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Excess Heat-Driven Carbon Capture at an Integrated Steel Mill – Considerations for Capture Cost Optimization

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Abstract:

Primary steelmaking in blast and basic oxygen furnaces is inherently carbon-intensive. Partial capture, i.e., capturing only a share of the CO_2, is discussed as an option to reduce the cost of carbon capture and storage (CCS) and to realize a near-term reduction in emissions from the steel industry. This work presents a techno-economic assessment of partial capture based on amine absorption of CO_2. The cost of steam from excess heat is assessed in detail. Using this steam to drive the capture process yields costs of 28 – 50 €/t CO_2-captured. Capture of CO_2 from the blast furnace gas outperforms end-of-pipe capture from the combined-heat-and-power plant or hot stove flue gases onsite by 3-5 €/t CO_2-captured. The study shows that partial capture driven exclusively by excess heat represents a lower cost for a steel mill owner, estimated in the range of 15-30 €/t CO_2-captured, as compared to full capture driven by the combustion of extra fuel. In addition, the full-chain CCS cost (capture, transport and storage) for partial capture is discussed in light of future carbon prices. We conclude that implementation of partial capture in the steel industry in the 2020s is possible and economically viable if policymakers ensure long-term regulation of carbon prices in line with agreed emission reduction targets beyond Year 2030.

Keywords: MEA, steel making, partial capture, CCS, excess heat, cost estimation

Nomenclature:

ASU	Air separation unit	HL	Heat level
BF	Blast furnace	HRC	Hot rolled coil
BFG	Blast furnace gas	HS	Hot stoves
Bio-CHP	Biomass-fired CHP plant	ICA	Intercooled absorber
BOF	Basic oxygen furnace gas	MEA	Monoethanolamine
BOFG	Basic oxygen furnace gas	MSR	Market Stability Reserve
CAPEX	Capital expenditures	NOAK	Nth-of-a-kind
CDQ	Coke dry quenching	OPEX	Operational expenditures
CHP	Combined heat and power	RSS	Rich solvent splitting
COG	Coke oven gas	c_{carbon}	Carbon price projection, €/t CO_2
DCC	Direct contact cooler	c_{NAC}	Net abatement cost, €/t CO_2
DSG	Dry slag granulation	c_{power}	Electricity price, €/MWh
EAC	Equivalent annualized capture cost	c_{steam}	Cost of steam, €/tonne steam
EDF	Enhanced detailed factor	$c_{t\&s}$	Transport and storage cost, €/t CO_2
EU ETS	EU emissions trading system	m_{steam}	Amount of recovered steam, tonne/annum
EUA	European Union Allowance	$P_{gain,BioCHP}$	Power generated from bio-CHP, MWh/annum
FGHR	Flue gas heat recovery	$P_{loss,CHP}$	Power loss linked to steam supply from CHP to capture unit(s), MWh/a

1 Introduction

The iron and steel industry emits about 8% of the global direct CO_2 emissions. More than 70% of the world's steel is produced in blast (BF) and basic oxygen (BOF) furnaces, which rely on fossil fuels for energy and for reducing the iron ore (World Steel Association, 2017). Amine absorption of CO_2 is a mature technology for CO_2 separation at a technology readiness level of 9 (IChemE Energy Centre, 2018), i.e. commercially available. The technology has therefore been proposed as a means for carbon capture and storage/utilization (CCS or CCU) for near-term reductions of emission from the steel industry (Eurofer, 2013; Fischedick et al., 2014; Wörtler et al., 2013). Carbon capture from the steel industry is low-cost compared to other industrial sources like petroleum refining (Bains et al., 2017; Leeson et al., 2017) due to high concentrations of CO_2 and large flows of off-gases emitted from integrated steel mills (Ho and Wiley, 2016; Leeson et al., 2017). Today, there is one large-scale (capture capacity of 0.8 Mt CO_2) demonstration plant from steel mill gases in operation – at the direct-reduced iron plant in Abu Dhabi (Global CCS Institute, 2018). There, the CO_2 is captured downstream of the shaft reactor, which is powered by syngas, and utilized for enhanced oil recovery.

The coal used in integrated steel mills (BF-BOF route) has multiple purposes, which make it a challenge to achieve deep carbon reduction. Integrated steel mills have several emission points. Yet, partial capture of CO_2 from the major stacks, i.e. power plant, hot stoves, coke ovens, sinter plant, and lime kiln, would reduce considerably the site emissions. Studies of capture from these stacks applying 90% separation rate in the absorber with a 30 wt.% aqueous MEA solvent have estimated a mitigation potential of 50%–80% of all site emissions at an avoidance cost of 60–100 $€_{2015}$ per tonne CO_2, depending on how many stacks are included and which assumptions are applied to the energy supply and cost parameters (Arasto et al., 2013; Cormos, 2016; Ho et al., 2013; IEAGHG, 2013; Tsupari et al., 2013). The present work focuses on the stacks with high gas flow and CO_2 concentration, and, thus, prospectively, with low capture cost, and adapts the capture rate to match the available excess heat.

In steel mills, it may be beneficial in terms of energy efficiency and process control to separate CO_2 from the process gases prior to their combustion, although > 20% of the carbon is in the form of CO. These process gases include the blast furnace gas (BFG), coke oven gas (COG), and basic oxygen furnace gas (BOFG), all of which are rich in CO, H_2 and CO_2. Currently, these gases are combusted for heat generation in the power plant, hot stoves, coke ovens, lime kilns, or in a walking beam furnace. Separation of CO_2 from these process gases would increase the gas heating value, decrease the gas volume that needs to be handled, and increase the reducing potential of the gas. BFG comprises around 70% of the CO_2 site emissions and is typically pressurized to around 2–3 bar; its relatively high CO_2 partial pressure makes it especially suitable for carbon capture. Carbon capture from BFG using amine absorption, without modifying the blast furnace to enable top gas recycling, has previously been studied (Dreillard et al., 2017; Ho et al., 2013). These studies have generally concluded that capture from process gases has lower specific capture cost but lower CO_2 reduction potential relative to capture from the stacks. Dreillard et al. have shown that the co-absorption of CO by MEA is negligible and that the CO_2/CO selectivity is high, with a CO_2 purity level of >99.5% being achieved (Dreillard et al., 2017). In the same study, the absence of oxygen in the BFG was shown to reduce solvent degradation compared to capture from the flue gases. Techno-economic studies of BFG capture with 30 wt.% MEA have reported 19%–30% reduction in site emissions at an avoidance cost of 54–72 €$_{2015}$ per tonne CO_2 (Dreillard et al., 2017; Ho et al., 2013; Kim et al., 2015; Kuramochi et al., 2012).

All the studies discussed above have assumed a 90% separation rate in the absorber and have sought to combine stacks or capture from the largest stacks to achieve an "as-high-as-possible" reduction in emissions. Usually, it is proposed that heat be provided by additional fossil fuel combustion, thereby incurring extra investment, operating costs, and CO_2 emissions. This approach, which in our previous work on partial capture for process industry was defined as the *full capture* approach, seeks to minimize the specific investment cost for carbon capture (Biermann et al., 2018). In contrast, *partial capture* seeks to reduce the operating cost and, thereby, the overall capture cost, by capturing only a share of the accessible CO_2 from a flue gas or process gas. The magnitude of this share is governed by economic factors, such as energy prices and policy-driven requirements. Situations that are potentially amenable to partial capture include, for example, industrial sites that have available, low-value excess heat or have multiple stacks that allow only the most suitable stacks to be targeted for capture. An integrated steel mill typically meets both criteria.

A previous study by the authors (Sundqvist et al., 2018) examined how the excess energy from the steel mill in Luleå, Sweden, that is currently used for district heating, process heat, and electricity production could be extended to drive also partial capture. The heat sources, which ranged from power plant steam (back-pressure operation) to the installation of excess heat recovery units, were mapped, and they allowed for a reduction of up to 43% in site emissions. It was found that partial capture from BFG gave a lower specific heat demand compared to end-of-pipe capture from the power plant. Furthermore, the increase in the heating value of BFG due to CO_2 removal allowed for re-allocation of the process gases in the steel mill, thereby releasing additional excess heat from certain process units to the capture process.

The present work extends our previous study (Sundqvist et al., 2018) to a techno-economic assessment of partial capture in the iron and steel industry through utilization of excess heat. The work illustrates how the reduction in emissions (capture rate) and the corresponding capture cost are governed by the CO_2 source and the level of available excess heat. The emphasis here is on the difference in cost between steam from excess heat and additional combustion. Three suitable CO_2 sources, hot stove flue gases, power plant flue gases, and BFG are analyzed for various capture rates and levels of heat supply. Partial capture scenarios are defined and compared with full capture benchmarks from the present study and from the literature. From this we discuss partial capture as a near-term mean option for carbon mitigation for the iron and steel industry. In addition, the time perspective and conditions in terms of carbon pricing for such near-term implementation are presented.

The Methods section describes the capture scenarios, process modeling, and cost estimation approaches. The Results section is divided into a technical section on capture performance and a section on economics. The latter highlights the cost of steam and Capital Expenditure (CAPEX) before aggregating both CAPEX and Operational Expenditure (OPEX) into a specific capture cost for different capture rates from the three main CO_2 sources in the steel mill. A sensitivity analysis highlights the main capture cost-driving parameters before the entire CCS cost chain (capture, transport and storage cost) is discussed for three carbon price projections. Finally, in the Discussion section, the findings are interpreted and compared to the results from the literature.

2 Methods

Figure 1 shows the setup and scope of the techno-economic assessment of the MEA CO_2-absorption unit integrated with an existing steel mill. Established modeling tools for the heat and mass balances of the steel mill and the capture unit are used (Sundqvist et al., 2018). In brief, the steel mill model determines the available excess heat and gas properties, which are used as inputs to the capture model. The capture model determines the achievable level of CO_2 capture and the lean gas compositions, which are used to iterate the flue gas flow and process gas composition to the steel mill model. To benchmark against full capture, two scenarios include external heat supply by an additional CHP plant fired with low-grade biomass are considered. The cost of erecting and operating the capture unit covers the costs for capture, CO_2 compression, heat supply, and the piping used to connect the CO_2-rich gases and steam to the designated capture site locations.

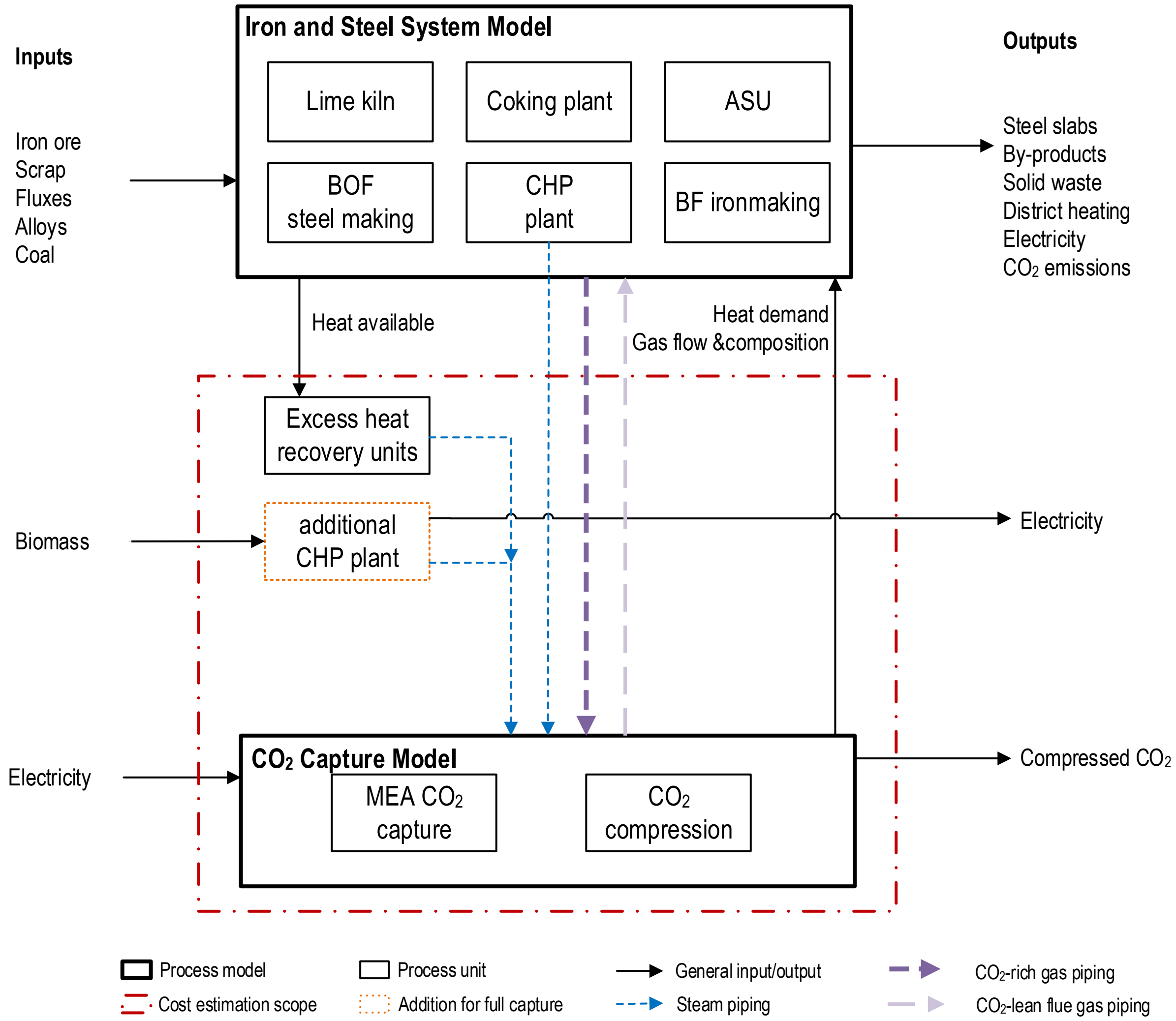


Figure 1: Overview of the methodology applied in the present work. Included are the scope of the steel mill model, the capture unit model, and the techno-economic assessment.

2.1 Capture scenarios studied

The SSAB site in Luleå has a production rate of around 2.0 Mt of primary slabs per year. In total, the plant site emits around 1.7 tonne CO_2/tonne steel slab produced. The major features of the SSAB plant that distinguish it from other integrated iron and steel plants are that: 1) the blast furnace is only charged with iron ore pellets (no sinter); and 2) downstream treatment of the steel slabs after casting does not take place onsite, but at a separate rolling mill and coating plant. Figure 2 shows the carbon balance of the Luleå site. Carbon is mainly expended for energy and iron ore reduction and only a small amount is found in the product, 98% of the carbon is emitted as CO_2. In line with the shown carbon balance, this work considers capture from the largest carbon sources, i.e., the blast furnace gas, CHP plant flue gases, and hot stove flue gases. The gas properties of these three CO_2 sources are listed in Table 1. The possible heat sources for powering the regeneration of the solvent at 120°C are considered in the following order:

1) Recovery of excess heat for which no additional direct emissions from combustion arise, and for which only the collection and distribution costs are considered. Table 2 lists five excess heat sources at the Luleå steel mill, as previously identified by the authors (Sundqvist et al., 2018).

2) Utilization of available capacity in the existing energy infrastructure. In this case, an augmented boiler capacity is omitted, since the boilers onsite already run at full load throughout the year.

3) Installation of an additional heat supply for which the emissions and costs for the extra primary energy consumption and the required investment are considered. Table 2 includes one additional external heat source in which the level of excess heat is insufficient to meet the capture target in the full capture scenarios

Note that the values in Table 2 are given as yearly averages. The order, from top to bottom, represents increased technical implications/decreased accessibility for recovering heat in the form of saturated steam at 3 bar (~133°C). Note that the amount of assessed heat for each heat source in Table 2 is valid for the Luleå reference mill without CO_2 capture. Importantly, Table 2 also provides the definitions for heat levels 1–6 in the two columns to the right. Starting with the first heat source (HL1), each progressive heat level includes the preceding heat sources, such that the total amount of recovered heat is accumulated, e.g., HL6 implies the utilization of all six heat sources.

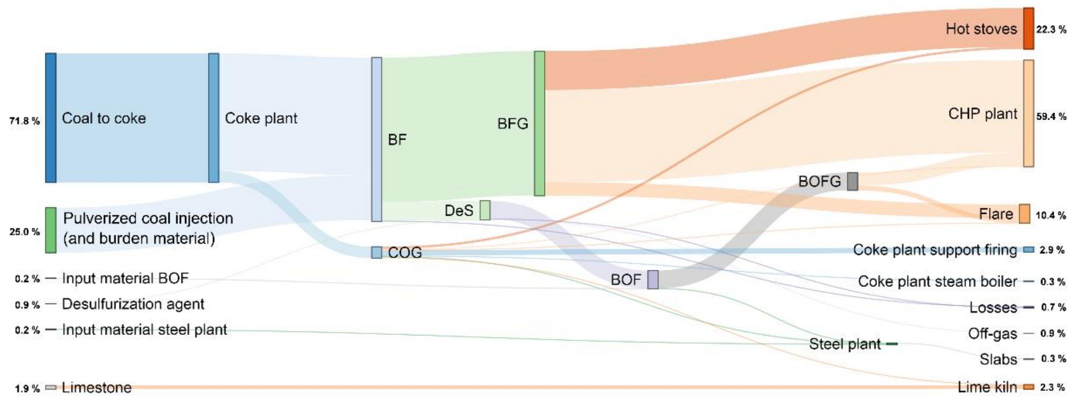


Figure 2: Carbon balance of the Luleå steel mill, as assessed with the iron and steel system model.

Table 1: Gas properties for the considered CO_2 sources at the Luleå steel mill, i.e. in the case without CO_2 capture.

	Unit	Hot Stoves flue gas	BFG	CHP flue gas
CO_2	mol.%	25.1	24.6	29.6
N_2	mol.%	66.4	49.6	64.4
O_2	mol.%	1.0	0.0	0.4
H_2O	mol.%	7.5	2.2	5.6
CO	mol.%	0.0	20.4	0.0
H_2	mol.%	0.0	3.2	0.0
T	°C	269	29	120
p	kPa	105	181.3	105
Flow	kNm^3/h	178.5	352.4	394.7

Table 2: Heat sources for partial capture of CO_2 with suitable heat recovery technology, estimated heat recovery efficiency, and heat amount for the Luleå steel mill under reference conditions, i.e. without carbon capture. Adapted from (Sundqvist et al., 2018).

Source	**Recovery method**	**Recovery efficiency**[1]	**Heat (source)**[2] **(GJ/h)**	**Accum. Heat (level)**[3] **(GJ/h)**	**Heat Level (HL)**[4]
CHP plant (excess heat)	Back-pressure operation	63%	228.1	228.1	1
Gas flaring (excess heat)	Steam boiler	93%	152.8	380.9	2
Hot stove flue gas (excess heat)	Heat recovery boiler	91%	32.9	413.8	3
Hot coke (excess heat)	Dry coke quenching + heat recovery boiler	67%	41.5	455.4	4
Hot slag (excess heat)	Dry slag granulation + moving bed heat exchanger +heat recovery boiler	65%	94.2	549.5	5
additional CHP plant (extra primary energy)	Biomass fired steam boiler + back-pressure steam turbine	85%[5]	419.5	977.7	6

[1] Potential to convert the excess energy into steam.

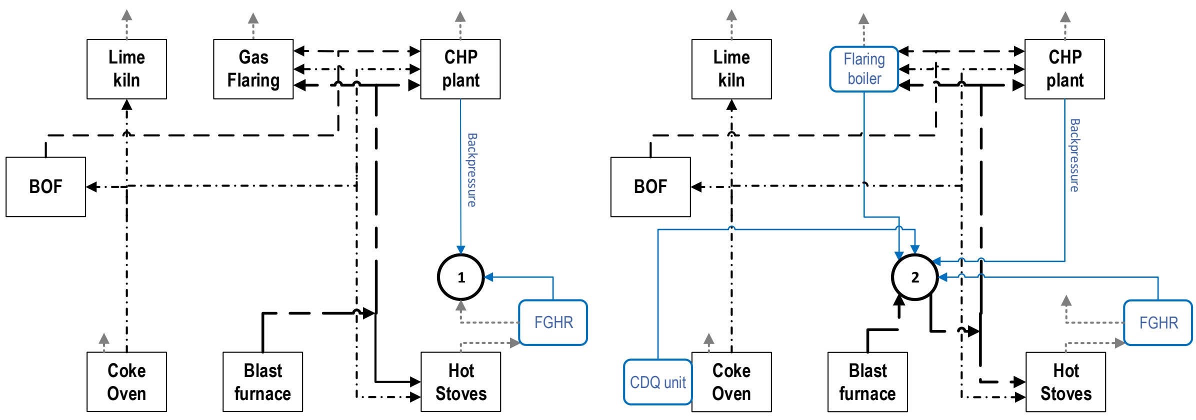
[2] Accessible energy from specific source at the investigated plant site.

[3] Accumulated accessible energy at the given HL at the investigated plant site.

[4] Rating according to level of accessibility (i.e., technology readiness) of the excess energy.

[5] The total efficiency (steam and electricity) is 85% and the electrical efficiency is 22.7%.

The present work considers five capture scenarios S1-S5. Each capture scenario includes one or more of the CO_2 sources listed in Table 1 and one or more of the identified sources of excess heat or *heat levels (HL)* from Table 2. Figure 3 presents an overview of the capture scenarios, showing the integration of the capture units into the steel mill. The considered heat levels that deliver steam to the capture site for each scenario are highlighted in blue. Table 3 summarizes key characteristics of the scenarios. Capture scenarios S1–S3 represent partial capture solely from the hot stove flue gas, BFG, and CHP flue gas, respectively. The heat supply level is based on the available excess heat, which sets the capture rate from the respective CO_2 source. In case sufficient amount of heat is available, the capture rate from a single CO_2 source is set to a limit of 90%, which resembles full capture and an associated minimum investment cost (Biermann et al., 2018) for enabling capture from that source. Scenarios S4 and S5 represent capture from more than one CO_2 source at capture rates of 90%. In S4 and S5, a biomass-fired CHP plant (Bio-CHP) powers the process in addition to the excess heat. The Bio-CHP plant is a back-pressure turbine that generates 3 bar of steam for the reboiler of the capture unit. No extra carbon emissions are allocated to the heat and power production from the Bio-CHP. Scenario S4 includes a capture unit with two absorbers and a common stripper, to avoid blending the BFG and hot stove flue gas. Scenario S5 includes a capture unit for the CHP plant flue gases in addition to the unit described in scenario S4. Thus, scenario S5 captures 90% of the CO_2 from all three sources and represents the full capture case in this work, i.e. similar to what was investigated by Ho et al. (Ho et al., 2013).



S1 S2

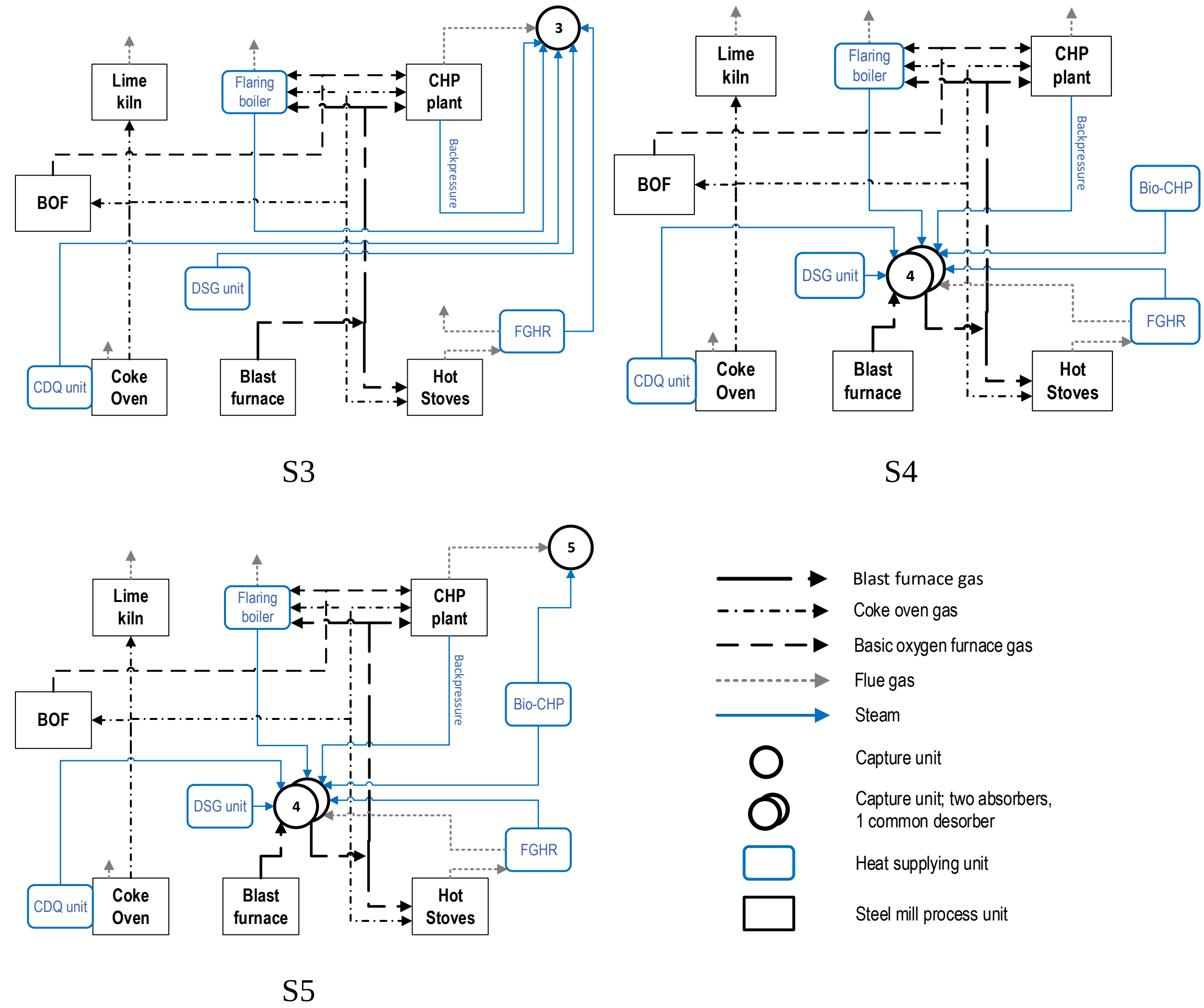


Figure 3: Integration of the heat supplying units (blue) and gas system (black) of the steel mill with the capture unit in scenarios S1–S5. The scenarios consider capture from: S1, hot stove off-gas); S2, blast furnace gas; S3, CHP plant flue gas; S4, hot stoves flue gas plus blast furnace gas; and S5, hot stoves flue gas plus blast furnace gas plus CHP plant flue gas. Circles denote capture units and type of design. Bio-CHP, biomass-fired CHP plant; BOF, basic oxygen furnace; CDQ, coke dry quenching; DSG, dry slag granulation; FGHR, flue gas heat recovery from hot stoves.

Table 3: Characteristics of the studied capture scenarios S1-S5. The capture rate depends on the heat that can be made available, only the highest capture rate investigated is shown for each CO_2 source. HL, heat level, see Table 2; FGHR, flue gas heat recovery; abs, absorber column; str, stripper column.

	S1	S2	S3	S4	S5
CO_2 sources	Hot stoves	BFG	CHP	BFG, Hot stoves	BFG, Hot stoves, CHP
Max. capture rate achieved from source	90%	90%	76.5%	90% each	90% each
Heat sources	FGHR combined with back-pressure	HL1-4	HL1-5	HL6	HL6
Number of capture units/ configuration	1/ 1x abs/1x str	1/ 1x abs/1x str	1/ 1x abs/1x str	1/ 2x abs/1x str	2/ 2x abs/1x str

	and 1x abs/1x str

2.2 Process modeling

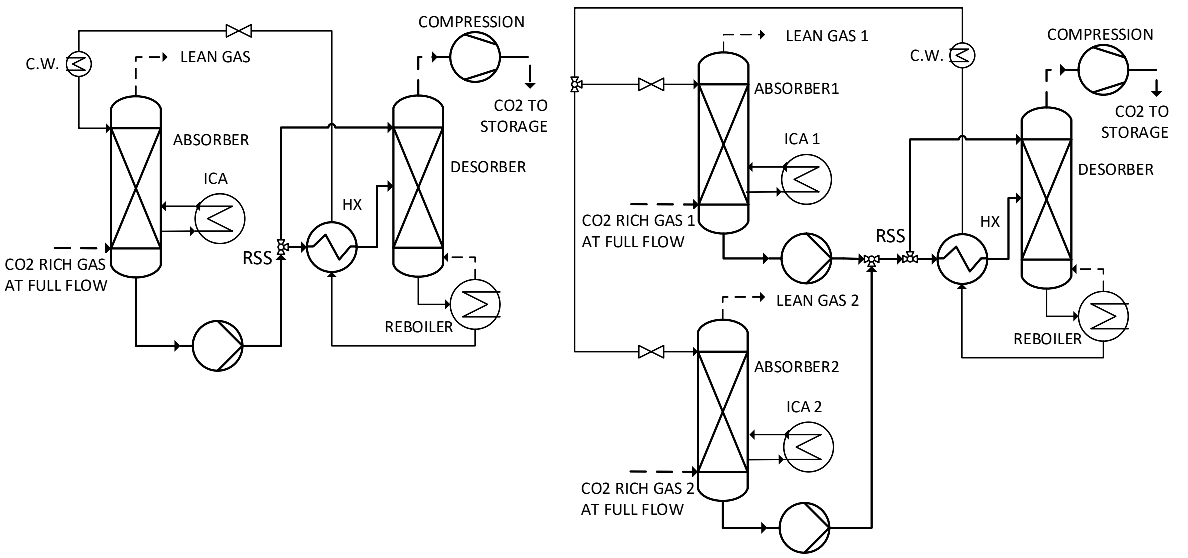
2.2.1 Iron and steel system model

The integrated iron and steel system is modeled using an in-house, 1-D static model composed of inter-linked mass and energy balances over the process units and includes a detailed model of the blast furnace with accompanying hot stove and burden calculation. Each unit operation (see Figure 1) is described by theoretical correlations and empirical relations from industry data, as described in previous works (Hooey et al., 2010; Sundqvist et al., 2018). The model has previously been used, for example, for integrated steel plant optimization modelling (Hooey et al., 2010) or to assess top gas recycling concepts as part of a techno-economic assessment (IEAGHG, 2013). The model requires calibration to an industrial site and, therefore, should be operated close to the calibration points. In the present study, the model is calibrated against data from the SSAB steel mill in Luleå for the reference year 2006.

2.2.2 CO_2 capture model

The capture process is assessed using an Aspen Plus model of a CO_2 absorption cycle with a 30 wt.% aqueous MEA solvent, based on the work by Garðarsdóttir et al. (Garðarsdóttir, 2017). Compared to other capture technologies, amine absorption is already commercially available (IChemE Energy Centre, 2018) and suitable for retrofitting (Voldsund et al., 2019). Both these aspects are important to a near-term realization of a partial capture, which is the focus of this paper. The choice of MEA as the amine solvent is based on it being a well-understood benchmark solvent. The likelihood of commercial or advanced solvents economically outperforming MEA adds a conservative perspective to costing results in this work. The model uses rate-based mass transfer correlations and kinetics for MEA reactions. The absorption cycle is designed for partial capture, which means that depending on the gas flow and CO_2 concentration, the removal of CO_2 from the feed gas will be a function of the available heat (given as a boundary condition, derived from the integrated iron and steel system model). The absorption cycle is optimized to maximize the capture rate by varying the liquid-to-gas ratio (L/G) through manipulation of the solvent circulation rate. For partial capture from CO_2-rich gases, it has been shown that it is more beneficial, in terms of specific reboiler heat demand and therefore possibly costs, to pass the entire process stream through the absorber rather than allow a split-flow of the gas to enter the absorber (Biermann et al., 2018; Øi et al., 2017).

Two process configurations, illustrated in Figure 4, are used in this work. A single absorber configuration is applied in capture scenarios S1–S3. Due to the proximity of the blast furnace and hot stoves, a double-absorber/common-stripper configuration is used for scenarios S4 and S5. Having an absorber for each gas avoids blending the BFG with the flue gas, which is not desired because the BFG is used as heating gas and a dilution to an even lower heating value is unpractical. A common stripper requires a lower level of investment. Both process configurations use intercooled absorbers (ICA) to enhance absorption, as well as a rich-solvent split (RSS) to augment stripper efficiency, as this reduces the specific reboiler heat demand, and, thus, can lead to lower capture cost (Biermann et al., 2018; Garđarsdóttir et al., 2015; Le Moullec et al., 2014). The modeling setup encompassing rich-split, ICA, and the absorption cycle, together with its key design parameters is described in previous work by the authors (Sundqvist et al., 2018).



a) b)

Figure 4: MEA absorption cycle configurations used for partial capture; a) Single absorber configuration. b) Double-absorber/common-stripper configuration;

2.3 Cost estimations

Cost estimations are performed with the Enhanced Detailed Factor (EDF) method (Ali et al., 2019) and are used to discuss the design of the partial capture system for retro-fitting to the Luleå steel mill with the boundary of the cost estimation as shown previously in Figure 1. The costs are aggregated on two levels:

1) the capture plant cost, i.e., the CAPEX of the capture plant including piping from the CO_2 source and all the OPEX related to the capture plant (maintenance, labor, utilities etc.), excluding the steam cost; and
2) the cost of steam, i.e., the CAPEX for piping system required for the steam supply and for the heat recovery equipment, as well as the OPEX related to the equipment and, in particular, any possible changes in power revenue due to excess heat recovery and additional energy supply.

Finally, both the capture plant cost and steam cost are aggregated into an equivalent annualized capture cost (EAC), given in € per captured tonne of CO_2 according to Eq. (1). The consideration of integration cost (piping) and steam supply cost is in line with recent developments in costing (van der Spek et al., 2019).

$$c_{\text{capture,EAC}} = \frac{(CAPEX + OPEX)_{\text{capture plant}} + m_{\text{steam}} \cdot c_{\text{steam,average}}}{m_{\text{CO}_2\text{,captured}}} \tag{1}$$

The cost estimation is made for high technology maturity and reflects the so-called "*n*th-of-a-kind" (NOAK) approach. Using the Aspen In-Plant Cost Estimator, the investment cost for each piece of equipment is estimated and multiplied by an individual installation factor that represents equipment type and size. These installation factors are retrieved from an in-house industry cost database available in the EDF-tool (Ali et al., 2019; Biermann et al., 2018; van der Spek et al., 2017). It is further assumed that all the equipment, except for major vessels such as tanks and columns, is placed in non-insulated buildings. Not included are the cost for purchase of land and piling and the costs for secondary buildings. This method of CAPEX estimation normally constitutes an uncertainty of ± 40% (80% confidence interval). Some of the equipment for heat supply could not be estimated by the individual installation factor method, so cost information from both the academic and grey literature have been used instead, as described in the Appendix in the section on steam cost A.1.2.

Table 4 summarizes the assumptions made regarding the cost estimations. The operational hours represent an annual availability of 95% for the capture plant and heat recovery equipment, which is motivated by high levels of availability of the blast furnace, hot stoves, and CHP plant. The electricity price is oriented towards the Nordic spot-price market (Nord Pool AS), which in the period 2013–2016 had an average electricity price of 29 €/MWh. Electricity required/produced by process units is first balanced within the investigated system shown in Figure 1 before there is purchasing from or selling to the grid. It is assumed that the personnel members operate both the capture plant and the heat supply equipment. The currency throughout this study is $€_{2015}$; external input is converted to $€_{2015}$ using Eurostat's consumer price index (Eurostat, 2018) and historical currency exchange rates.

The cost of steam, c_{steam}, for each recovery technology is determined by a bottom-up approach according to Eq.(2) and includes:

- CAPEX for the equipment that converts heat into steam and piping for delivering the steam to the capture site or to connect to the existing network;
- OPEX including the costs for electricity, cooling water, and maintenance, as obtained from mass and energy balances in Aspen Hysys;
- Revenue loss from electricity sales linked to steam supply from the steel mill CHP plant;
- Revenue gain from electricity sales linked to the additional biomass-fired CHP.

$$c_{\text{steam}} = \frac{(P_{\text{loss,CHP}} - P_{\text{gain,BioCHP}}) * c_{\text{power}} + CAPEX + OPEX}{m_{\text{steam}}} \quad (2)$$

Details of the assumptions made regarding the equipment included to calculate c_{steam} for each heat level are described in Appendix A.1 in Section A.1.2. Appendix A.1 also decribes the equipment included in the capture plant cost (A.1.1).

In order to investigate the conditions for economic viability of the capture scenarios studied, we calculate the net abatement cost, which is the full-chain CCS cost (capture, transport and storage) related to a carbon price, as calculated in Eq. (3). The net abatement cost represents the remaining cost for the plant owner after receiving credit for the captured carbon, either by capitalizing on not having to buy allowances, or by selling off free allocated allowances on the market. The transport and storage cost, denoted $c_{\text{t\&s}}$ in Eq. (3), represent ship transport from the Bothnian Bay to a storage site in the Baltic Sea, and lie within 17 – 27 €/t CO_2 depending on scale (Kjärstad et al., 2016). It should be noted that CO_2 storage in the Baltic Sea may not be considered as mature. However, storage in the North Sea may be considered as mature (Gassnova SF, 2019) and the cost estimation by Kjärstad et al. shows that the cost is similar for both options as transport cost only increase slightly with distance, as long as ship-transport is considered (Kjärstad et al., 2016). Three carbon price projections are examined, denoted c_{carbon} in Eq. (3), as described in Appendix A.1.3.

$$c_{NAC} = c_{\text{capture,EAC}} + c_{\text{t\&s}} - c_{\text{carbon}} \qquad [€/t_{CO_2}] \quad (3)$$

Table 4: Economic parameters assumed in this study

Cost year	-	Year 2015

Plant life time	Years	25
Construction	Years	2
Rate of return	%	7.5
Maintenance	% inst.cost/annum	4.0
Plant availability	h/annum	8,322
Electricity	€/kWh	0.030
Cooling	€/m^3	0.022
MEA	€/m^3	1,867
Sludge disposal	€/m^3	333.3
Biomass price	€/kWh	0.016
Labor		
One engineer	k€/annum	158
Six operators	k€/annum	111

3 Results

3.1 Technical capture performance

This section gives a brief overview of the technical performances of the capture units in the investigated scenarios. Figure 5 shows that the heat requirement for solvent regeneration is dependent upon the CO_2 source and achieved capture rate. A general increase in specific heat demand at a higher rate of CO_2 removal (lower partial pressure of CO_2 in the gas leaving the absorber) is evident. Using MEA absorption, the benefits in terms of heat demand of partial capture are limited to a saving of up to 10% in required heat per tonne of CO_2 captured. Of the three CO_2 sources examined, BFG shows the lowest specific heat demand due to its higher pressure, which results in improved CO_2 absorption. Capture from the flue gases of the hot stoves shows a slightly higher heat demand than capture from CHP plant flue gas, which is due to lower concentrations of CO_2 in the hot stove flue gas.

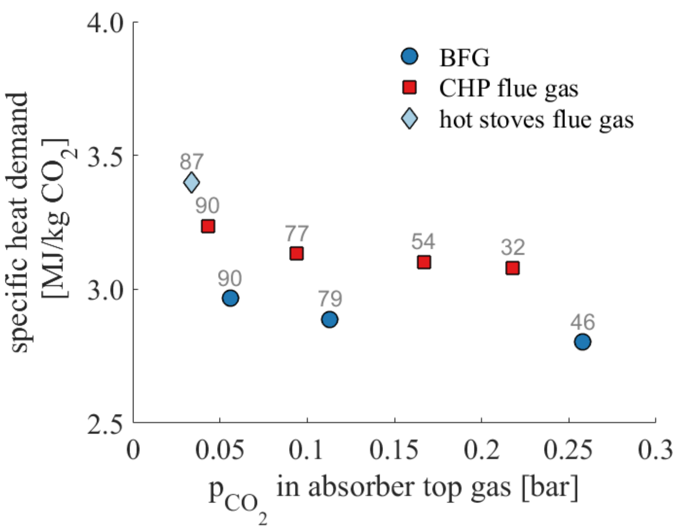


Figure 5: Heat requirement for CO_2 separation from BFG, CHP and hot stove flue gas plotted against partial CO_2 pressure in the absorber overhead gas. The numbers in grey show the achieved separation rate of CO_2 in the absorber in %; Note that ordinate does not start from zero.

The performance of the system is shown in Table 5 for the five capture scenarios S1–S5 – each at their maximum heat recovery level. The three CO_2 sources considered represent almost 85% of the total site emissions, and full capture from all three sources (S5) yields a total site emission reduction of 76.3%. Full capture from hot stoves alone can mitigate about half as much as full capture from BFG. Utilizing all the retrievable excess heat allows for partial capture of 76 % of the CO_2 in the CHP plant flue gases, which corresponds to about 51% of the total site emissions. The total energy input to the system increases, as compared to the reference without capture, and the system becomes a net importer of electricity from the grid at capture rates >20–22 %. The increased electricity demand is predominantly due to the demand for power for CO_2 compression and the need to compensate for the loss of electricity production due to back-pressure operation. It is noteworthy that capturing from BFG (S2) increases the heating value of the BFG and allows for a process gas re-allocation, i.e. greater usage of BFG in the hot stoves and coke oven gas in the CHP (Sundqvist et al., 2018), unlocking a potential of 2–3 MW of excess heat that can be used for carbon capture compared to the steel mill with no capture. This re-allocation of process gases decreases the energy demand and the system becomes more energy-efficient than the reference case without capture, albeit at the expense of power generation. The net power output improves in S4 and turns positive in S5 with additional fuel input in the form of biomass being supplied to the system.

Table 5: System performance in terms of reduced emissions reduction, power generation, and total energy input for each capture scenario (S1–S5), with the highest level of supplied heat (HL) tested. Ref, No capture; S1, hot stoves; S2, BFG; S3, CHP; S4, BFG + hot stoves; S5, BFG + hot stoves + CHP.

	unit	Ref	S1	S2	S3	S4	S5
Heat level (highest tested)		-	HL1m	HL4	HL5	HL6	HL6
Total site reduction	% CO_2	0	19.0	38.8	43.2	51.0	76.3
Specific heat demand	MJ/kg CO_2	0	3.40	2.90	3.12	3.04	3.15
Heat supplied to reboiler	GJ/h	0	262	457	549	629	978
Additional biomass input	GJ/h	0	0	0	0	113	674
Net power output	GJ/h	30	4	-30	-36	-25	62
Total energy input	TJ/h	6.26	6.26	6.17	6.29	6.28	6.88

3.2 Economic efficacy

First, the CAPEX and the cost of steam are presented separately. Thereafter, the total annualized cost for the Luleå plant case is discussed. The total annualized cost is then analyzed for sensitivity towards selected cost parameters.

3.2.1 Investment cost of the capture plant

The installed cost for a capture plant increases with the amount of CO_2 captured and, thus, the capture rate. However, due to economy of scale, the specific CAPEX for each tonne of CO_2 captured decreases with scale for the captured CO_2. Figure 6 shows the magnitudes of these effects on scenarios S1 HL1, S3 HL2 and S2 HL2. The cost break-down highlights the compressor, cross heat exchanger, reboiler, and gas piping as the most expensive items of equipment. The relative proportions of the cost categories vary with scale, CO_2 source and plant design. For instance, the cost of the compressor is merely a function of scale, the gas piping depends highly on the CO_2 source, and the separation columns obviously account for a larger share of the cost in the cases designed to include two absorbers and one stripper. A more detailed break-down of installation cost per equipment type is appended in section A.2, Table A.4, in which a partial capture scenario (S2 HL3) is compared with the full capture scenario (S5 HL6).

Capture from BFG (S2 HL2) requires an investment that is lower by ca. 3 €/tonne CO_2 than capture from CHP plant flue gases (S3 HL2). The slightly higher pressure of the BFG allows for smaller diameters of the columns and piping compared with capture from CHP or HS flue gases and this yields a lower CAPEX. Capture from the hot stoves (S4 HL6) or the CHP (S5 HL6) in combination with capture from the BFG is relatively inefficient, as BFG is the main fuel feed to the hot stoves and the CHP. The concentration of CO_2 drops from 25% and 30% to 17% in the hot stoves and CHP flue gas, respectively, when 90% of the CO_2 in the BFG is captured. The lower inlet concentration increases solvent circulation and decreases CO_2 loading, causing the equipment to be less cost-effective per tonne of CO_2.

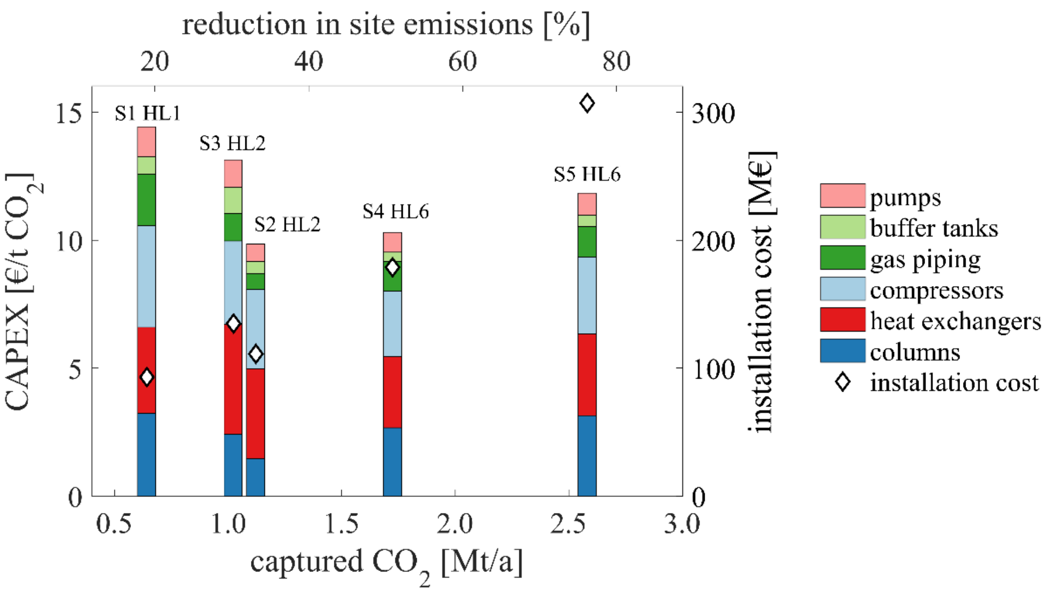


Figure 6: Installation cost (diamond) and specific CAPEX (bars with cost categories) of the CO_2 capture plant versus captured CO_2 for selected capture scenarios

3.2.2 Cost of steam supply

Figure 7 shows the factors governing the cost of steam calculated according to Eq.(2). The cost is primarily determined by the type of heat-recovery technology used (cf. Table 2), the distance to the capture site, and the amount of retrievable steam. A substantial amount of steam, 220–228 GJ/h on average, may be obtained by operating in back-pressure mode for the entire operational year at a cost <2 € per tonne of steam. The cost is dominated by the loss in power revenues. The recovery of steam from flare gases generates a cost of 7 (±2) €/tonne steam, mainly due to the cost of the piping required to lead the flare gases to the additional steam boiler. Heat recovery from hot stove flue gases supplies relatively low levels of steam (~32 GJ/h), although at a low cost of 2–4 €/tonne. The distinct difference in steam cost for FGHR between capture from BFG (S2) and CHP flue gas (S3) is attributable to the longer piping distance in the CHP scenario. Using coke dry quenching (CDQ) to generate low-pressure steam comes at a relatively high costs of 45–55 €/tonne due to the large investment required. Here, the BFG scenario (S2) is more expensive because the steam production is matched to the capture rate cap of 90%, whereas more steam is recovered from excess heat in the CHP flue gas scenario (S3), which captures 64% of the CO_2 at a similar capital expense. Dry slag granulation (DSG) has a comparatively low cost for steam, ca. 5 €/tonne, and a higher capacity than CDQ. However, the cost for DSG is uncertain, as it is not a commercial technology. Additional primary energy supply in the form of a biomass-fired CHP plant can generate steam at a cost of 28 (±5.1) €/tonne and 18 (±2.7) €/tonne for S4 and S5, respectively. The difference in cost is due to economy of scale. In both scenarios, the costs are dominated by the cost of fuel, although the produced electricity helps to reduce the steam cost by 5–6 €/tonne. This also implies that an investment that is solely motivated by power revenues does not pay off. The electricity price would have to be at least 102 €/MWh and 138 €/MWh for S5 and S4, respectively, for the investment to break even.

Figure 8 shows the average steam costs for the successive deployment of the discussed heat recovery technologies, with excess heat recovery being deployed before additional combustion. The increments in steam cost represent the deployment of the next heat-supplying technology with costs (CAPEX and OPEX) at the respective scale of heat supply (in MW). The average steam cost increases from 1 (±0.05) €/tonne for utilizing only the heat available as back-pressure from the existing steam cycle to 12 (±2) €/tonne for full capture powered by the installation of an additional steam cycle (Bio-CHP). Note that if all the steam were to be generated through a biomass-fired steam boiler the cost of steam would be around 14–30

€/tonne. The average cost of steam is similar for the three CO_2 sources in S1–S3, with the differences mainly seen for back-pressure operation and gas flaring. The cost of supplying steam for BFG capture (S2) is higher because the loss of power-related revenue is greater and increases beyond the first heat recovery level (back-pressure). The more heat is retrieved, the more CO_2 can be captured and the BFG is upgraded in terms of its heating value, allowing for extended use of BFG in other steel mill units at the expense of electricity generation in the CHP plant (cf. (Sundqvist et al., 2018)).

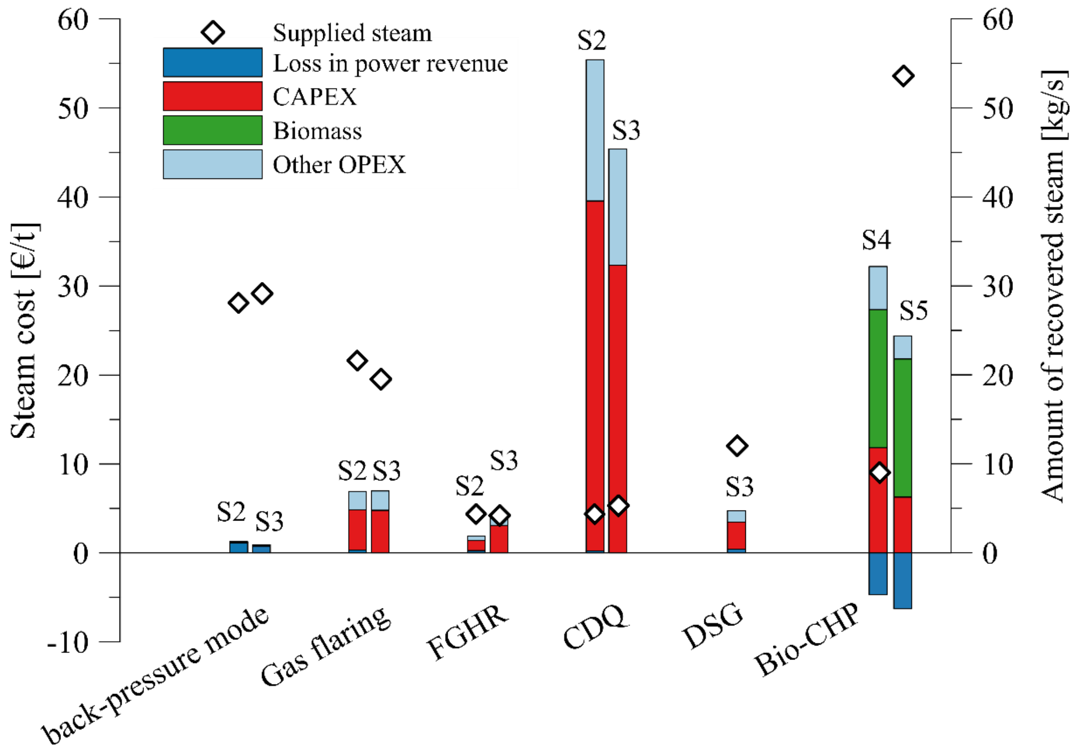


Figure 7: The costs of steam recovered in capture scenarios S2 and S3 via CHP back-pressure operation, gas flaring, flue gas heat recovery (FGHR), coke dry quenching(CDQ), and dry slag granulation (DSG), as compared to the costs of steam produced in additional biomass-fired CHP (Bio-CHP) in capture scenarios S4 and S5.

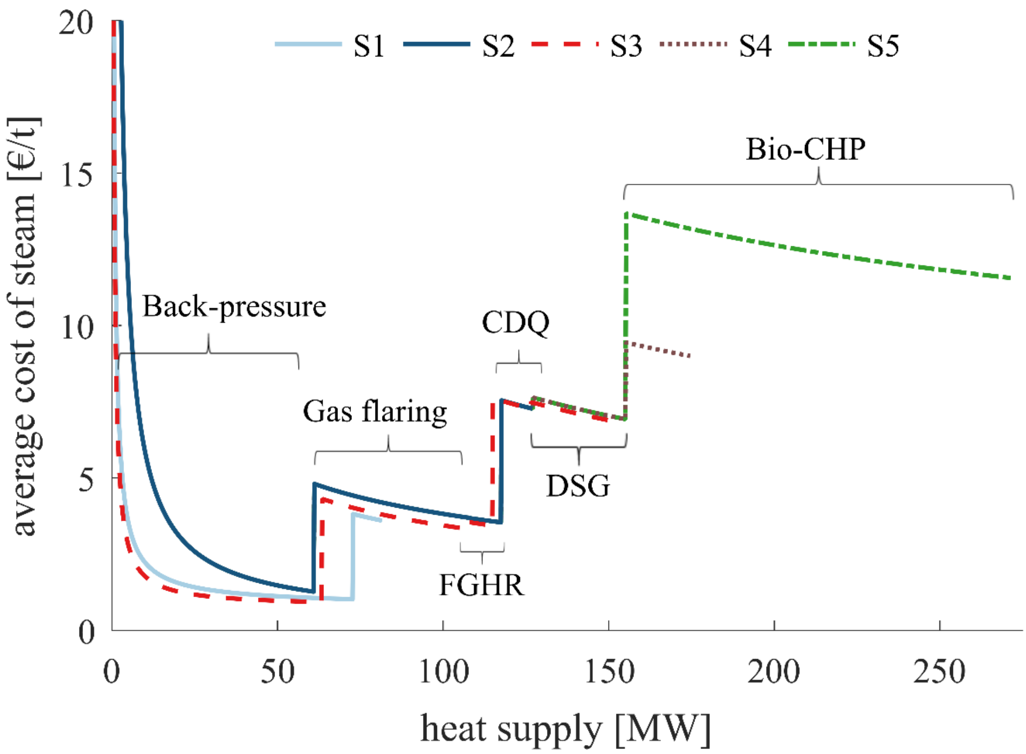


Figure 8: Average costs of steam for capture scenarios S1–S5 in relation to the amount of steam available for capture: FGHR, flue gas heat recovery; CDQ, coke dry quenching; DSG, dry slag granulation; Bio-CHP, biomass-fired CHP plant. The parenthesis in the figure represent the recovery technology being implemented successively with increasing steam amount.

3.2.3 Equivalent annualized capture cost

The equivalent annualized capture cost (EAC) is aggregated from the capture plant cost and steam cost according to Eq. (1). The annualized absolute cost including CAPEX and OPEX are in the range of 20.6 (±4.1) M€ to 111.9 (±14.8) M€ for the smallest and largest annual capture capacities of 0.64 Mt CO_2/annum and 2.58 Mt CO_2/annum, respectively. Figure 9 demonstrates that the capture costs for the studied scenarios vary within the range of 28–50 €/tonne CO_2-captured depending on the amount of CO_2 captured. A range of low-capture costs is observed for 0.7–1.2 Mt CO_2/annum, corresponding to a 19–36% reduction in site emissions, after which the capture cost increases with capture rate as more expensive heat recovery equipment is installed. The lowest capture cost of 28 (±4) €/tonne CO_2-captured is observed in scenario S2 HL3, i.e., capture from BFG with heat supplied from back-pressure operation, gas flaring, and flue gas heat recovery (FGHR), achieving a 36% (ca. 1.2 Mt CO_2/annum) reduction in site emissions. The full capture scenario S5 HL6, i.e., 90% capture from BFG, hot stoves, and CHP plant flue gases, shows a rather high cost of 43 (±6) €/tonne CO_2-captured, although it achieves a reduction in site emissions of 76% (ca. 2.6

Mt CO_2/annum). Furthermore, it is clear that capture from BFG is more economic by 3 € or 5 € per tonne CO_2-captured (on average) compared to capture from hot stove or CHP flue gases, respectively, which is within the margin of uncertainty for the cost estimation.

Figure 10 shows the cost breakdowns for the most cost-effective BFG capture scenario (S2 HL3) and the full capture scenario S5, which have annual costs of 33.6 (±5.1) M€ and 111.9 (±14.8) M€, respectively. In the partial capture scenario, CAPEX makes up one-third of the cost, followed by fixed OPEX (maintenance and labor), and the cost of steam recovered from excess heat. In the full capture scenario, steam generation from both excess heat and additional fuel input is the dominating cost with a share of 39%, followed by CAPEX at 27%.

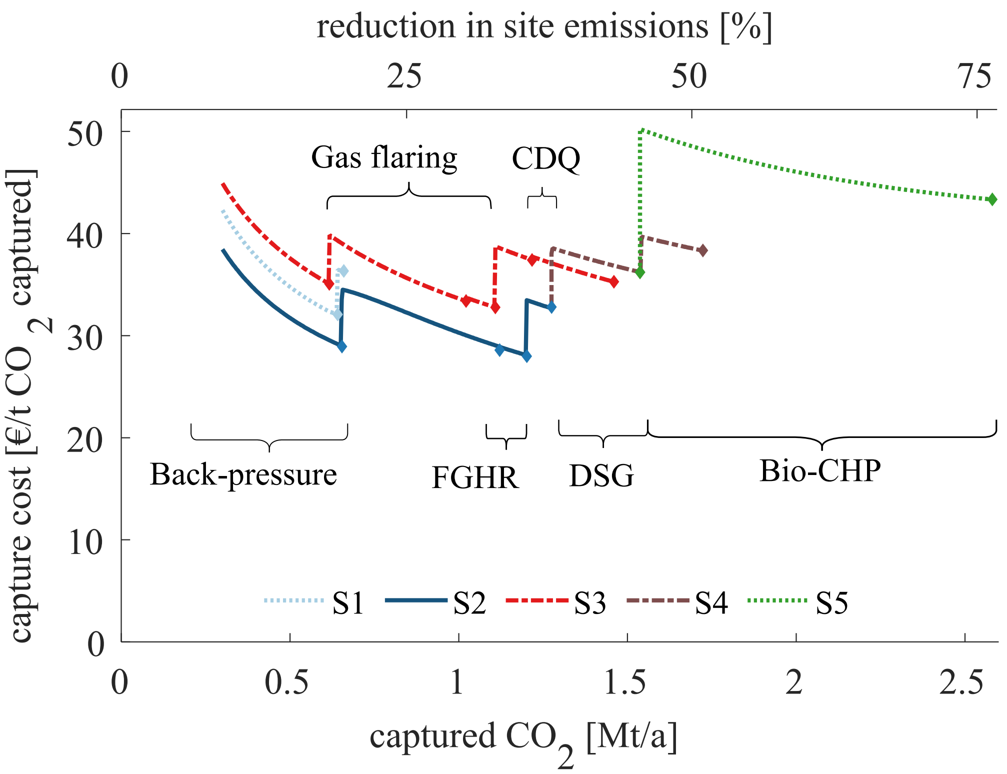


Figure 9: Capture costs for scenarios S1–S5 depending on annually captured CO_2
The parentheses and diamonds indicate the successive deployment of heat recovery technologies; FGHR, flue gas heat recovery; CDQ, coke dry quenching; DSG, dry slag granulation; Bio-CHP, biomass-fired CHP plant.

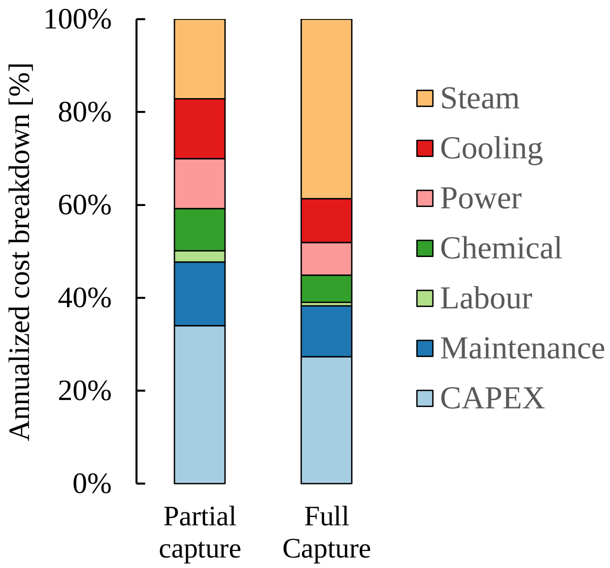


Figure 10: Comparison of the annualized cost breakdowns of the partial capture scenario (S2 HL3) and full capture scenario (S5 HL6). CAPEX represents the capital expenditures for the CO_2 capture plant.

3.2.4 Sensitivity analysis

The influences of underlying cost parameters (cf. Table 4) on annualized cost are illustrated in Figure 11 for the partial capture scenario S2 HL3 and the full capture scenario S5 HL6. The listed parameters are altered by ±50% one at a time. The figure reveals that operational hours, lifetime of the plant, rate of return and external energy (electricity and biomass) are the factors most sensitive to change. Maintenance rate, cooling water supply, and the assumed length of the gas and steam piping influence the cost by <9%. Overall, the partial capture scenario demonstrates a higher sensitivity than the full capture scenario, as its annual cost is more dependent upon the investment (cf. Figure 10). The exception to this is the cost for external energy, which is more sensitive in the full capture scenario because it relies not only on power imports but also on biomass supply. The electricity price and biomass price are treated as coupled parameters, which is likely to be the case for future electricity systems that rely on renewables with a significant share of biomass (Johansson et al., 2019). Figure 12 shows the net abatement cost, i.e., the full-chain cost for CCS (capture, transport and storage) minus the carbon price, for various carbon and electricity prices over a larger range, and couples the biomass price to the electricity price at a constant ratio for the full capture scenario. In all cases, partial capture is more cost-efficient and less-sensitive to variations in the price of the external energy supply. In general, carbon prices of around 50–60 €/tonne CO_2 and 50–80 €/tonne CO_2

are required for the net abatement cost to become negative for the partial capture scenario and full capture scenario, respectively.

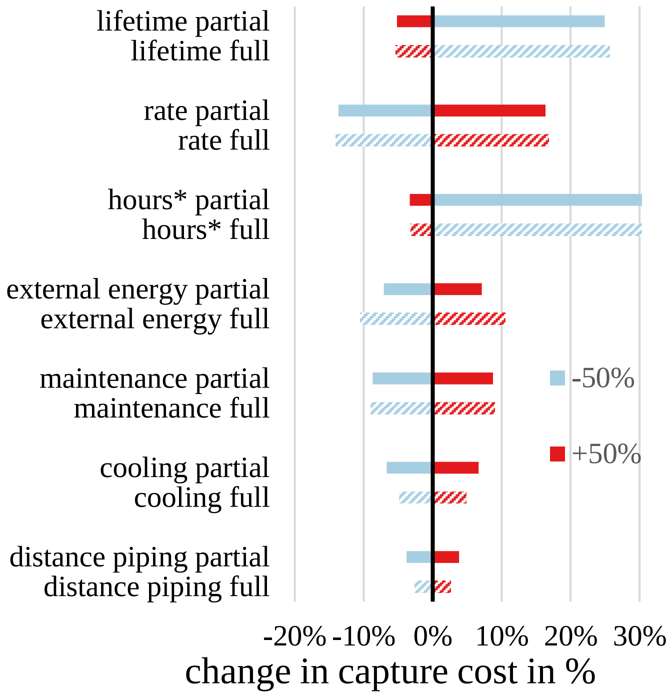


Figure 11: Sensitivity of the annualized capture cost with respect to the main cost parameters for a partial capture scenario (S2 HL3, full bar, base value 28 €/tonne CO_2) and a full capture scenario (S5 HL6, striped bar, base value 43 €/tonne CO_2). * Increase in hours limited to 100% annual operation, the decrease in hours not shown fully due to scale: cost increase by 67% and 64% for partial and full capture scenario, respectively.

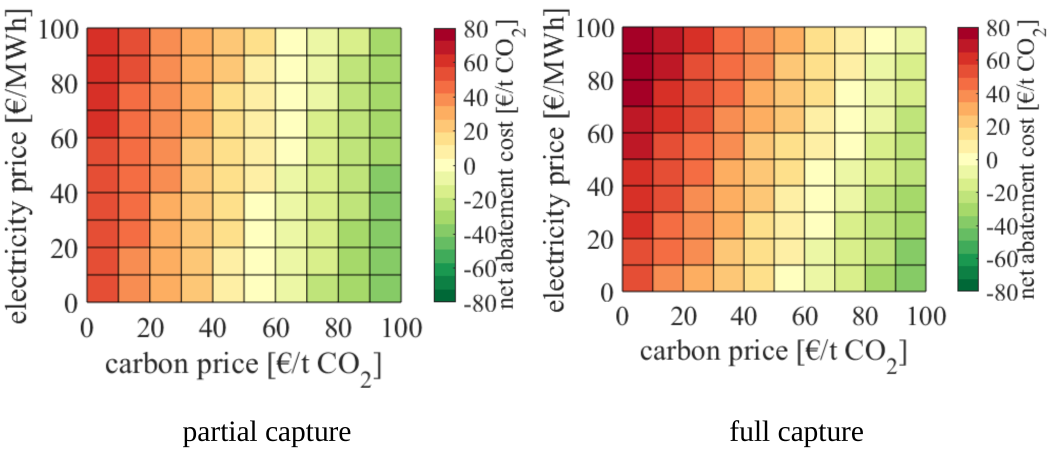


Figure 12: Sensitivity of the net abatement cost towards the electricity price and carbon price for partial capture (S2 HL3) and full capture (S5 HL6).

3.2.5 Time perspective on the abatement cost

Figure 13 shows the net abatement cost trajectories for partial capture from BFG for the period 2018–2040, based on three carbon-pricing projections. CO_2 prices for advanced economies in line with IEA's sustainable development scenario (WEO 2 °C) would make partial capture at the Luleå steel mill economically viable in Year 2025. Less ambitious policy-driven carbon pricing in the early 2020s will postpone this to Year 2029 (WEO&NEPP). Following the price projection for the EU ETS by Refinitiv (Qin, 2018), a company providing financial market data, the market does not foresee negative net abatement cost in either the 2020s or in the 2030s when extrapolating the data to the 2030s (see Appendix Table A.3). It should be noted that the applied EU ETS projection does not foresee the carbon price levels necessary to meet the sustainable 2°C target (WEO).

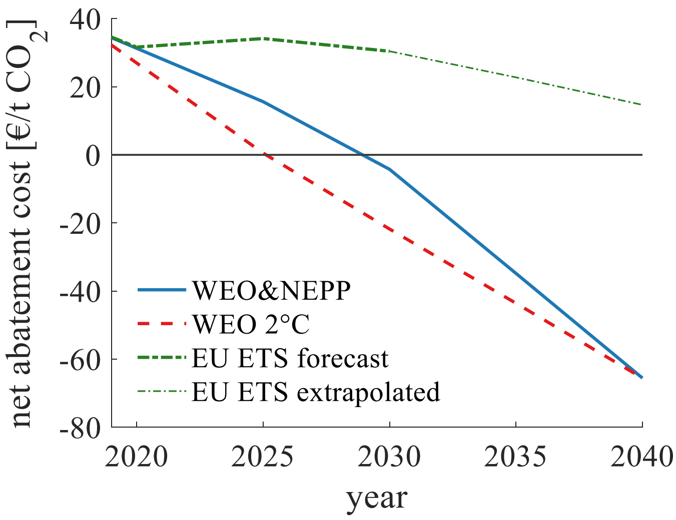


Figure 13: Net abatement costs for the steel industry based on partial capture of CO_2 from BFG (S2 HL3) with excess heat from back-pressure operation, flue gas heat recovery, flare gases, and three carbon price projections: sustainable development projection (WEO 2°C), moderate development projection (WEO &NEPP), and a carbon-market projection (EU ETS forecast). The carbon price for the EU ETS has been extrapolated for the period 2030–2040.

4 Discussion

This section is divided into three parts. First, the excess heat sources used for partial capture and their limitations are discussed. Second, the full capture benchmark is compared to the data in the literature and its external heat supply is debated. Third, near-term implementation of partial capture in the iron and steel industry is explored.

4.1 Limitations on excess heat recovery for partial capture

The above given techno-economic assessment has found that partial capture with excess heat can be more economic than full capture, provided that low-cost and mature heat recovery technology is implementable. Such technologies include back-pressure operation and flue gas heat recovery, either of which can use the existing infrastructure or relatively low-cost heat recovery units. Flare gas utilization provides steam rather intermittently, and an extra buffer tank may be required to allow continuous heat production, which was not taken into account in the equipment cost. The increase in process complexity is reflected in a higher steam cost from CDQ, though less so for DSG, due to uncertainties in how the costs will turn out once commercialization is achieved.

In all, the excess heat from back-pressure operation and flue gas heat recovery will likely be deployed first, followed by the installation of a new boiler fired by flare gases and additional fuel, e.g., biomass or other. Since steam from CDQ is found to be more expensive than additional combustion (cf. Figure 7), investment in CDQ cannot be motivated based on steam production alone. It should be noted that the steam cost in the present study does not represent secondary effects, such as efficiency gain by capturing from BFG (reduced fuel consumption in the steel mill) or improved quality of the slag due to DSG or avoidance of water pollution and reduction of water consumption due to CDQ. Note that carbon capture and the required heat recovery units are operated continuously at constant load. Martinez Castilla et al. (Martinez Castilla et al., 2019) performed a dynamic modeling study of capture unit operation with seasonal and hourly variations and they found that typical variations are manageable through the implementation of an appropriate capture unit design and control scheme, and that a capture performance close to constant load can be achieved.

4.2 The full capture benchmark and comparison with the literature

The comparability of the cost results within the literature is often low due to the high variability of applied methods and scopes. From a literature review on capture cost from the steel industry applying 30 wt.% aqueous MEA solvents, a cost range for capture from BFG was found to be

54–72 €/tonne CO_2, which is comparable with and even lower than the cost for end-of-pipe capture, which is around 60–100 €/tonne CO_2 (see Table A.5 in the Appendix for a list of cost data from the literature reviewed). The techno-economic assessment carried out in the present study confirms that carbon capture from BFG is more cost-effective than end-of-pipe capture from hot stoves or the CHP plant onsite. Compared to the literature, this study concludes that there is a lower cost for full capture, i.e., separating 90% of the CO_2 from BFG, hot stove and CHP plant flue gases, at 43 (±6) €/tonne CO_2 (cf. Figure 9). The reason for this is the use of excess heat to cover 57% of the heat supply. The supply of heat exclusively from natural gas or coal at a price of 20–22 €/tonne steam (Ali et al., 2018) would entail a cost of 56–58 €/tonne CO_2-captured, which is at the lower end of the cost range reported in the literature. Yet, such fossil fuel-based heat supply would increase CO_2 emissions, which would also have to be taken into account.

The use of low-grade biomass to provide the remaining 43 % of the required heat for full capture that is not supplied by excess heat, would require roughly 300,000 tonnes (dry) of biomass per year, which is at the scale of the world's largest biomass pelletization plants currently in operation (Kuparinen et al., 2014), so this might pose challenges in terms of production and supply of CO_2-neutral biomass. Furthermore, the use of biomass to generate heat for CCS and some electricity may not represent the 'best' option for using a limited resource. Other options even exist in the iron and steel industry for a more-efficient use of biogenic carbon, e.g., as a bio-reductant fed directly to the blast furnace via tuyère injection, thereby replacing pulverized coal injection (Mousa et al., 2016; Wiklund et al., 2017).

4.3 Partial capture and conditions for near-term implementation

In anticipation of the Market Stability Reserve (MSR), the CO_2 price in the EU ETS has increased to >20 €/tonne in 2018 after a period of low prices due to oversupply following the financial crisis in Year 2008. The MSR will remove a large share of superfluous emission certificates in the early 2020s, and thus, will likely maintain CO_2 price levels at >20 €/tonne (Qin, 2018). Importantly, the capture cost found in this study for partial capture in the steel industry is close to the expected carbon price levels in the near future (Qin, 2018), and thereby cover a large share of the entire full-chain cost. The full-chain cost, including ship transport to the storage site in the Baltic Sea minus a carbon price, i.e. the net abatement cost (cf. Eq. (3)), have been analyzed for different carbon price projections (cf. Figure 13). The market-oriented projection, i.e., the current EU ETS system, is unlikely to trigger the implementation of even a low degree of capture before the Year 2030. Given the strict emission limits foreseen for

Europe, partial capture will not be sufficient for the period 2040–2050, and the economic lifetimes of the capture units will be rather short if implemented in the 2030s or later. However, with policies that assign a higher value to carbon (cf. Figure 13), the economic viability of partial capture looks promising over the entire lifetime of ca. 25 years, starting from the 2020s.

Note that the applied transport and storage costs are quite high, as they account only for the CO_2 emissions at a single and rather remote site. Prices closer to 10 €/tonne CO_2 or lower for less-remote sites or sites connected to a transport hub allowing for pipeline transport (Kjärstad et al., 2016) could result in lower full-chain cost, and, thus, an earlier implementation. It should be noted that the net abatement cost uses electricity price estimates that are based on annual averages and do not cover large price variations in the electricity system, which may be expected in future electricity systems with a large share of renewables (Johansson et al., 2019).

Allocating the cost for CO_2 capture, transport, and storage to the steel product (excluding any carbon credit), would lead to an increase in production cost in the range of 20–80 €/t steel (hot rolled coil, HRC) for the investigated scenarios. Relative to an estimated production cost of 466 $€_{2015}$/t HRC (IEAGHG, 2013), partial capture with excess heat (S2 HL3) and full capture (S5 HL6) would cause an increase in production cost of about 6% and 17%, respectively. For context, the U.S. tariffs on steel imports were increased by 25% in 2018, leading to a turmoil on the global steel market with an increase in HRC prices of about 27% in the U.S. and a drop by 11% in Europe within a year (MEPS International Ltd., 2019). Possibly triggered by the more protectionist global trade atmosphere, there have been recent calls for border carbon adjustments (ArcelorMittal, 2018; Mehling et al., 2019), such as a carbon tax for imported goods, which may level competition for domestic manufacturers who face carbon prices and may help incentivize the investment into mitigation technologies, such as CCS. Note that the allocation of CCS cost to the steel product alone is not a priori – costs and reduced CO_2 emissions could be allocated to all products including electricity, district/industrial heating, and minerals (slag). The implications of such allocation schemes on the cost and emission intensity of a product-portfolio depend, amongst others, on the choice of mitigation technology and economic conditions, and is a matter of ongoing research.

In addition to the uncertainties surrounding economic viability, the long investment cycles in the steel industry may be a decisive factor for the timing of implementation of partial capture. For example, the refractory lining of a blast furnace lasts 15–20 years and it is highly likely that the blast furnace will be used for the entire life time of the lining. Thus, investments made on relining in the period 2020–2030 are likely to be continued until a time of strict carbon constraints when alternative carbon-free production technologies (e.g. hydrogen reduction) may be a competitive alternative to the blast furnace route.

In summary, as a mature and low-cost technology, partial capture of CO_2 has a time-window for implementation in the coming 10–15 years (or within one more investment cycle), after which the lifetime of the capture unit will most likely be too short until policies will require close to 100% decarbonization, which will favor other options for CO_2 mitigation from steel manufacturing. However, partial capture could evolve towards full capture over time and achieve low or even near-zero emissions, as required from the power sector to limit warming to 2 °C (Feron et al., 2019), through onsite technology development, such as solvent improvement, additional capture units, and/or in combination with other measures, such as biomass, electrified heating, and energy efficiency (Biermann et al., 2018). Early implementation of partial capture would initiate large-scale emissions reductions and decrease the risk of other technologies failing to arrive on time and at scale to meet reductions targets. This is an important argument in favor of partial capture since it is the accumulated CO_2 emissions which govern if the world will comply with the Paris agreement of staying well below 2 °C. Thus, unless there are full capture or other zero-emission steel making processes made available economically or technically in the near term, partial capture can constitute a first drastic cut of emissions contributing to significantly lower the accumulated emissions.

5 Conclusions

A techno-economic assessment of partial capture in primary steelmaking is conducted at the example of a Swedish steel mill. Excess heat from various sources in the steel mill, quantified in a previous work (Sundqvist et al., 2018), is recovered in the form of low-pressure steam to drive a 30 wt.% amine-based absorption process to separate CO_2 from the off-gases of the steel mill. An established cost estimation method is applied together with literature sources to determine the CAPEX and OPEX for the capture unit, the cost of the required gas and steam piping, and the cost for steam production from excess heat.

This study finds that for the steel industry, partial capture of CO_2 with excess heat is more low-cost in terms of both the absolute and specifics cost per tonne CO_2 than full capture of CO_2. The lowest capture cost of 28 (±4) € per tonne CO_2 is found for capture from blast furnace gas with excess heat from the CHP, hot stove flue gas heat recovery and flare gas utilization. This corresponds to a reduction of 36% in site emissions. The full capture benchmark, i.e., 90% CO_2 separation from three CO_2 sources, achieves a reduction of around 76% at a cost of 43 (±6) €

per tonne CO_2-captured. Full capture relies more on the external energy supply making OPEX the dominating cost factor. Partial capture powered by excess heat is dominated by CAPEX and is less-sensitive to fluctuations in the price of external energy.

Capture from the BFG yields a cost which is 3–5 € per tonne CO_2 lower than end-of-pipe capture from either CHP or hot stoves. This is due to the higher pressure in BFG, which reduces the heat demand and allows for a more cost-efficient design.

The bottom-up method applied in this work finds that the cost of steam from excess heat depends on the quantity involved and the recovery technology utilized. Back-pressure operation, heat recovery from hot stove flue gases, and the utilization of flare gases for steam production are available, and implementable heat supply options, with the steam costing <2 €, 2–4 €, and approximately 7 € per tonne of steam, respectively. Retrieving additional excess heat via coke dry quenching or dry slag granulation becomes more expensive and complex. Instead, further heat supply via combustion of additional fuel is likely to yield a lower cost of steam of around 14–28 €/t.

An analysis relates the full-chain abatement cost for partial capture of CO_2 (capture, transport, storage) to different carbon price projections. Early implementation of partial capture of CO_2 in the 2020s is possible and economically viable, if policymakers enact and enforce long-term and predictable regulation of carbon prices beyond Year 2030. Over the lifetime of the capture plant, carbon prices will have to be in the range of 40–60 €/tonne CO_2 on average to justify the investment from the plant owner's perspective.

Notes

The authors declare no competing financial interest.

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Appendix

A.1 Detailed cost estimation

The following sections, which are an extension to Section 2.3, describe in detail the assumptions made and the calculation of the capture plant cost, steam cost, and net abatement cost.

A.1.1 Capture plant cost

The individual installation factor method described in Section 2.3 is applied to estimate the installation costs for the equipment of the MEA capture plant. Figure A.1 depicts the most relevant items of equipment considered for a single-absorber configuration with gas treatment. The double-absorber/common-stripper configuration (not shown) is identical but includes additional gas treatment, an absorber and washer column, an intercooling arrangement, a rich pump, and a lean cooler. Importantly, the direct contact cooler (DCC) is omitted for the blast furnace gas, since its temperature is about 30 °C (De-SO_x/De-NO_x already in place at the site).

Note that gas piping from the CO_2 source to the capture plant is considered as item of equipment. The cost of piping installation includes basic fittings, valves and insulation and is based on the site-derived distances for the capture scenarios listed in Table A.1, the gas properties and flow in Table 1, an assumed gas velocity of 40 m/s, and the piping material (SS-316L).

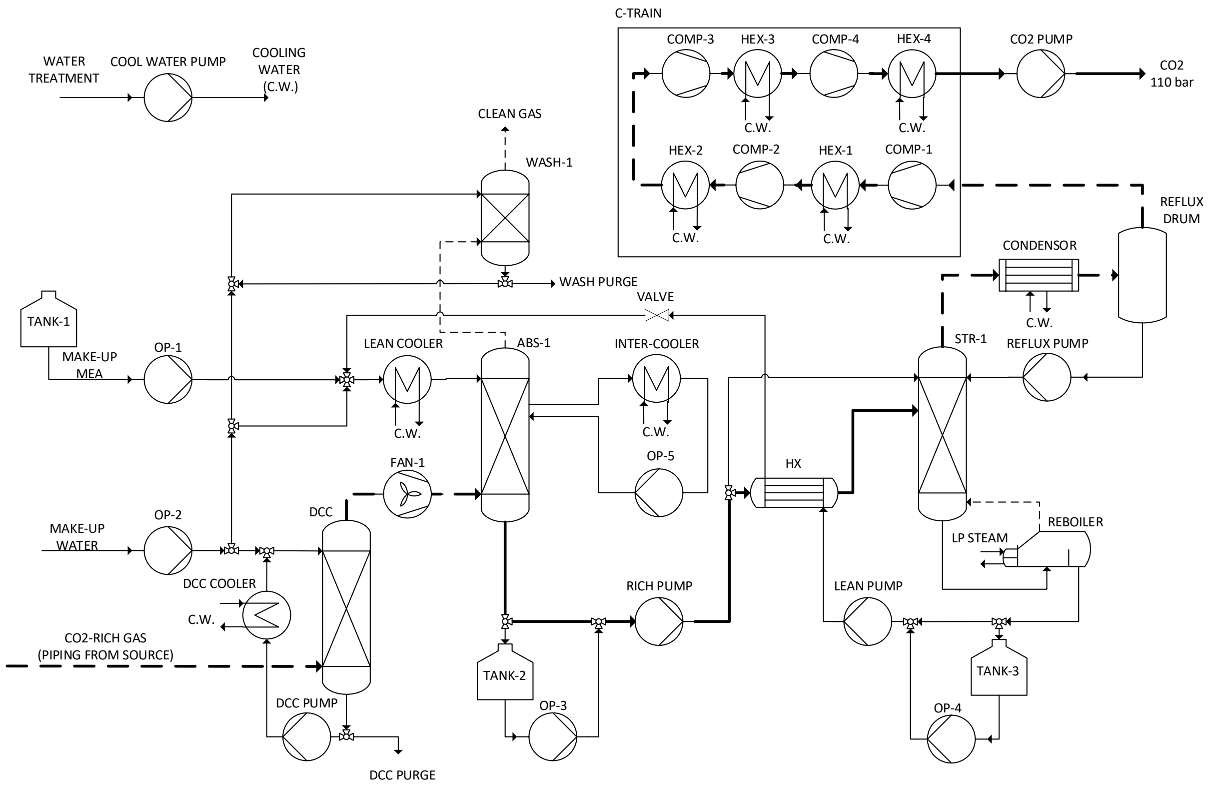

Figure A.1: Major items of equipment included in the installation cost estimation for the capture plant. Shown is an exemplary flowsheet for a single-absorber design with gas piping and gas treatment (DCC) and CO_2 compression to 110 bar.

Table A.1: Lengths of gas piping considered in capture scenarios S1–S5

Capture scenario	S1 HS	S2 BFG	S3 CHP	S4 BFG+HS	S5 BFG+HS+CHP
Length (m)	50	100	75	175	225

A.1.2 Cost parameters for heat recovery equipment

The items of equipment considered at each heat level are listed in Table A.2. Steam from turbine back-pressure operation does not require any recovery equipment. For gas flaring, FGHR, and DSG, the cost methodology for heat recovery networks described previously (Ali et al., 2018) is followed. For gas flaring, additional gas piping is required to connect the flare gases to a new steam boiler site. The cost for CDQ and the additional CHP plant is based on external sources. The scaling factor to obtain adjusted installation costs with the power law is 0.65. For CDQ, the capacity was is to 80 tonnes of coke/h. For DSG, the annual slag production at the site from both the blast furnace and basic oxygen furnace is assumed to be 550,000 tonnes. For the Bio-

CHP, the thermal capacity is set to match the amount of heat required to meet the full capture requirement in scenarios S4 and S5. If more than one heat recovery option is utilized, the steam cost is based on the average cost $c_{\mathrm{steam,average}}$.

Table A.2: Assumptions made regarding the cost parameters for the heat-supplying equipment.

	Heat recovery					Extra energy
Heat source	Back-pressure operation	Gas flaring	FGHR from hot stoves	Coke dry quenching (CDQ)	Dry slag granulation (DSG)	Biomass-fired CHP (Bio-CHP)
First introduced in	HL1	HL2	HL3	HL4	HL5	HL6
Steam piping (m) velocity 30 m/s	50	100	700/50	3000	100	100
Equipment						
Steam boiler	-	✓	✓	n.a.	✓	n.a.
Condenser/cooler	-	✓	✓	n.a.	✓	n.a.
Condensate pump	-	✓	✓	n.a.	✓	n.a.
Condensate tank	-	✓	✓	n.a.	✓	n.a.
Air fan	-	✓	-	n.a.	-	n.a.
Flare gas piping (m)		200				
Special equipment	-	-	-	CDQ plant[1]	DSG plant[2]	CHP plant[3]
Scaling size	-	-	-	100	300	132
Unit				t coke/h	kt slag/yr	MWth
Cost (k€$_{2015}$)	-	-	-	40,250	8,057	80,000
Reference	-	-	-	[4]	[5]	[6]

n.a., Does not apply/considered in special equipment.

[1]CDQ: cooling vessel, recovery boiler, gas circulation system, steam cycle.

[2]DSG: dry granulator, moving bed heat exchanger, blower, off-gas system.

[3]Bio-CHP plant: back-pressure turbine, steam cycle with biomass boiler.

[4] (SSAB EMEA AB, 2012)

[5] (Norgate et al., 2012; U.S. DOE Energy Efficiency & Renewable Energy, 2016)

[6] (Haaker, 2007)

A.1.3 Net abatement cost and carbon price projections

The net abatement cost is calculated (cf. Eq. (3)) for three carbon price projections for the period 2020–2040: 1) a sustainable development scenario in line with the 2°C target (WEO 2°C); 2) an adapted moderate development scenario by NEPP (WEO & NEPP); and 3) a market-oriented EUA forecast (EU ETS forecast). For the same time period, the electricity price projection for Sweden is taken from the latest results of the NEPP project. The underlying price assumptions are listed in Table A.3.

Table A.3: Carbon prices (CO_2) and Swedish electricity price scenarios for the period 2020–2040

Year	Carbon price $€_{2015}$/t CO_2			Electricity price $€_{2015}$/MWh
	WEO & NEPP	WEO 2°C	EU ETS forecast	
2018	17.7	17.7	17.7	41.6
2020	24.1	28.4	23.7	42.4
2025	40.0	55.1	21.5	44.5
2030	60.0	77.5	25.3	45.6
2035	91.2	100.0	33.6[1]	50.5
2040	122.4	122.4	42.2[1]	54.2
source	(IEA, 2018; NEPP, 2019)	(IEA, 2018)	(Qin, 2018)	(Rydén and Unger, 2018)

[1] Extrapolated values from estimated prices for period 2026–2030.

A.2 Break-down of capital expenditures for CO2 capture plants

Table A.4: Capital expenditures in k€$_{2015}$ (thousands) of the capture plants for two scenarios: partial capture from BFG with excess heat (S2 HL3), and full capture from BFG, hot stoves and CHP plant (S5 HL5). The 'ID' corresponds to equipment in Figure A1, '#' stands for quantity of each equipment, 'size' for the aggregated size of an equipment type except for vessels, where '/' denotes the ratio between height and diameter.

			Partial capture S2 HL3				Full capture S5 HL6			
Equipment	ID	type	#	size		cost k€	#	size		cost k€
Rotary										
Rich solvent pump	P-RICH	Centrifugal	1	21	kW	410	3	130	kW	2330
CO2 pump	P-CO2	Centrifugal	1	250	kW	990	2	550	kW	2890
MEA make-up pump	OP-1	Centrifugal	1	>0	kW	20	2	>0	kW	30
Make-up water pump	OP-2	Centrifugal	1	0	kW	30	2	3	kW	70
Absorber buffer pump	OP-3	Centrifugal	1	40	kW	150	2	80	kW	340
Lean solvent pump	P-LEAN	Centrifugal	1	300	kW	1210	2	710	kW	2630
Stripper buffer pump	OP-4	Centrifugal	1	30	kW	190	2	70	kW	500
Stripper reflux pump	P-RFLX	Centrifugal	1	1	kW	60	2	10	kW	150
Cooling water pump	P-CW	Centrifugal	1	850	kW	4800	2	2080	kW	10640
Intercooler pump	OP-5	Centrifugal	1	70	kW	190	3	150	kW	1940
DCC circulation pump	P-DCC	Centrifugal	1	-	kW	-	2	200	kW	880
Flue gas fan	FAN-1	Blower	1	360	kW	570	3	810	kW	1480
Four-stage compressor	COMP-1 - COMP-4	Centrifugal	1	12540	kW	35790	2	31410	kW	76750
Vessels										
Absorber column	ABS-1	SS316	1	22/8	m	5600	3	-	m	11930
Absorber packing	Sulzer Mellapak 250Y		1	15/8	m	3220	3	-	m	19250
Stripper column	STR-1	SS316	1	28/7	m	3380	2	-	m	7470
Stripper packing	Sulzer Mellapak 250Y		1	20/7	m	1600	2	-	m	10730
Washer column	WASH-1	SS316	1	2/8	m	2380	3	-	m	6790
Washer packing	Sulzer Mellapak 250Y		1	1.4/8	m	780	3	-	m	6968
MEA make-up tank	TANK-1	SS316	1	10	m^3	300	2	-	m^3	680
Absorber buffer tank	TANK-2	SS316	1	10	m^3	290	2	-	m^3	680
Stripper buffer tank	TANK-3	SS316	1	10	m^3	340	2	-	m^3	680
DCC column	DCC	SS316	0	-	m	-	2	-	m	5080
DCC packing	Sulzer Mellapak 250Y		0	-	m	-	2	-	m	13576
Condenser KO drum	RFLX	SS316	1	6/4	m	1400	2	-	m	2370
Knock-out drum		SS316	4	5/3	m	2480	8	-	m	5850
Heat exchangers										
DCC circulation cooler	HX-DCC	Shell&Tube	0	-	m^2	-	3	2220	m^2	2360
Cross heat exchanger	HX-ECO	Shell&Tube	17	16,000	m^2	16730	32	32680	m^2	32070
Stripper condenser	COND	Shell&Tube	1	510	m^2	650	3	2580	m^2	2360
Stripper reboiler	REB	Thermosyphon	13	12,290	m^2	13910	32	30840	m^2	26920
Lean solvent cooler	HX-LEAN	Shell&Tube	3	2,720	m^2	2660	5	3870	m^2	3860
Absorber intercooler	HX-ABS	Shell&Tube	2	1,130	m^2	1360	5	4290	m^2	3770
Intercooler 1	HX-1	Shell&Tube	1	460	m^2	610	2	990	m^2	1580
Intercooler 2	HX-2	Shell&Tube	1	460	m^2	670	2	990	m^2	2420
Intercooler 3	HX-3	Shell&Tube	1	520	m^2	930	2	1110	m^2	1630
Intercooler 4	HX-4	Shell&Tube	2	1,440	m^2	3880	4	3080	m^2	5570
Other										
Pre and post filter			2	-		260	4	-		520
Active carbon filter			1	-		240	2	-		480
Gas piping column		SS316	1	100	m	6920	2	230	m	30940
Total installation cost						115000				307160

A.3 Comparison with data from the literature

Table A.5: Comparisonsof the data in the literature for absorption of CO_2 using 30 wt.% aqueous MEA solvent.

Study		Arasto/Tsupari	IEAGHG		Cormos	Ho		Kuramochi	Kim	Dreillard
Site		Raahe Steel Mill, FI	conceptual western Europe			Ijmuiden, NL		n.a.	n.a., KR	IFPEN mini pilot, FR
Site characteristic		existing, district heating	greenfield, access to Rotterdam; no export of energy (no district heating)			integrated site; district heating		integrated	integrated	Arcelor Mittal data
CO_2 source		HS + CHP	HS + CHP	HS + CHP + coke ovens	HS + CHP + coke ovens + lime kiln	HS + CHP +coke ovens + sinter	BFG	BFG	BFG	BFG
Capture rate (CO_2 source)	%	90	90	90	90	90	90	n.a.	90	90
Capture rate (site)	%	50–75	50	60	50–60	80	30	19	n.a.	n.a.
Scale	Mt CO_2/a	2–3	5.0	6.1	5–6.5	8	3.2	1.3	0.7	n.a.
Heat source		power plant renewal; off-gases	CHP plant fired with NG, BFG, BOFG		NGCC power plant	CHP plant fired with NG, BFG, BOFG		n.a.	CHP fueled by off-gas only	external steam
Specific heat demand	MJ/kg CO_2	3.40	3.03	3.03/3.18	2.95	n.a.	n.a.	4.40	n.a.	3.3–3.6
CO_2 compression	bar	60	110	110	120	100	100	110	150	6
Cost year		2012	2010	2010	2016	2010	2010	2007	2011	2018
Rate of return	%	10	10	10	n.a.	n.a.	n.a.	10	8	n.a.
Life time	years	20	25	25	n.a.	25	25	20	20	n.a.
Cost avoided	[currency]/ tonne CO_2	84–114[1] [EUR]	74 [USD]	81 [USD]	100–150 [EUR]	80 (75–96) [AUD]	76 [AUD]	64 [EUR]	71.7 [USD]	63.6 [EUR]
Cost avoided - levelized	€ 2015/ tonne CO_2 avoided	86–116[1]	60	66	100–150	60 (56–72)	57	72	54	62
Reference		(Arasto et al., 2013; Tsupari et al., 2013)	(IEAGHG, 2013)		(Cormos, 2016)	(Ho et al., 2013)		(Kuramochi et al., 2012)	(Kim et al., 2015)	(Dreillard et al., 2017)

[1] includes transport and storage and carbon credit (EUA)

n.a., Not available