

Green steel: design and cost analysis of hydrogen-based direct iron reduction

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Abstract

Hydrogen-based direct reduced iron (H₂-DRI) is an alternative pathway for low-carbon steel production. Yet, the lack of established process and business models defining “green steel” make it difficult to understand what the respective H₂ price has to be in order to be competitive with commercial state-of-the-art natural gas DRI. Given the importance of establishing break-even H₂ prices and CO₂ emission reduction potentials of H₂-DRI, this study conducted techno-economic analyses of several design and operation scenarios for DRI systems. Results show that renewable H₂ use in integrated DRI steel mills for both heating and the reduction of iron ore can reduce direct CO₂ emissions by as much as 85%, but would require an H₂ procurement cost of \$1.63 per kg H₂ or less. When using H₂ only for iron ore reduction, economic viability is reached at an H₂ procurement cost of \$1.70 per kg, while achieving a CO₂ emission reduction of 76% at the plant site. System design optimization strategies around excess H₂ ratios in the DRI top gas and the H₂ recycle pressurization can further improve performance and economics. Low H₂ excess ratios are particularly attractive as they reduce pre-heating energy requirements and offer integration opportunities with static recycle ejectors if H₂ is supplied at sufficiently high pressure. The potential of utilizing the EAF off-gas is shown to be much more synergistic with H₂ DRI than NG DRI and can increase the break-even H₂ procurement cost by up to 7¢ per kg H₂. Such findings are critical for setting technical performance criteria for H₂ supply and storage in the iron and steel sector.

Introduction

Iron and steelmaking are essential parts of the U.S. economy, generating 87 million metric tonnes (MMT) of steel and \$88 billion in total revenue as of 2023¹. These industries rely on fossil resources as heat sources and as reducing agents for the reduction of iron oxide in ore. Global iron and steelmaking are responsible for 8% of the global final energy demand and represent 7% of the energy sector’s carbon dioxide (CO₂) emissions.² The U.S. steel industry used 431 trillion BTU of energy for nonfuel purposes and 1,100 trillion BTU of energy as fuel in 2006.³ In the U.S., energy consumption per 2000 dollar value of shipments declined by 30% between the years 1991 and 2002, suggesting that energy intensity of the industry dropped.³ Energy efficiency improvements have been achieved by increasing the recycling of steel, which consumes less energy per unit product, and by lowering the energy consumption of iron and steelmaking^{4,5}. These efficiency improvements tend to be motivated by high and uncertain energy prices but additionally contribute to reducing the direct CO₂ intensity of the industry, where the average carbon intensity of crude steel production from iron ore is 1,990 kg of CO₂ per metric tonne of steel compared with scrap steel recycling at 791 kg of CO₂ per metric tonne of steel.^{4,6} While steel recycling has proven to be energy efficient and low in CO₂ emissions, primary steel will still be necessary to meet the growing demand for steel.⁵

Further minimizing CO₂ emissions of the steel industry has attracted significant interest over the years and numerous companies – accounting for 17% of the global steel production – have adopted net-zero-emission targets.⁷ While technology upgrades and improvements in heat recovery can reduce CO₂ emissions,⁸ these reductions become more incremental as technology matures. Carbon capture at steel facilities has proven to be challenging due to the high concentrations of carbon monoxide (CO), nitrogen (N₂), and steam in the flue gases, as well as the unsteady nature of furnace off-gases (i.e. the electric arc furnace; EAF), calling for a reinvention of current steel production. As the majority of current steel facilities will reach their end of life by 2030, it is crucial to start investing in green steel technologies during the 2020s.⁹ By using shaft furnaces, iron oxide can be directly reduced by methane (CH₄)-derived syngas, a mixture of CO and H₂. This can reduce CO₂ emissions by 61% compared to conventional coke-based iron production processes. When replacing CH₄ with renewable H₂, CO₂ emission reductions of 97% are possible.¹⁰ In a recent assessment of the steel industry, the International Energy Agency (IEA) further highlights the need for new low-emission steel technologies such as H₂-based direct reduced iron (H₂-DRI).⁵ Using renewable H₂ is currently still expensive, making the H₂-DRI process highly dependent upon the availability of low-cost clean electricity and/or the implementation of a CO₂ emission tax;¹¹ nevertheless, H₂-DRI shows advantages over other renewable carbon-based drop-in fuels.¹⁰

Preliminary comparisons of decarbonization pathways for iron and steelmaking suggest that hydrogen (H₂)-based iron oxide reduction in shaft furnaces has the greatest potential for lowering CO₂ emissions from primary steel production when using renewable electricity for H₂ production and serving ancillary power needs^{6,12}. Other advantages of H₂-based direct reduction of iron oxide include faster reaction kinetics compared to CO, which can reduce equipment size.¹³ However, the endothermic nature of the iron oxide reduction reaction with H₂ makes the heat integration more challenging and a pre-heater for the H₂ is needed.¹⁴ While this opens interesting heat integration options to improve the overall system efficiency, e.g. with the EAF off-gas (an often wasted resource), the high sensitivity of H₂-based steel plants to the electrolyser efficiency typically leads to lower overall systems efficiencies when compared to the traditional blast furnace and basic oxygen furnace route.¹⁵ Nevertheless, H₂-based steel production offers an enormous CO₂ emission reduction potential and technologies for direct reduced iron (DRI) production using H₂ have been successfully demonstrated with no adverse effects on steel quality.^{16–18}

This presents the opportunity for renewably generated H₂, from sustainable regional energy resources, to be coupled with the steel industry. Due to the fluctuations in the availability of renewable power and the continuous nature of the steel-making process, there is a need for H₂ storage to supply cheap, low-carbon emission H₂ continuously during operation.¹⁹

Today short-term compressed H₂ storage in steel plants – to shift the electricity use to off-peak hours – is shown to not off-set the electrolyzer investment.²⁰ However, higher price fluctuations in a future grid with high renewable penetration are expected to make onsite H₂ storage a valuable asset increasing profit margins.²¹ A drawback of compressed gas H₂ storage systems is their relatively high energy demand associated with gas compression. Thus, alternative H₂ storage systems, such as liquid organic H₂ carriers (LOHCs), are currently being considered for integration into steel facilities.²²

Advantages of LOHCs include their easy handling and storage, as they are typically liquid at ambient temperature and pressure, which makes them highly compatible with current fossil fuel infrastructure. Liquid organic H₂ carriers can be centrally produced, to make use of the economies of scale, and transported to end-use locations. Methanol is shown to be an attractive option for such scenarios which offers a similar economic performance as compressed H₂ storage.²³ When thermally integrated with end-use H₂ applications, LOHCs are shown to be thermodynamically and economically advantageous over compressed H₂ storage solutions.²⁴ Particularly, the lower capital investment cost of LOHC systems are shown to economically outperform lined rock cavern H₂ storage at H₂-DRI steel facilities.²⁵ In the same context, using methanol as LOHC is found to be more cost competitive compared to formic acid, ammonia, and perhydro-dibenzyltoluene, which suffer from large thermodynamic barriers for dehydrogenation and/or higher investment costs.²⁶

The focus of this work is to characterize plausible hydrogen end use in iron and steelmaking in a manner that can inform coupling with H₂ generation and storage systems. In this study, break-even levelized cost of hydrogen (LCOH) targets for decarbonizing the steel industry with H₂-DRI are established by comparing H₂-DRI to the commercial natural gas-based direct reduced iron (NG-DRI) process. To enable this comparison, a detailed techno-economic analysis, supported by rigorous process modelling, is conducted. Detailed process models of integrated mills with NG-DRI and H₂-DRI are developed to establish material balances, energy balances, balance-of-plant, direct and indirect CO₂ emissions, etc. This information is then used to derive the economic performance; whereby, the NG-DRI acts as the state-of-the-art reference case used to benchmark the H₂-DRI performance. Detailed breakdowns of capital expenditure and cost-driving factors are discussed and their respective impact on the levelized steel production cost (LSPC). By replacing the feedstock of the various NG users in the mill with H₂, the decarbonization potential of the mill is evaluated together with its economic performance expressed in break-even LCOH. Additionally, key operating parameters of the H₂-DRI process are studied to evaluate the economic impact of operating states upon the break-even LCOH and highlight opportunities for improving process design parameters. Specifically, EAF off-gas utilization is an important topic in this area as it contains substantial amounts of energy in the form of heat and CO, as well as smaller amounts of H₂. The highly dynamic nature of this off-gas makes it difficult to utilize. In this work, we present dynamic simulations of the EAF off-gas to investigate their utilization potential in NG-DRI and H₂-DRI applications, as well as shed light on the economic value of the respective utilization option.

In summary, while H₂-DRI has been investigated in literature, this study provides for the first time a comprehensive techno-economic comparison of NG-DRI and H₂-DRI. While most studies integrate the electrolyzer into their analysis, it needs to be recognized that economical H₂-production and economical H₂-DRI operation are two connected but distinctively separate concerns. Along those lines, we provide for the first time insights into the pure H₂-DRI economics, study H₂-DRI operating parameters and their impact upon economics as well as derive break-even/target costs of H₂ to enable economical operation.

Methodology

The system boundary is set around the physical plant site of an integrated DRI steel mill with EAF. The analysis includes the processes shown in **Figure 1** and **Figure 2** as well as EAF off-gas treatment, heat and electricity demand for ladle refining operations,

and the cooling water system. Fuels, chemicals, and feedstocks, such as electricity, NG, H₂, carbon, lime, and iron ore pellets, are considered to be delivered to the plant site at their respective procurement costs. Emissions from their production are not included with those of the plant site but are accounted for as indirect emissions in the emission analysis. Direct emissions are emissions generated within the modelled system and cross the system boundary, which is drawn around the plant's stack. Other streams leaving the system boundary include the steel slab product, the slack, and the cooling tower blowdown. In this work, we study medium-sized integrated steel facilities with an annual production capacity of 1,162,000 metric tonnes of steel slabs, or an annual production of 1,046,000 metric tonnes considering downtime due to maintenance and other outages (90% capacity factor). An NG-based DRI facility with the same production capacity is used to benchmark the technical and economic performance of the H₂-based DRI facility. Additional scenarios are modelled reflect plausible variations on how H₂ is used in the facility, and how EAF off-gas could be utilized; the former focusing on the use of H₂ to replace NG in heating applications, and the latter exploring opportunities for lowering the energy intensity of steel facilities. The reference integrated steel mill uses NG in four different processes, which could all in theory be served with H₂: I) the shaft furnace uses NG-derived syngas as reductant, II) NG is used by the reformer as fuel for the firebox, and III) NG is used in the EAF and IV) ladle refining processes.

All scenarios evaluated in the techno-economic analysis are developed in the process simulation software ProSim Plus.²⁷ We estimate total upfront investment, material and energy efficiency, and predict the maximum LCOH that is permissible in order to break even with the LSPC of the NG-DRI base case scenario. This means that a high break-even LCOH is desirable as it constitutes a scenario that is economically viable even if the cost of H₂ is high. This section provides an overview of the data and modelling approaches used to establish the performance metrics of the aforementioned technologies.

Direct Reduced Iron Steel Mill

The DRI plants investigated in this study are integrated mills with hot link, meaning that the sponge iron produced in the DRI process is not cooled to room temperature and is still hot when loaded into the EAF for further processing. A simplified flowsheet of the NG-DRI steel mill with its main process units and material streams is provided in **Figure 1**. The corresponding state-point stream data with information about temperatures, pressures, flow rates, and composition can be found in the ESI **Table S1**. A simplified flowsheet of the H₂-DRI steel mill with its main process units and material streams is provided in **Figure 2**. The corresponding state-point stream data can be found in the ESI **Table S2**.

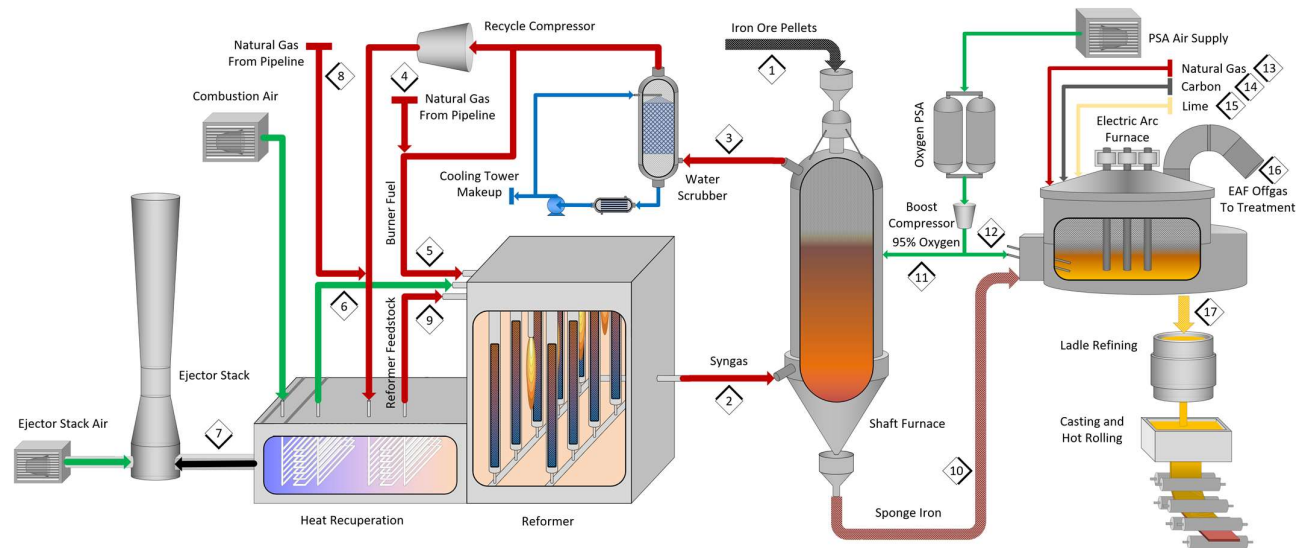
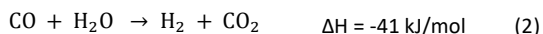
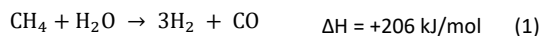


Figure 1: Simplified flowsheet of the integrated natural gas-based DRI steel mill (NG-DRI-B).

Natural Gas Reformer

The NG-based system, which acts as benchmark scenario, uses reformed NG or syngas in the shaft furnace as reducing agent for the iron oxide. The reformer operates on a mixture of NG and recycled syngas from the shaft furnace. The recycle scrubber temperature is adjusted to 70 °C to provide sufficient quantities of H₂O to the reformer¹⁴ to minimize the risk of carbon deposition. The reformer feedstock is pre-heated against the reformer's flue gas to a temperature of 500 °C prior to entering the reforming section. Inside the reformer, the steam

reforming reaction, Equation (1), and water gas shift reaction, Equation (2), create a syngas high in H₂ and CO, the reductants for reducing iron oxide to metallic iron.

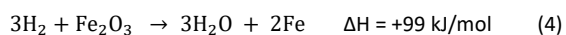
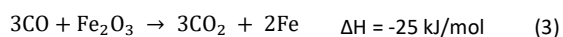


The catalyst used in this process is an alumina-supported nickel catalyst, i.e. Ko et al.²⁸ Due to the integration with the shaft furnace and to minimize recycle compression work, the reformer operates at a pressure of 2.9 bar. The outlet gas (syngas) consists of 51 mol.-% H₂, 35 mol.-% CO, 8 mol.-% H₂O, 1 mol.-% CH₄ and 5 mol.-% CO₂, which is in good agreement with literature values.^{14,29–31}

The firebox operates with an excess air of 15% and maintains a thermodynamic temperature of greater than 1000 °C in the reforming section to facilitate the endothermic reforming reactions. To maximize the efficiency of the firebox, the combustion air is pre-heated to 500 °C against the flue gas before it enters the stack. Due to the high flue gas temperature, even after recuperation, an ejector stack is used.

DRI Shaft Furnace

The syngas enters the shaft furnace at a temperature of 850 °C^{29,30,32,33} where H₂ and CO reduce the iron oxide stepwise to metallic iron. The overall reducing reactions occurring in the shaft furnace are summarized in Equations (3) and (4):



The shaft furnace is loaded with iron ore pellets at the top and the feedstock slowly moves downwards over time as it heats up. On the way to the bottom of the furnace, the iron oxide comes in contact with hot CO and H₂ which reduces the Fe₂O₃ first to Fe₃O₄, and FeO before converting it into metallic Fe. At the same time, CO and H₂ are converted to CO₂ and H₂O. To maintain a reducing atmosphere, excess CO and H₂ are needed. Higher excess ratios increase the chemical potential/driving force for the chemical reactions inside the shaft furnace; however, this increases the recycle stream and work associated with recompression. Typical top gas concentrations of the shaft furnace range from 33–49 mol.-% for H₂, and 19–26 mol.-% for CO.^{29–31} The temperature of the top gas leaving the shaft furnace ranges from 300–450 °C.^{29,30,33} Thereafter, the top gas is quenched in a water scrubber to remove water produced by the iron ore reduction process as well as dust, and to lower the gas temperature for the recycle compressor. About 1/3 of the top gas is sent to the reformer's firebox where it is burned. This purge is necessary to eliminate the build-up of CO₂ and other compounds in the recycle.

The productivity of the shaft furnace can be improved by introducing high-purity O₂ into the furnace (or upstream of the shaft furnace) to raise the operating temperature of the shaft furnace, which improves reaction kinetics. Oxygen addition can vary greatly³⁴, and a value of 0.1 kmol/metric tonne of sponge iron has been chosen in this work.

Electric Arc Furnace

The sponge iron leaves the shaft furnace at a temperature of 700 °C in this hotlink configuration and does not utilize methane injection at the bottom of the furnace to cool the sponge iron. The hot sponge iron is directly transferred to the EAF where it is melted using NG, carbon, and electricity as energy input to the EAF. EAF operating parameters are based on industrial operating data^{35,36} and best practice values.^{37,38} The EAF operates in batch mode; however, for simulation purposes, time-averaged values are used (at the process unit boundaries). At tapping, the liquid steel has a temperature of 1650 °C, and also the slag is assumed to be removed at this temperature. The average off-gas temperature was determined to be 1025 °C based on operational data available.^{35,36} However, EAF operation can vary greatly based on steel type and debottlenecking efforts to increase productivity. Natural gas, carbon, oxygen, and lime consumption are based on best practice values³⁷ slightly modified to meet energy demands (cooling accounts for 10% of energy input).³⁸ Air leakage and average gas concentrations are based on industrial operating data.^{35,36} The EAF off-gas is extracted from the EAF via water-cooled ducts and fed to a post-combustor where the off-gas is fully oxidized followed by further cooling, particle removal and venting using an induced draft fan. Ladle furnace operation, baghouse, casters, ladle heating and auxiliaries consume 22 kWh of natural gas per metric tonne of steel and 88 kWh of electricity per metric tonne of steel.³⁷

H₂-Based Steelmaking

The H₂-based DRI steel mill (Figure 2) shares many similarities with the NG-based system. Sponge iron processing downstream of the shaft furnace remains unchanged except from some operating conditions to produce the same grade of steel (0.6 wt.-% C), i.e. more carbon is needed in the EAF since the sponge iron in the H₂-DRI plant does not contain any carbon. The largest change compared to the NG-DRI plant is that the H₂-DRI plant does not need a reformer; however, due to the endothermic nature of the iron ore reduction reaction with H₂, the H₂ needs to be sufficiently pre-heated before entering the shaft furnace. This H₂ pre-heater is fueled by purged top gas from the shaft furnace (to avoid build-up of trace components such as N₂), and makeup natural gas to meet the heat load. Moreover, the oxygen addition into the shaft furnace is increased to raise the operating temperature. Having both a pre-heater and O₂ injection increases operational flexibility. The

size of the pre-heater, which pre-heats the H₂ to 775 °C is strategically chosen so that approximately 50% of the energy is added to the shaft furnace directly (oxidation of H₂) and the other 50% in the pre-heater to maximize flexibility and the ability to potentially buffer H₂ supply chain issues, which is particularly important if considering H₂ generation from 100% renewables. Furthermore, this design will allow the integration and utilization of the EAF off-gas a currently mostly wasted resource (more discussion on this in the Results and Discussion section). In the H₂-base case, the top gas outlet contains approx. 30 mol.-% excess H₂ and has a temperature of 350 °C. This operating condition has been chosen based on the reduction potential of the reducing gas. Above approx. 500 °C, the shaft furnace inlet conditions are more reducing than in the NG-DRI scenario; however, outlet conditions of the H₂-DRI scenario are less reducing than in the NG-DRI scenario.

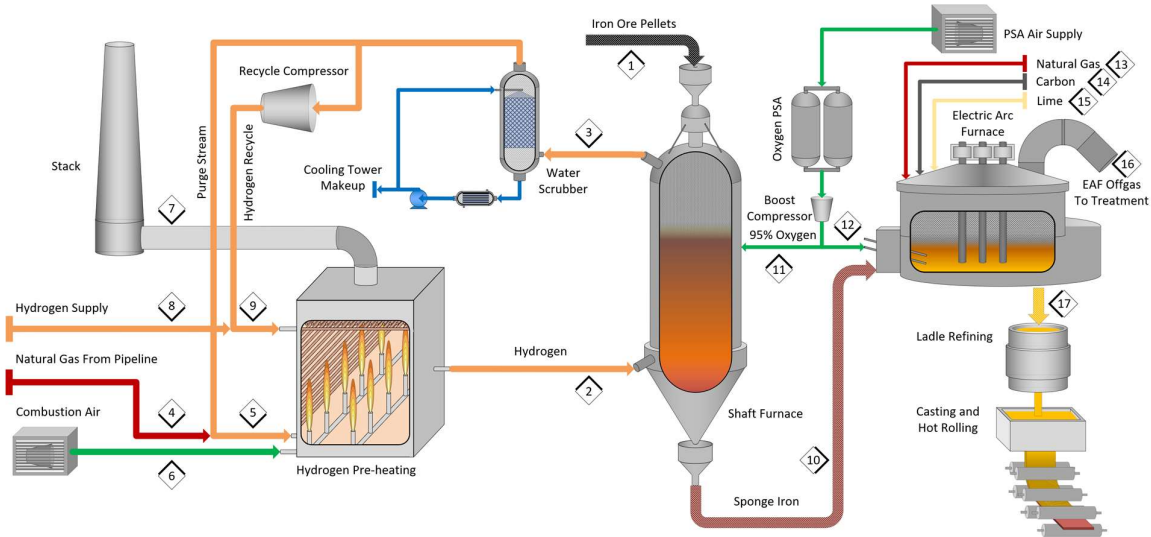


Figure 2: Simplified flowsheet of the integrated hydrogen-based DRI steel mill (H₂-DRI-B).

Design Scenarios

In the following section, we provide an overview of the different design scenarios and sensitivity analysis studied. In the course of this analysis, we present scenarios and sensitivities to gauge the significance of certain H₂-DRI operating parameters and to highlight critical areas in the research and development of H₂-based DRI. As previously mentioned, the shaft furnace operation and various heat applications in the steel facility are switched from NG to H₂ to investigate break-even prices of H₂ and CO₂ emission reduction potentials. Descriptions of the studied scenarios are summarized in **Table 1**.

Table 1: Design Scenarios

Scenario	Description
NG-DRI-B	NG-DRI base case: NG use for syngas production, reformer heating, EAF heating, ladle heating.
NG-DRI-R	NG-DRI reformer case: NG use for syngas production, EAF heating, ladle heating. H ₂ use for reformer heating.
NG-DRI-E	NG-DRI EAF case: NG use for syngas production, reformer heating, ladle heating. H ₂ use for EAF heating.
NG-DRI-L	NG-DRI ladle case: NG use for syngas production, reformer heating, EAF heating. H ₂ use for ladle heating.
NG-DRI-T	NG-DRI total case: NG use for syngas production. H ₂ use for reformer heating, EAF heating, ladle heating.
H ₂ -DRI-B	H ₂ -DRI base case: H ₂ use for shaft furnace. NG use for H ₂ pre-heating, EAF heating, ladle heating.
H ₂ -DRI-P	H ₂ -DRI pre-heater case: H ₂ use for shaft furnace, H ₂ pre-heating. NG use for EAF heating, ladle heating.
H ₂ -DRI-E	H ₂ -DRI EAF case: H ₂ use for shaft furnace, EAF heating. NG use for H ₂ pre-heating, ladle heating.
H ₂ -DRI-L	H ₂ -DRI ladle case: H ₂ use for shaft furnace, ladle heating. NG use for H ₂ pre-heating, EAF heating.
H ₂ -DRI-T	H ₂ -DRI total case: H ₂ use for shaft furnace, H ₂ pre-heating, EAF heating, ladle heating.

Secondly, a sensitivity study is conducted for the H₂-DRI-B and H₂-DRI-T cases to investigate the impact of varying H₂ excess ratios in the shaft furnace. Due to limited data availability, it remains difficult to judge the exact amount of excess H₂ needed. This sensitivity analysis will shed light on the economic impact of this operating variable together with an alternate ejector-based top gas recirculation option.

Lastly, EAF off-gas utilization is studied using the NG-DRI-B and H₂-DRI-T cases, and two exemplary EAF off-gases. EAF off-gas composition, temperature, and mass flow for EAF off-gas #1 and EAF off-gas #2 are shown in the ESI **Figure S1** and **Figure S2**. For each case (NG-DRI-B and H₂-DRI-T), four sub-scenarios are investigated: 1) EAF off-gas #1 with single train EAF, 2) EAF off-gas #1 with two contracyclical EAF trains, 3) EAF off-gas #2 with single train EAF, 4) EAF off-gas #2 with two contracyclical EAF trains.

Economics

The basis for the economic analysis is the year 2022. The LSPC is evaluated over an assumed 30-year plant operational period with a capital expenditure period of 3 years (33 years total). The total overnight cost is assumed to be 100% depreciable over 20 years at a 150% declining balance.³⁹ After-tax weighted average cost of capital for an investor-owned utility with 55% debt financing is 4.73% (real).³⁹ Tax rates are 21% (federal) and 6% (state).³⁹ This financing structure results in a capital charge factor (CCF) of 0.0776 and Equation (5) can be used to determine the LSPC.

$$LSPC = \frac{(CCF)(TOC) + OC_{fix} + (CF)(OC_{var})}{(CF)(MTPY)} \quad (5)$$

LSPC represents the cost of producing steel in the first year, calculated by taking into account factors such as the capital charge factor (CCF), the total cost of building the facility (TOC), fixed and variable annual operating costs (OC_{fix} and OC_{var}), the plant's capacity utilization (CF), and the expected annual production of steel at full capacity (MTPY). The TOC is the total overnight capital expenditure and includes the total plant cost (TPC) as well as pre-production costs, inventory capital, financing costs, land, and other owner's costs (for details see reference).⁴⁰

Fixed operating costs (OC_{fix}) include property tax and insurance at 2% of the TPC and operating labour. Operating labour for the integrated steel mill at the relevant scale is estimated with 51 skilled operators paid at an hourly rate of \$40.85 and 93 shift workers paid \$30.00 per hour. It is estimated that the labour burden accounts for 30% of the operating costs, and an additional 25% will be allocated for overhead expenses. Maintenance-related labour expenses make up 35% of the maintenance costs, and administrative and support labour are 25% of the combined operating and maintenance labour costs.⁴⁰

Variable operating costs (OC_{var}) such as maintenance expenses are dependent on the availability of the plant. Other variable costs to consider include items like fuel, sorbents, and catalysts that are consumed during the production process. A summary of the consumables used in the steel mills is provided in **Table 2** (for the analysis all costs are escalated to the year 2022 using an annual escalation factor of 3%).

Table 2: Cost Summary of Fuel and Consumables

Fuel/Consumables	Value	Unit	Cost Year	Ref.
Natural Gas	33.85	\$/MWh	2022	41
Electricity	93.40	\$/MWh	2022	42
Iron Ore Pellet	130.00	\$/tonne	2022	43,44
Slag Disposal	30.00	\$/tonne	2011	45
Solid Waste Disposal	200.00	\$/tonne	2017	46
Raw Water	0.44	\$/m ³	2011	45
Carbon	179.47	\$/tonne	2019	47
Lime	100.00	\$/tonne	2021	48
Reforming Cat. (Ni-Al ₂ O ₃)	17.52	\$/litre	2014	49

Scaling costs to the relevant analysis year can be achieved via Equation (7). To obtain cost estimates at plant scale Equation (7) is used.

$$SC = RC \cdot (1 + AER)^{SY-RY} \quad (6)$$

$$SC = RC \left(\frac{SP}{RP}\right)^u \left(\frac{TS}{TR}\right)^{0.9} \quad (7)$$

The scaled cost (SC) is determined by using the reference cost (RC), annual escalation rate (AER), scaled year (SY), and respective reference year (RY). To scale the equipment size, the scaling parameter (SP) and the reference parameter (RP) at reference scale are used along with the scaling exponent (u) which can be found in literature⁵⁰ for various types of plant equipment. The number of trains or quantity of equipment for the scaled plant is represented by TS, while TR represents the number of trains or quantity of equipment in the reference case. Additionally, an exponent of 0.9 is used to account for cost reduction when multiple units of the same equipment are purchased and installed. The expected accuracy of this methodology for cost estimation is between -30% to +50%, but scaling by more than a factor of two may increase the error margin. Capital cost estimates are based on values reported in literature⁵⁰⁻⁵⁴ and ProSim economic evaluation. A reduced order model of the CAPEX is presented in **Table 3**.

Table 3: Total Plant Cost Correlations* (\$2022)

Process Unit	Scaling Parameter X	Correlation
EAF & Casting	Liq. Steel kg/h	$1,132,370 \cdot X^{0.4560}$
Shaft Furnace	Pig Iron, kg/h	$49,080 \cdot X^{0.6538}$
Oxygen Supply	O ₂ Product Stream, kg/h	$30,622 \cdot X^{0.6357}$
Reformer	Furnace Heat Exchange, MW	$4,930,889 \cdot X^{0.6505}$
H ₂ Pre-Heater	Furnace Heat Exchange, MW	$228,860 \cdot X^{0.7848}$
Recycle Compressor	Power, MW	$6,151,202 \cdot X^{0.7100}$
Cooling Tower	Water, m ³ /h	$60,812 \cdot X^{0.6303}$
Electrical & Instrumentation	Liq. Steel kg/h	$69,819 \cdot X^{0.5584}$
Buildings, Storage, Water Service	Liq. Steel kg/h	$6,320 \cdot X^{0.8000}$
Other Miscellaneous Cost	Liq. Steel kg/h	$174,548 \cdot X^{0.5583}$
Integrated NG-DRI Steel Mill (total)	Liq. Steel kg/h	$785,087 \cdot X^{0.5857}$
Integrated H ₂ -DRI Steel Mill (total)	Liq. Steel kg/h	$800,884 \cdot X^{0.5647}$

*TPC includes EPC, process contingencies, project contingencies, etc.

Direct and indirect CO₂ emissions

Direct emissions are emissions that occur within the system boundary defined earlier. These emissions originate from the use of fuels and feedstocks. The basis for direct emissions is the individual consumption rates to satisfy process specifications, heat loads, and chemistry. Indirect emissions are emissions associated with the production and procurement of materials used in the production process but occur outside the modelled system boundaries. Indirect CO₂ emissions are estimated from upstream life-cycle processes associated with raw material feedstocks and grid electricity based upon individual emission factors obtained from the EIA,⁵⁵ Ecoinvent 3 database: electricity 386 kg_{CO2}/MWh, natural gas 0.44 kg_{CO2}/kg_{NG}, metallurgical coal 0.98 kg_{CO2}/kg_C, lime 0.05 kg_{CO2}/kg_{Lime}, iron ore mining 0.02 kg_{CO2}/kg_{Ore}, iron ore pelletizing 0.16 kg_{CO2}/kg_{Ore}.⁵⁶ Hydrogen emissions = 0 from the GREET model. While emissions from nuclear, solar, and wind-powered H₂ generation from water electrolysis can be non-zero, emission factors are typically less than 1 kg_{CO2eq} per kg H₂. As such, no emissions are associated with H₂, as suggested by the GREET model for solar H₂.⁵⁷

Results and Discussion

State-of-the-art natural gas DRI steel mill

The NG-DRI reference plant is analysed to determine benchmark values for efficiency, CO₂ emissions, and economic performance. In order to produce 1,045,000 metric tonnes of steel per year the shaft furnace processes 1,704,000 metric tonnes of iron ore annually, which consumes 11,388,000 GJ of natural gas. Additionally, 606,500 MWh of electricity are consumed whereby the EAF is responsible for over 62% of the electric load. This corresponds to a specific power consumption of 355 kWh/metric tonne of liquid steel whereby energy savings due to hotlink are 98 kW/h, which is in good agreement with literature.³⁷ The carbon consumption in the EAF is 27 kg_C/tonne_{Steel(lq.)}, and lime is 50 kg_{Lime}/tonne_{Steel(lq.)}. About 10% of the EAF's total energy is lost through the reactor walls³⁸ which is captured by the cooling system and rejected in the cooling tower. The cooling tower consumes 1,337,000 metric tonnes of raw water per year, whereby 373,000 metric tonnes of cooling tower blowdown end up as surface discharge. Another disposal stream is the slag which amounts to 182,000 metric tonnes per year. Considering the total energy input, the thermal efficiency of this process is 54.3%-LHV. A summary of the energy balance and balance-of-plant power consumption is shown in the ESI **Table S3**. Direct plant CO₂ emissions are 0.675 kg_{CO2}/kg_{Steel}. The EAF accounts for 0.125 kg_{CO2}/kg_{Steel} and the DRI process for 0.551 kg_{CO2}/kg_{Steel}. The total (direct + indirect) CO₂ emissions of the plant are 1.318 kg_{CO2}/kg_{Steel}. Indirect emissions are dominated by the iron ore pelletizing process (0.261 kg_{CO2}/kg_{Steel}) followed by grid emissions (0.224 kg_{CO2}/kg_{Steel}). A breakdown of the emission sources and their contribution to the overall emissions is provided in **Figure 3**.

The economic analysis shows that the TPC is \$795.5M which translates to a specific plant cost of \$685 per metric tonne of steel (based on annual production capacity). Considering preproduction costs, inventory capital costs, and other owner costs such as land, financing, etc., the total overnight capital cost is \$1,074.5M. Variable operating costs are \$414.9M per year (at 90% capacity factor), whereby the iron ore feedstock costs account for \$221.5M. Natural gas and electricity expenses are \$106.8M and \$56.6M, respectively. Other consumables such as water, carbon, and lime are responsible for \$12.1M. Maintenance materials add another \$10.8M annually and disposal costs for slag and other solids cost \$7.1M per year. The fixed operating costs account for an annual expenditure of \$110.6M, which is dominated by the operating labour with \$69.4M followed by tax and insurance costs of \$15.9M. Maintenance labour and administrative labour are \$6.4M and \$18.9M. The resulting LSPC is \$582.18 per metric tonne of steel. Hot-rolled steel traded at around \$670 per metric tonne towards the end of the year 2022.⁵⁸ However, in recent years, steel prices experienced large fluctuations due to market dynamics originating from tight supplies and high demand.⁵⁹ A breakdown of the individual cost-driving factors is shown in **Figure 3**.

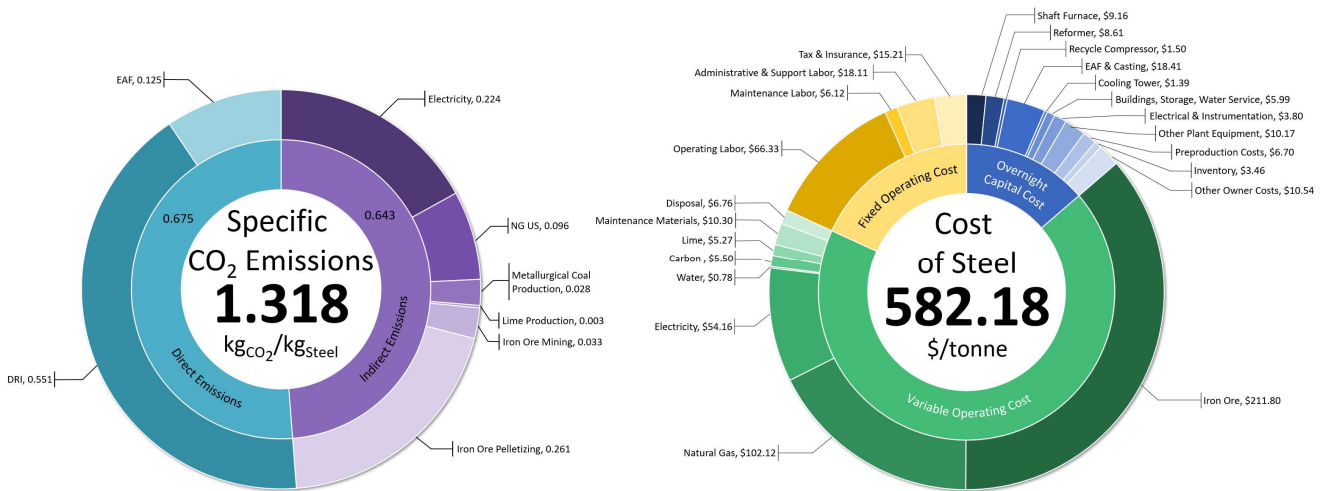


Figure 3: Breakdown of carbon dioxide emissions (left) and levelized cost of steel (right) for the integrated NG-DRI steel mill.

Hydrogen in DRI steel mills

By switching the shaft furnace operation from natural gas-derived syngas to 100% renewable H₂ (H₂-DRI-B), an onsite CO₂ emission reduction of 76.3% can be realized. Additionally, indirect emissions are decreased by 14.3% primarily due to reduced NG use; however, increased carbon (coal)-use in the EAF counteracts some of these effects. The great advantage of using H₂ as reductant is the reduced capital investment as the reformer becomes obsolete. This reduced capital investment as well as savings in operating costs (no catalyst needed) help to increase the break-even LCOH to \$1.70/kg_{H2} making this scenario an economically viable option long before the U.S. DOE target of \$1.00/kg_{H2} is reached (without any CO₂ credits).⁶⁰

Next, we discuss replacing NG with H₂ as heat source. In the case of an NG-DRI, the ladle refining operation (NG-DRI-L) has little impact on the CO₂ emissions due to the small quantities of NG used in the process and only a CO₂ emission reduction of 0.6% can be achieved (direct emissions). To economically achieve this emission reduction, an LCOH of \$1.14/kg_{H2} or lower is needed. Replacing the NG used in the EAF for heating with renewable H₂ (NG-DRI-E) can lower the direct CO₂ emissions by 2.8% at a break-even LCOH of \$1.06/kg_{H2}. The amount of NG used in the reformer as heat input to the firebox is substantially higher and replacing this energy carrier with H₂ can reduce the direct CO₂ emissions of an integrated NG-DRI steel mill by 15.9% (NG-DRI-R). Similarly to the previous heat applications, the break-even cost of H₂ is \$1.19/kg_{H2}. This indicates that in order to economically replace NG with H₂ in heat applications in an NG-DRI steel mill, the LCOH needs to be close to the U.S. DOE target of \$1.00/kg_{H2}. Combining all these measures (NG-DRI-T), a reduction of direct CO₂ emissions of 19.4% is achievable at a break-even LCOH of \$1.16/kg_{H2}. Indirect emissions are only minimally impacted by replacing NG in heat applications with H₂, a reduction of 0.021 kg_{CO2}/kg_{Steel}.

Using H₂-DRI and replacing NG as heat source for ladle refining, EAF, and H₂ pre-heating (H₂-DRI-T case) can further the reduction of direct CO₂ emissions of the H₂-DRI-B case from 76.3% to 84.9%. Indirect emissions are only reduced by 0.009 kg_{CO2}/kg_{Steel} (0.02%) compared to the H₂-DRI-B case and emissions from electricity generation and iron ore pelletizing account for over 73.5% of the total CO₂ emissions (direct + indirect). As previously discussed, replacing NG in heat application is expensive compared to replacing NG as reductant in the shaft furnace; however, the large quantities of H₂ needed in the shaft furnace help

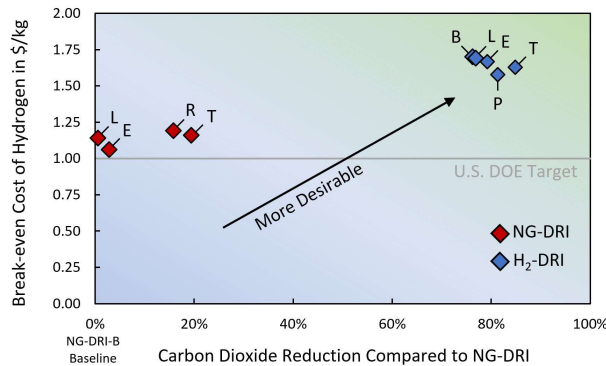


Figure 4: Carbon dioxide emission reduction potential and associated H₂ break-even prices for the various NG consuming processes in an integrated DRI steel mill.

to stabilize the break-even LCOH at \$1.63/kg_{H2} when adding H₂-based heat applications in an H₂-DRI steel mill. Hence, the most economical way to decarbonize DRI steel mills is to start with switching the shaft furnace operation from NG-derived syngas to renewable H₂. Additional information on H₂-DRI performance and economics can be found in the ESI.

Sensitivity Analysis

The amount of excess H₂ needed in the shaft furnace is an active area of research. The iron ore reduction potential of H₂ is greater than the reduction potential of syngas at shaft furnace inlet conditions explaining some of the fast reaction kinetics observed at high H₂ mole fractions.¹³ However, at outlet conditions the opposite is true. At shaft furnace outlet conditions, the syngas has a higher reduction potential than the H₂-DRI top gas. Nevertheless, operating temperature also plays a crucial role in process design and provides an opportunity to optimize operating conditions and the excess H₂ required. **Figure 5** shows the break-even cost of H₂ for a range of excess H₂ mole fractions in the shaft furnace top gas considering a constant H₂ reductant feed rate (Stream 8). The red line represents the H₂-DRI-B case with NG used to pre-heat the H₂ reductant fed into the shaft furnace, and the blue line represents an H₂-DRI-T case that uses H₂ to pre-heat the H₂ for the shaft furnace. Using higher excess ratios, while maintaining the same tail gas temperature, has three major effects on the plant operation: I) a higher H₂ excess ratio leads to a larger H₂ recycle stream increasing recycle compression power, II) a larger recycle stream requires more energy for pre-heating, and III) a larger feed stream into the shaft furnace reduces the O₂ demand in the shaft furnace as the thermal inertia increases. While phenomena I) and II) lead to an economic performance decrease, some of this is counteracted by the phenomena described in III). From an economic perspective it is desirable to operate the shaft furnace at low H₂ excess ratios and past the 40 mol.-% mark a steeper decrease in economic performance is observed. By reducing the H₂ excess from 59 mol.-% to 21 mol.-%, the break-even cost of H₂ can be increased by 10¢ per kg of H₂. The absolute change remains the same for the NG-pre-heater case (H₂-DRI-B) and the H₂ pre-heater case (H₂-DRI-T); however, considering the lower break-even cost in the H₂-DRI-T case this change is of higher relative importance with variations of -5.4% (59 mol.-% H₂ excess) and +0.9% (21 mol.-% H₂ excess) when compared to the base case scenario with 29 mol.-% H₂ excess.

Lower H₂ excess ratios have another advantage with respect to the plant design. Hydrogen excess ratios of approximately 30% and lower support the use of inexpensive static ejectors to facilitate the recycling of the unused excess H₂. If the primary H₂ feedstock is available at sufficiently high pressures, i.e. 30 bar which is typical for PEM electrolyzers, this pressure can be utilized to recompress the recycle stream without the need for a mechanical compressor. Capital cost savings on the compressor alone are over \$7.6M. Additionally, reducing the electrical load by 1.18 MW reduces the annual electricity costs by \$0.9M. As a result, using an ejector instead of a mechanical compressor can improve the break-even LCOH by another 3¢ per kg of H₂. While high-pressure H₂ generation (around 30 bar) has become common practice, this pressure requirement might pose challenges for certain H₂ storage technologies that might be used to buffer intermittencies of renewable electricity generation to ensure a steady supply of H₂.

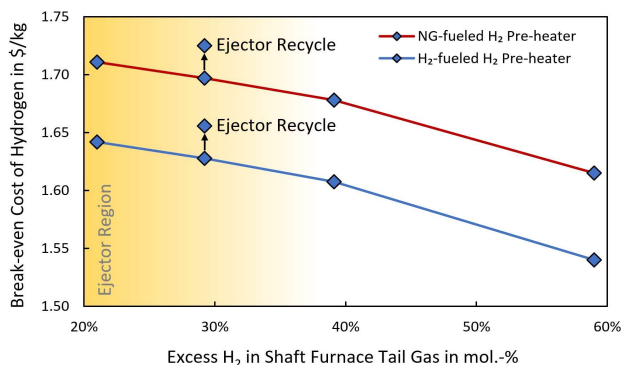


Figure 5: Break-even cost of H₂ as function of excess H₂ in H₂-DRI tail gas for NG-fueled H₂ pre-heater and H₂-fueled H₂ pre-heater. At low excess ratios the recycle compressor can be replaced with an ejector if the primary H₂ is supplied at sufficiently high pressure indicated by the yellow region.

EAF off-gas utilization scenarios

The off-gases from the EAF contain a considerable amount of energy in the form of heat and chemical energy (CO, H₂). Utilization of the off-gas has shown to be challenging because the off-gas composition not only varies substantially between different batches,⁶¹ but also, during the different phases of operation (loading, heating, oxygen blowing, pre-tapping, tapping, etc.; see ESI). While typical off-gases contain on average 7–12 mol.-% CO and 0–3 mol.-% H₂, momentary CO and H₂ concentrations can be as high as 60 mol.-% and 20 mol.-%, respectively.^{35,62} Due to these large fluctuations, EAF off-gases are commonly just cooled, oxidized, further cooled, and treated before the off-gas is eventually emitted into the atmosphere without utilizing any of its heat

or caloric value. In the following section, the aforementioned two exemplary EAF off-gases are used to study their utilization potential in the NG-DRI-B scenario and in the H₂-DRI-T scenario.

In the NG-DRI case, heat is needed in the firebox of the reformer to drive the endothermic reforming reactions, which requires a total heat input of approximately 202.1 MW-LHV. In the NG-DRI-B case, this energy input is partially provided by the DRI top gas purge and partially by a supplemental NG support fuel stream. In the scenario with EAF off-gas utilization, the off-gas is cooled, treated, and compressed (without oxidizing), and added into the reformer's firebox. The NG support fuel flow is then adjusted to meet the thermal load of the reformer. Due to the highly dynamic nature of this operation, the off-gas is analyzed dynamically in 30s intervals. Using EAF off-gas #1 (**Figure S1**) in the reformer firebox leads to an increase in fuel consumption by 1.7% confirming current industry practice as best practice scenario. Challenging for the utilization of the off-gas is its overall low heating value, which lowers the adiabatic flame temperature, making it more difficult to provide large quantities of high-temperature heat for the reforming reactions (over 1000 °C). Heating the non-combustible gases in the EAF off-gas to these temperatures adds a thermal penalty of 10.7 MW-LHV while the off-gas itself only contains 9.5 MW-LHV, leading to an overall increase in fuel consumption. Since most of the combustible gas output is present in the second half of the EAF batch operation, one could try to only use the EAF off-gas when its heating value reaches a certain threshold value; however, this would require more advanced control strategies as well as additional equipment for off-gas treatment during times when the off-gas is not sent to the reformer introducing new economic uncertainties.

Operating two EAFs in parallel, with their cycles 50% offset, can lead to a steadier off-gas; however, this does not change the time-averaged composition of the off-gas leading to the same 1.7% increase in fuel consumption. Off-gas #2, as shown in **Figure S2**, has a higher CO mole fraction and lower mass flow rate per tonne of liquid steel compared to off-gas #1. As a result, the thermal penalty associated with gas heating is reduced to 6.3 MW-LHV. With an off-gas energy content of 9.4 MW-LHV, the NG support fuel consumption can be reduced by 4.2% in this scenario.

In the H₂-DRI-T case, heat is needed in the H₂ pre-heater. The pre-heater requires 40.6 MW-LHV. In the H₂-DRI-T case, this energy is partially provided by the DRI top gas purge and partially by supplemental H₂ fuel. Similarly, to the previously-discussed NG case, the EAF off-gas is cooled, treated, and compressed (without oxidizing), before it is combusted in the pre-heater firebox. To meet the heat load, the supplemental H₂ fuel flow is adjusted as needed. Using EAF off-gas #1 in a single train setup shows a substantial 30.8% reduction in the pre-heater's fuel consumption. Since the H₂ pre-heater operates in a much lower temperature window and can also make use of low-quality heat compared to the reformer (H₂ is pre-heated from 55 to 775 °C), the penalty associated with heating non-combustible gases is less problematic and reduces to 2.5 MW-LHV. A new observation specific to the H₂-DRI pre-heater cases is that due to the high CO concentrations at certain times of the EAF operation and the relatively low heat demand of the pre-heater, some energy contained in the off-gas is wasted. Between minutes 50' and 58' as well as between minutes 66' and 68', the energy provided by the off-gas exceeds the heat necessary to pre-heat the H₂-DRI feed as seen in **Figure 6e**). To capture more of the caloric value contained in the EAF off-gas, the EAF operation can be performed in two parallel trains, with their cycles 50% offset. With this more balanced off-gas, excessive heating can be almost completely eliminated which increases the fuel savings in the pre-heater to 41.5% (**Figure 6f**). Similar behavior is observed for off-gas #2 with pre-heater fuel savings ranging between 34.1–40.4%. An overview of the different EAF off-gas utilization scenarios is provided in **Figure 6**.

The NG-DRI-B case shows that with EAF off-gas utilization the LSPC remains almost unchanged. With a 4.3% reformer fuel flow reduction in that scenario, the LSPC reduces by less than 0.1% resulting in a cost of \$581.70 per metric tonne of steel. The benefits of EAF off-gas utilization are much more apparent in the H₂-DRI cases. EAF off-gas utilization in the NG-DRI-B case with NG as support fuel for the H₂ pre-heater can increase the break-even LCOH by 5¢ from \$1.70 to \$1.75 per kg of H₂. This effect is even more pronounced if H₂ is used as support fuel in the H₂ pre-heater (NG-DRI-T case), since H₂ is an expensive fuel for heating applications compared to NG. In that case, EAF off-gas utilization is able to increase to break-even LCOH by 7¢ from \$1.63 to \$1.70 per kg of H₂. This confirms that the economic value of EAF off-gas utilization in NG-DRI is minimal at best; however, in H₂-DRI, EAF off-gas utilization is shown to be a valuable asset to reduce support fuel consumption and improve economic performance.

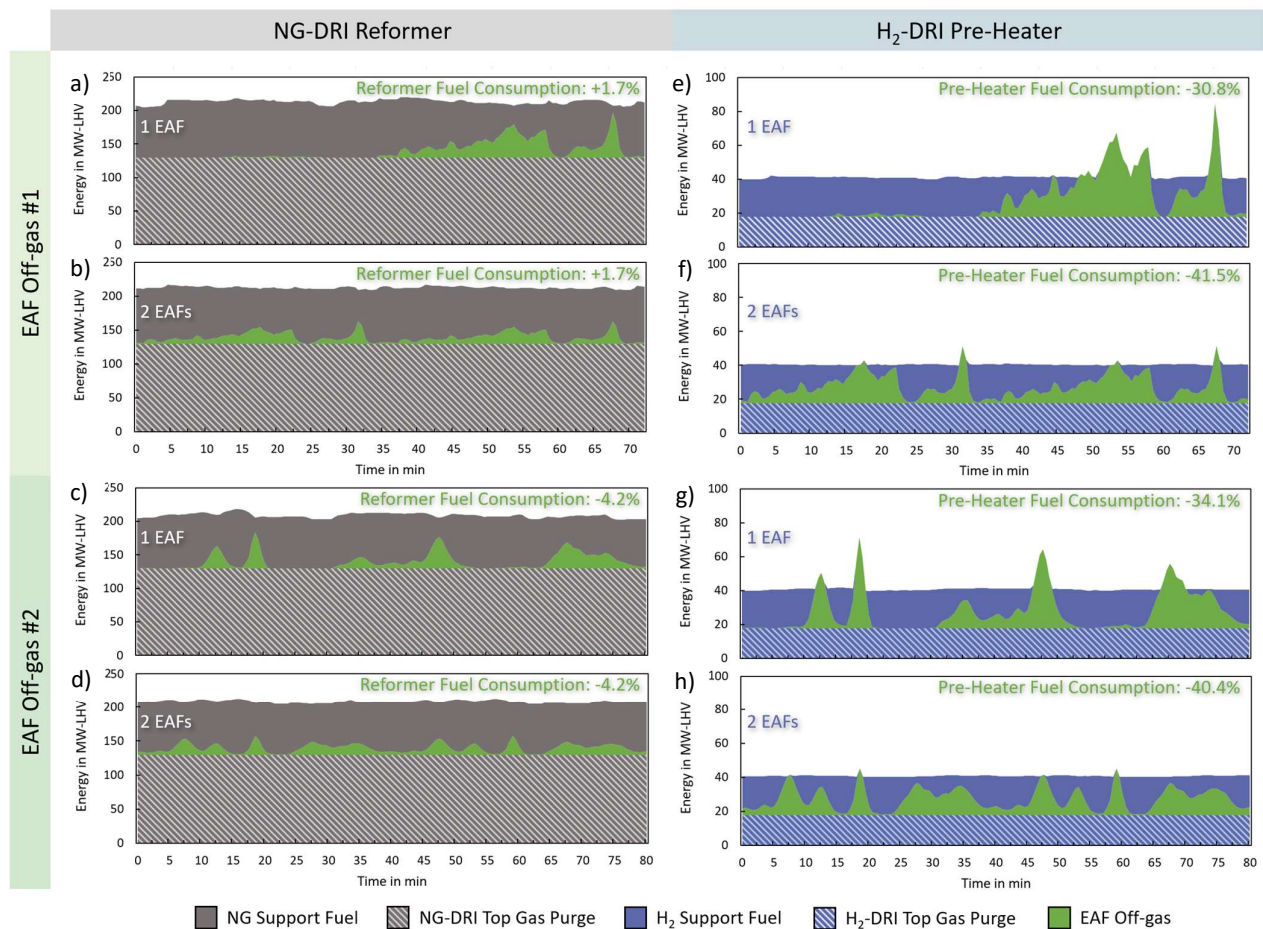


Figure 6: EAF off-gas utilization analyses for the NG-DRI and H₂-DRI cases. Figures a–d) showing the fuel energy input (LHV) into the reformer firebox by fuel source and Figures e–h) showing the fuel energy input (LHV) into the H₂ pre-heater by fuel source. Two different EAF off-gases are studied. Off-gas #1 is presented in Figures a,b,e,f) and off-gas #2 is presented in Figures c,d,g,h). For each of the combinations 2 EAF setups have been considered. Single train EAF operation is presented in Figures a,c,e,g) and two-train operation with 50% cycle offset is presented in Figures b,d,f,h).

Conclusions

In this work, the authors investigated the performance of integrated steel mills with NG-DRI and H₂-DRI to establish target costs for renewable H₂ production. The target values are break-even prices of H₂ that need to be achieved in order to maintain the identical LSPC as in the NG-DRI reference case.

The various NG users in the reference case were replaced with H₂, showing that replacing NG-based heat applications in the NG-DRI mill with H₂ leads to a relatively small reduction in CO₂ emissions while requiring a low cost of H₂ ranging from \$1.06–\$1.19 per kg of H₂. However, switching the shaft furnace operation from NG to H₂-shaft is shown to reduce capital costs, due to the omission of the reformer, leading to a break-even cost of H₂ of \$1.70 per kg, while reducing direct CO₂ emissions by 76.3%. Furthermore, converting the shaft furnace from NG to H₂ helps to stabilize the break-even cost of H₂ when switching the remaining NG heat applications to H₂. As a result, a CO₂ emission reduction of 84.9% is reached at an H₂ break-even price of \$1.63 per kg of H₂. After switching all NG users to H₂, the largest CO₂ emissions originate from indirect emitters; predominately, the iron ore pelletizing process and electricity generation. This suggests that renewable electricity, and H₂-DRI that can operate on iron ore fines rather than pellets are needed to further reduce CO₂ emissions. The third largest CO₂ source in H₂-DRI (with H₂ heat applications) is direct emissions from the use of coking coal in the EAF.

Furthermore, lower H₂ excess ratios are shown to support higher break-even prices despite an increase in oxygen demand. Over a range of 21% to 59% H₂-excess ratios, a 10¢ change in break-even cost of H₂ is observed. Additionally, excess ratios of around 30% or less allow the use of static ejectors to facilitate the H₂ recycle, eliminating the need for large recycle compressors if the primary H₂ supply pressure is sufficiently high, further boosting the break-even cost of H₂ by 3¢ per kg of H₂.

Lastly, EAF off-gas utilization has been investigated. The results show that utilization of the off-gas in NG-DRI is difficult due to the low heating value and the need for high-temperature heat in the reformer, confirming that the flaring of EAF off-gases can be considered as best practice. However, for the H₂-DRI scenario, it is found that due to the very different thermal load profile of the H₂-pre-heater, EAF off-gas utilization can reduce the primary fuel consumption by up to 41.5% or up to 7¢ per kg of H₂ in terms of break-even costs.

Author Contributions

Conceptualization FR, HB, TA, KB, RA, DP; Methodology FR, HB; Software FR; Validation FR, KB, AT, DP, HB; Formal analysis FR, KB, KY, DP; Investigation FR; Resources HB; Writing - Original Draft FR, HB, KB, TA, RA; Writing - Review & Editing FR, HB; Visualization FR; Supervision HB; Funding Acquisition HB, TA, KB, DP, RA

Conflicts of interest

There are no conflicts to declare.

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