configurations, being the feasible number of heat exchangers between 5 and 9. Fig. 6 shows the composite curve of the Pinch analysis and the selected network of five heat exchangers. Assuming a BF producing 320 t_{HM}/h, the major heat exchanger would require an effective area of 8.8 m² (Table 6).

The selected configuration does not recover heat from reactor 2 (block R2) because of the low amount of energy available, nor from the separator 2 (SEP2) because of the low temperature level. Producing steam with the remaining heat is also not worth it because of the low energy flow recoverable compared with the energy available from other processes within the steel plant (investment in additional pipes for this steam distribution would not be reasonable). Another alternative to heat integration would be to increase the size of the CO₂ capture plant to match the heat available in methanation. However, this option is discarded because it would imply the need for geological storage for the extra CO₂, which is not considered in Case 1.

Thanks to the heat integration, we avoid using part of the BFG in the heat requirements of the amine scrubbing and in the preheating of the methanation plant. Thus the total energy that can be sold in the form of fuel gases is 3391 MJ/t_{HM} for Case 1 (I&S integrated with PtG), which is

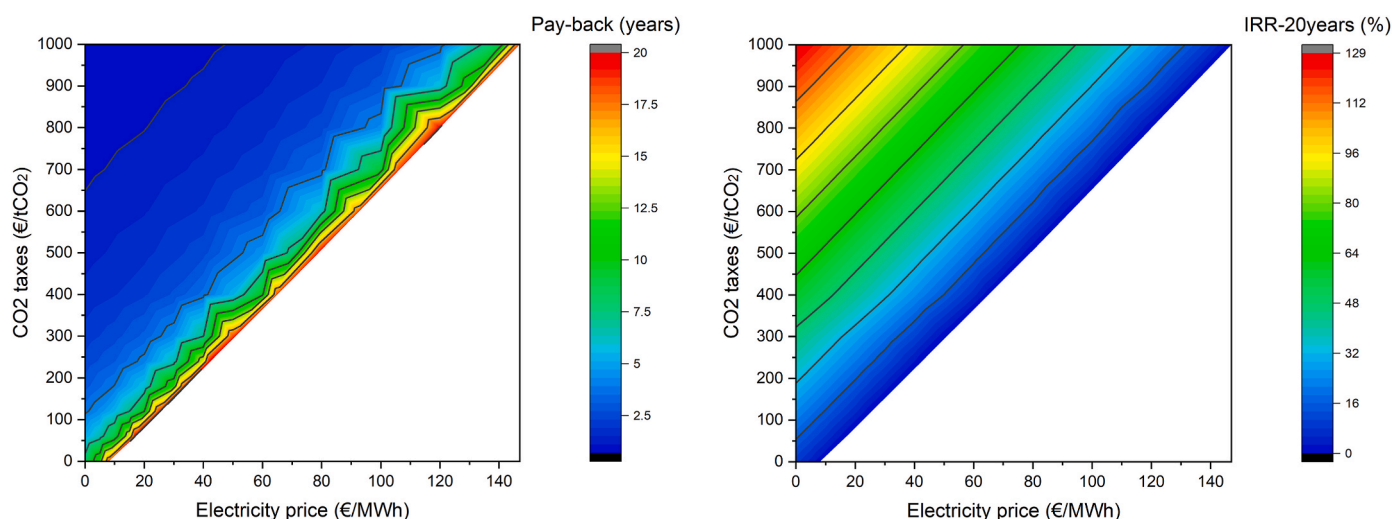


Fig. 8. Pay-back and Internal rate of return (IRR) as a function of electricity price and CO₂ taxes for the proposed plant with PtG and amine scrubbing (Case 1).

Table 8
Economic analysis for PtG and amine scrubbing integration in the iron and steel industry (Case 1).

	Cost (M€) or (M€/y)	Cost equation (M€) or (M€/y)	Parameters α, β, γ	Ref.
CAPEX				
Amine Plant				
Overall Cost	7.40	$26.094 \cdot (\alpha/408)^{0.65}$	CO ₂ captured [t/h]	[57]
Electrolysis				
Overall Cost	142	$400 \cdot 10^{-6} \cdot \alpha$	Power [kW]	[58]
Methanation				
H2 compressor	0.036	$0.267 \cdot (\alpha/445)^{0.67}$	Power [kW]	[59]
CO2 compressor	0.017	$0.267 \cdot (\alpha/445)^{0.67}$	Power [kW]	[59]
Reactors	0.245	$300 \cdot 10^{-6} \cdot \alpha$	SNG power [kWSNG]	[60]
Catalyst	4.391	$0.1875 \cdot \alpha$	Volume of catalyst [m ³]	[59]
Heat exchangers				
Aspen Energy Analyzer				
HE-A	0.137	$(9016.3 + 721.3 \cdot (\alpha^{0.8})) \cdot 10^{-6}$	Area [m ²]	–
HE-B	0.120	$(9016.3 + 721.3 \cdot (\alpha^{0.8})) \cdot 10^{-6}$	Area [m ²]	–
HE-C	0.058	$(9016.3 + 721.3 \cdot (\alpha^{0.8})) \cdot 10^{-6}$	Area [m ²]	–
HE-D	0.155	$(9016.3 + 721.3 \cdot (\alpha^{0.8})) \cdot 10^{-6}$	Area [m ²]	–
HE-E	0.147	$(9016.3 + 721.3 \cdot (\alpha^{0.8})) \cdot 10^{-6}$	Area [m ²]	–
Other direct costs				
Installation	15.47	$10\% \cdot \alpha$	Total equipment costs [€]	[57]
Instrumentation & control	6.19	$4\% \cdot \alpha$	Total equipment costs [€]	[57]
Piping	24.75	$16\% \cdot \alpha$	Total equipment costs [€]	[57]
Electrical	7.74	$5\% \cdot \alpha$	Total equipment costs [€]	[57]
Building	7.74	$5\% \cdot \alpha$	Total equipment costs [€]	[57]
Land	1.55	$1\% \cdot \alpha$	Total equipment costs [€]	[57]
Indirect costs				
Engineering	15.27	$7\% \cdot \alpha$	Total direct costs [€]	[57]
Legal expenses	5.37	$2\% \cdot \alpha$	Total CAPEX	[57]
Construction expenses	10.73	$4\% \cdot \alpha$	Total CAPEX	[57]
Contingency	18.78	$7\% \cdot \alpha$	Total CAPEX	[57]
TOTAL CAPEX	268.3			
OPEX				
Amine renovation	0.16	$3204 \cdot 10^{-6} \cdot \alpha$	MDEA renovation [tCO ₂ /y]	[61]
Catalyst renovation	0.66	$15\% \cdot \alpha$	Initial catalyst cost [M€]	[43]
Electricity	210.97	$342.5 \cdot 10^{-6} \cdot \alpha \cdot \beta$	Electricity cost [€/MWh], Operating hours [h/y]	–
Water	0.87	$1.47 \cdot 10^{-6} \cdot \alpha$	Water consumption [m ³ /y]	[62]
O&M	8.05	$3\% \cdot \alpha$	Total CAPEX	[57]
TOTAL OPEX	220.71			
INCOMES				
Coal	20.42	$130.43 \cdot 10^{-6} \cdot \alpha$	Coal generated [tCO ₂ /y]	[63]
Oxygen	25.13	$80 \cdot 10^{-6} \cdot \alpha$	O ₂ generated [tO ₂ /y]	[55]
CO2 taxes	39.59	$84.28 \cdot 10^{-6} \cdot \alpha$	CO ₂ consumed [tCO ₂ /y]	[54]
Steel gases	16.83	$(15 + \alpha \cdot \beta / 13) \cdot 10^{-6} \cdot \gamma$	Natural gas price [€/MWh] Steel gases LHV [kWh/kg] Steel gases generated [MWh/y]	–
TOTAL INCOMES	101.97			
ANNUAL BENEFIT	–127.73			
PAY-BACK	>20 years			
CARBON AVOIDANCE COST	52 €/t _{HM} or 283 €/tCO ₂			

greater than the energy sold in the base case scenario, 2415 MJ/t_{HM}. The increase in the energy sold is due to the higher energy content of the BFG when injecting SNG into the blast furnace (higher H₂ content and lower CO₂ content). This leads to more efficient processes when BFG is used (sintering, BF and power plant), namely, less BFG is consumed for the same energy production. Approximately, in Case 0, 10% of the total energy input is just sold to other industries nearby, while in Case 1 it is sold the 14% (Fig. 7).

The opposite situation it is found in the case of electricity. In the conventional I&S plant (Case 0), the total electricity consumption is 950 MJ/t_{HM}, and the electricity produced by the power plant is 1122 MJ/t_{HM} (34.4% efficiency), hence being a self-sufficient industry (Table 7). When integrating PtG, the electrical consumption remarkably increases to 4965 MJ/t_{HM}, mainly due to the electrolyser (4021 MJ/t_{HM}). The electrical production in the power plant is assumed to remain constant, as greater productions (by consuming the sold gases) would require investing in a new power plant or retrofitting the existing one. Therefore, the industry is no longer self-sufficient, and 3843 MJ/t_{HM} of green electricity from a renewable source are needed in Case 1. The net energy penalty of the PtG integration is 16.2 MJ_e/kgCO₂.

4.4. Economic analysis

The economic analysis of the PtG integration in the iron and steel industry together with the cost equations are presented in Fig. 8, using economic parameters for year 2022. An operation of 8000 equivalent hours is considered [43], with an electricity cost of 77 €/MWh, a CO₂ allowances price of 84 €/tCO₂ [54], a selling price for O₂ of 80 €/tO₂ [55] and a variable selling price for the steel gases depending on its heating value and the natural gas price (see Table 8). The amortization of the loan is considered to happen in 20 years with an annual interest of 4%. Moreover, the volume of the reactors is calculated assuming a GHSV of 5000 h⁻¹ [56], and the catalyst is supposed to occupy 60% of that volume. Under the current scenario, the annual benefit of the PtG integration is negative (128 M€/y annual loss), which implies that the loan cannot be amortized. Therefore, the carbon abatement cost is 52 €/t_{HM} or 283 €/tCO₂.

An economic parametric study was performed looking for those combinations of electricity prices and CO₂ allowances that allow the Power to Gas to be economically viable for the proposed solution. The calculated pay-back and internal rate of return (IRR) are presented in Fig. 8. It can be seen that even at electricity prices of 40 €/MWh, a minimum of 240 €/tCO₂ is required to amortize the investment (14.6

years of pay-back and 2.6% of IRR). This economic analysis suggests that profitable scenarios would require that either the electricity is obtained at the cost of production (i.e., the renewable park belongs to the I&S industry) or subsidies are given to the purchased electricity or power purchase agreements (PPAs) are signed in order to fight against global warming. For example, assuming an electricity cost of 51 €/MWh (i.e., cost production for solar panels in 2022 [64]), the PtG integration would be profitable when the CO₂ taxes are above 294 €/t_{CO2}. If the cost of the electricity is 35 €/MWh (i.e., cost of production for wind power in 2022 [64]), profits are obtained for CO₂ taxes greater than 177 €/t_{CO2}. Furthermore, for the current scenario of CO₂ allowances prices (84 €/t_{CO2}) and grid's electricity prices (77 €/MWh), the minimum subsidy to make PtG integration profitable would be 54.6 €/MWh. If the CO₂ taxes increases to 128 €/t_{CO2} in the coming years (estimated price by 2050 by Strefler et al. [65]), the necessary subsidy would be reduced to 49 €/MWh. By 2100, Strefler estimates CO₂ taxes between 384 €/t_{CO2} and 907 €/t_{CO2}, needing a 13.7 €/MWh subsidy and no subsidy respectively. In the last evaluation (907 €/t_{CO2}) the pay-back would be 2 years and the IRR 52%.

5. Conclusions

A novel concept of integrating power-to-gas (PtG) technology in the ironmaking process is presented. Synthetic natural gas (SNG) from methanation is injected into the blast furnace to reduce the pulverised coal consumption, thus recycling CO₂ in a closed loop. The PtG plant is sized to inject as much SNG as possible, as long as the AFT does not fall below 2000 °C, accounting for 53.6 kg_{SNG}/t_{HM}. The electrolyser and the amine scrubbing plants are sized according to the SNG flow rate. This concept is compared with a conventional blast furnace – basic oxygen furnace (BF-BOF) steelmaking plant (Case 0).

The base case model includes the sintering, coke oven, blast furnace, ASU, basic oxygen furnace, casting and power plant. For PtG integration, an electrolyser, amine scrubbing and methanation plant are added to the reference model, and the blast furnace is run under SNG injection. The SNG injection replaces pulverised coal injection (PCI) with a ratio of 1.14 kg_{Coal}/kg_{SNG}. Mass flows, compositions and thermal and electricity consumptions are calculated through Aspen Plus® software simulations.

Optimization of the heat integration between the amine plant (endothermic) and the PtG (exothermic) was carried out through the pinch analysis method. In this way, no extra heat is needed for the amine plant, which means no energy penalisation is added for capturing the CO₂.

Savings achieved in CO₂ emissions with the PtG integration are 9.4% (equivalent to 4.7·10⁵ t_{CO2}/year), with a reduction in coal fuel of 9.2% (61 kg_{Coal}/t_{HM}, representing a 41% replacement of the PCI). The electric energy required to avoid these emissions is 16.2 MJ/kg_{CO2} of renewable power for electrolysis. This significant energy penalty is mainly due to a 355 MW electrolyser.

Under the current economic situation, the concept is not economically feasible, reaching a carbon abatement cost of 52 €/t_{HM} or 283 €/t_{CO2}. However, sensitivity analyses have shown that the concept would become economically feasible under certain conditions,

Appendix A. Aspen Plus modelling blocks considered in hierarchies

The main modelling blocks are listed in Table 9, showing the hierarchy, the block ID, the type of block (for Aspen Plus modelling), the description and the reactions.

depending on the CO₂ taxes, the electricity price or the subsidy amount.

A limitation of the economic analysis presented is the uncertainty in future electricity and carbon prices, as these parameters have a significant impact on the payback period and IRR of the proposed process concept. With regard to potential technical limitations, each steel-making process plant has unique equipment/space layouts that can impact the practical implementation of energy integration, including heat integration.

In general, this novel concept has the advantage of reducing fuel consumption and eliminating geological storage of the recirculated CO₂, which are additional economic and practical benefits compared to conventional carbon capture and storage. This PtG configuration also allows the industry to be indirectly electrified since fossil fuel is replaced by synthetic natural gas, which is derived from hydrogen produced by renewable sources. Other concepts of PtG integration in the blast furnace ironmaking will be considered in future work to reduce energy penalties and costs.

Credit author statement

Conceptualization, J.P., M.B., and B.P.; methodology, J.P., M.B. and B.P.; model, J.P. and M.B.; formal analysis, J.P.; writing—original draft preparation, J.P. and M.B.; writing—review and editing, M.B., B. P., L.M.R., and V.E.; visualization, J.P. and M.B.; supervision, M.B. and B.P.; project administration, M.B. and L.M.R.; funding acquisition, M. B. L.M.R. and V.E. All authors have read and agreed to the published version of the manuscript.

Declaration of competing interest

The authors declare the following financial interests/personal relationships which may be considered as potential competing interests: Luis Miguel Romeo reports financial support was provided by University of Zaragoza. Luis Miguel Romeo and Valerie Evely report financial support was provided by Khalifa University. Manuel Bailera reports financial support was provided by EU Framework Programme for Research and Innovation Marie Skłodowska-Curie Actions.

Data availability

Data will be made available on request.

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Table 9
List of main Aspen Plus modelling blocks.

Hierarchy	Block ID	Type	Description	Reactions
Sinter	HE01	Heater	Preheats the mix and the BF off-gas (BFG)	n/a
Sinter	HEATER	RStoic	Heats the raw mix by burning the BFG with air (stream AIR1) to 800 °C	$H_2 + 0.5O_2 \rightarrow H_2O$ $CO + 0.5O_2 \rightarrow CO_2$
Sinter	SEP1/SEP2	Sep	Separates the flue gases from the solid phase	n/a
Sinter	COAL-CONV	RStoic	Converts the nonconventional coal to the individual components	$Coal \rightarrow C + H_2 + O_2 + N_2 +$ Ash
Sinter	SINTER	RStoic	Burns the coal creating the sinter	$C + O_2 \rightarrow CO_2$ $H_2 + 0.5O_2 \rightarrow H_2O$
Sinter	HE02	Heater	Recovers heat from the sintered ore to preheat the inlet (HE01)	n/a
Coke oven	HEATER	RStoic	Indirectly heats the coal by burning COG, until the COAL3 stream reaches 1100 °C	$H_2 + 0.5O_2 \rightarrow H_2O$ $CO + 0.5O_2 \rightarrow CO_2$ $CH_4 + 2O_2 \rightarrow CO_2 + 2H_2O$
Coke oven	COKEOVE1	RStoic	Converts the nonconventional coal to the individual components	$Coal \rightarrow C + H_2 + O_2 + N_2 +$ Ash
Coke oven	COKEOVE3	RStoic	Converts the initial components obtained from coal decomposition in COKEOVE1 to the desired COG components	$C + 0.5O_2 \rightarrow CO$ $C + O_2 \rightarrow CO_2$ $C + 2H_2 \rightarrow CH_4$
Coke oven	SEP	Sep	Separates the gases from the solid phase	n/a
BF	COALCONV	RStoic	Converts the nonconventional coal to the individual components	$Coal \rightarrow C + H_2 + O_2 + N_2 +$ Ash
BF	COMPR	Turbine	Compresses the inlet air to 5 bar	n/a
BF	HE1	Heater	Heats the hot blast up to 1200 °C	n/a
BF	COMB	RStoic	Indirectly heats the hot blast by burning BFG	$H_2 + 0.5O_2 \rightarrow H_2O$ $CO + 0.5O_2 \rightarrow CO_2$
BF	BF01	RStoic	Reduces the iron ore to pure iron	$Fe_2O_3 + 3CO \rightarrow 2Fe + 3CO_2$ $Fe_2O_3 + 3C \rightarrow 2Fe + 3CO$
BF	BF02	RStoic	Reacts the coal and gases to form the BFG	$C + 0.5O_2 \rightarrow CO$ $CO + 0.5O_2 \rightarrow CO_2$ $CH_4 + 0.5O_2 \rightarrow CO + 2H_2$ $H_2 + 0.5O_2 \rightarrow H_2O$
BF	SEP1	Sep	Separates the gases from the solid phase	n/a
BF	SEP2	Sep	Separates the slag from the pig iron	n/a
BF	TRT	Turbine	Top Recovery Turbine: Recovers energy from the BFG	n/a
BF	HE2	Heater	Heat exchange between the BFG and the solids on the top	n/a
BF	LOSSES1	Heater	Thermal losses to the ambient temperature	n/a
BF	LOSSES2	Heater	Thermal losses of the pig iron out of the BF	n/a
BOF	COMB	RStoic	Indirectly heats the O ₂ stream by burning COG	$H_2 + 0.5O_2 \rightarrow H_2O$ $CO + 0.5O_2 \rightarrow CO_2$ $CH_4 + 2O_2 \rightarrow CO_2 + 2H_2O$
BOF	HEATER	Heater	Heats the O ₂ stream to 1650 °C	n/a
BOF	BOF	RStoic	Reduces the C content and forms the BOFG	$C + 0.5O_2 \rightarrow CO$ $CO + 0.5O_2 \rightarrow CO_2$ $H_2O \rightarrow H_2 + 0.5O_2$
BOF	SEP1	Sep	Separates the gases from the solid phase	n/a
BOF	SEP2	Sep	Separates the slag from the hot steel	n/a
BOF	LOSSES2	Heater	Thermal losses to the ambient temperature	n/a
Power plant	SPLIT1	FSplit	Separates excess gases	n/a
Power plant	COMP1-4	Compr	Compresses the air and the energetic gases	n/a
Power plant	COOL1-2	HeatX	Cools down the air and the gas streams	n/a
Power plant	COMBUST	RStoic	Burns the mixed gases	$H_2 + 0.5O_2 \rightarrow H_2O$ $CO + 0.5O_2 \rightarrow CO_2$ $CH_4 + 2O_2 \rightarrow CO_2 + 2H_2O$
Power plant	GASTURB	Compr	Turbines the flue gases producing electricity	n/a
Power plant	HE01	MHeatX	Heats two steam flows	n/a
Power plant	HE04-06	HeatX	Vaporizes water streams	n/a
Power plant	FLASH1	Flash2	Separates the steam from the water	n/a
Power plant	BLP/BIP/BHP	Pump	Pumps water to low, intermediate and high pressure	n/a
Power plant	HPT/IPT/LPT	Compr	Turbines the steam to high, intermediate and low pressure	n/a
Power plant	CONDENS	Heater	Cools the steam condensing the water	n/a
Amine plant	CONDENSE	Flash2	Condenses the water in the inlet stream	n/a
Amine plant	ABSORBER	RadFrac	Reacts the MDEA with the CO ₂ , forming the rich solvent	n/a
Amine plant	HX1	Heater	Heats the rich solvent by cooling the lean solvent	n/a
Amine plant	STRIPPER	RadFrac	Heats the rich solvent releasing the CO ₂	n/a
Amine plant	FLASH	Flash2	Condenses the water and MDEA in the CO ₂ stream at 25 °C	n/a
Amine plant	HX2	Heater	Cools the lean solvent by heating the rich solvent	n/a
Amine plant	RECCOOL	Heater	Cools the lean solvent to 40 °C	n/a
Amine plant	PUMP	Pump	Pumps the solvent to 1.5 bar	n/a
PEM	W-PUMP	Pump	Pumps the inlet water to 1.5 bar	n/a
PEM	W-HEAT	Heater	Heats the water to 80 °C	n/a
PEM	REACTOR	RStoic	Splits the water into H ₂ and O ₂	$H_2O \rightarrow H_2 + 0.5O_2$
PEM	SEP1	Sep2	Separates the H ₂ and O ₂ in two different streams	n/a
PEM	H2-COND	Flash2	Condensates the water dragged by the H ₂	n/a

(continued on next page)

Table 9 (continued)

Hierarchy	Block ID	Type	Description	Reactions
PEM	O2-COND	Flash2	Condensates the water dragged by the O ₂	n/a
PEM	O2-COM	Compr	Compresses the O ₂ to 5 bar	n/a
Methanation	COMP1	Compr	Compresses the CO ₂ to 5 bar	n/a
Methanation	COMP2	Compr	Compresses the H ₂ to 5 bar	n/a
Methanation	HE1	Heater	Heats the gases to 250 °C	n/a
Methanation	R1	RGibbs	Reacts the inlet gases to form SNG at 350 °C	4H ₂ + CO ₂ → CH ₄ + 2H ₂ O
Methanation	SEP1	Flash2	Condensates the water formed in R1 at 100 °C	n/a
Methanation	HE2	Heater	Heats the gases to 250 °C	n/a
Methanation	R2	RGibbs	Reacts the inlet gases to form SNG at 300 °C	4H ₂ + CO ₂ → CH ₄ + 2H ₂ O
Methanation	SEP2	Flash2	Condensates the water formed in R2 at 25 °C	n/a

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