Optimization of Top Gas Recycling Conditions under High Oxygen Enrichment in the Blast Furnace

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Primary steelmaking is among the most energy intensive industrial processes in the world and being mainly coal-based it substantially contributes to the global fossil CO₂ emissions. It is therefore important to study potential ways of suppressing the use of fossil reductants and the rate of emissions in the process. This paper analyzes by simulation and optimization the concept of recycling CO₂-stripped top gas in the blast furnace under massive oxygen enrichment, and its impact on the production economy and emissions of the steel plant. The effect of CO₂ emissions and stripping/storage costs on the optimal states is presented. The system studied is demonstrated to exhibit complex transitions between the optimal states. The findings throw light on the importance of selecting a proper state of operation for achieving a cost-efficient production of steel with reduced environmental impact.

KEY WORDS: blast furnace; top gas recycling; economic optimization; CO₂ emissions.

1. Introduction

The steelmaking industry is one of the most energy intensive sectors and due to the fact that coal is used as the primary source of energy and reduction potential it contributes significantly to the global CO₂ emissions. In primary steelmaking, most of the unit processes, including the blast furnace (BF), are well established and have already evolved to a mature state. Therefore, it is difficult to substantially reduce the energy demand and the emissions. Instead, new innovative processes or sources of energy, or fundamentally new ways of operating existing unit processes have to be found.

In order to suppress the emission rates, the use of non-fossil reductants, e.g., biomass, "hot" BF top gas recycling and CO₂ capture and storage have been proposed as potential solutions.

Even though top gas recycling in the blast furnace is an old idea, and has been practiced already in the 1980’s in several campaigns at Toulachermet in Russia, the awakened interest in the technique is due to the potential of suppressing the coke rate and dramatically reduce emissions if the stripped CO₂ can be stored.

Top gas recycling has been studied extensively with a multi-fluid model developed by the Tohoku University group. Austin et al. studied the options of top gas recycling with simple recycling of cold gas (with reduction in blast volume), recycling with increased oxygen enrichment of the (remaining) blast and hot gas recycling where CO₂ was stripped and both the recycled gas and the oxygen-enriched blast were heated to the blast temperature (1 200°C). The investigators found the concept with hot gas recycling and high oxygen blast promising, with a maximum production rate increase of 25% and a reduction in the reductant rate by 20%. Nogami et al. used a standard case of operation of the furnace as a reference in their study. The simulated raceway conditions (flame temperature and bosh gas volume) and hot metal temperature (at the slag layer) were maintained constant for the novel cases of operation, with plastics injection or with plastics injection combined with top gas recycling. In the latter case, the oxygen and the recycled top gas were blown through the tuyeres at ambient temperature (25°C). The authors found that recycling 37% of the top gas yielded a production increase of about 6% while the reduction agent rate slightly increased. In a later study, Nogami et al. reported detailed results on material and energy flows for a simulated steel plant with BF top gas recycling and concluded that the energy requirement was lower and that the emissions also decreased compared to the standard operation state if the stripped CO₂ would be stored. Murai et al. studied different novel ways of operating the blast furnace for minimizing the emissions and considered top gas recycling with tuyere and shaft injection and massive plastics injection as one option. The authors pointed out the need of an overall analysis of the material and energy flows in the whole plant for assessing the economic advantage of the proposed innovative ways of operation.

Recently, the technique of hot top gas recycling after CO₂ stripping was proven within the ULCOS program where the experimental pilot blast furnace in Luleå, Sweden, was operated under top gas recycling during a test campaign, applying vacuum pressure swing adsorption (VPSA) for CO₂ removal. The results were encouraging, which makes it meaningful to make further investigations of the potential of the technique.

In the work reported in this paper, the top gas recycling
concept with tuyere injection of the CO$_2$-stripped and heated top gas together with oxygen-enriched blast is evaluated by simulation. The performance of the process is optimized with respect to oxygen injection, top gas recycling rate and temperature, and oil injection rate using a linearized model of the blast furnace. However, since top gas recycling dramatically affects the material and energy flows in the whole plant, the entire system should be studied in order to evaluate the overall benefits. The present study therefore comprises the coke plant, sinter plant, hot stoves, blast furnace, stripping unit, the basic oxygen furnace and power plant, paying attention to the costs and the emissions of the process, and the optimal states of operation of the system are studied with different price settings for CO$_2$ emissions and stripping/storage.

2. The System

We study the units in an integrated steel works from the raw material processing units through one blast furnace to the basic oxygen furnace (BOF), with liquid steel as the main product (Fig. 1). The models were developed on the basis of process data from a Finnish integrated steel works. The core of the analysis is the blast furnace with oil injection and top gas recycling, which is modeled in more detail. In the next subsection, the models are briefly described. For more details, the reader is referred to Refs. 15, 16).

2.1. Process Models and CO$_2$ Emissions

The main focus of the study is put on the blast furnace, for which a more detailed description is used while the other unit processes are described in a simplified way (see Appendix). The sinter plant is modelled by expressing the produced sinter, the required coke and limestone as well as the recovered heat as linear functions of the ore inflow. The coke plant is described by linear relations between the mass flow rate of feed coal and the mass flow rate of coke and volume flow rate of (purified) coke oven gas (COG). The (internal) flow rate of coke, available for the blast furnace, is given by the difference of the coke produced and the coke consumed in the sinter plant. The hot stove model considers the compression and heating of the strongly oxygen enriched blast and the recycled and CO$_2$-stripped top gas to the “blast” temperature. Because of the possibility to recycle practically all BF top gas, the hot stoves are assumed to be heated by burning coke oven gas. The stove system was modeled as a single continuous counter-current heat exchanger in steady state, giving the share, $\alpha$, of the available coke oven gas that must be burnt for reaching the aim blast temperature. The basic oxygen furnace model expresses the mass flow of liquid steel ($ls$), and the required volume flow rate of oxygen and the resulting volume flow rate of off-gases as functions of the mass flow of hot metal ($hm$). In the power plant a mixture of the arising gases (remaining parts of the blast furnace top gas, coke oven gas and part ($k$) of the BOF gas) is used to produce electricity and hot water for district heating.

For the blast furnace, a thermodynamic first-principles model$^{17,18}$ was applied to develop a linear model, where it was assumed that the part of the top gas, $\beta$, that is recycled goes to a stripping unit with 95% separation of carbon dioxide. The stripped CO$_2$ is taken to be stored, while the stripped gas is compressed and then heated in the hot stoves to the “blast” temperature ($T_{bl}$) together with the blast, i.e., the pure (99%) oxygen possibly mixed with air. The option of using a mixture of air and oxygen turned out to be necessary for enabling different recycling degrees of the top gas, still keeping the BF operation within the imposed constraints, as observed by Austin et al.$^6$ The oxygen content of the blast (i.e., nitrogen–oxygen mixture) was allowed to
vary in the interval $[80\%, 99\%]$. The thermodynamic BF model was run under a large number of input combinations, with input variables uniformly distributed (with 4–6 different values of each) within their admissible regions (Table 1). This also resulted in infeasible furnace states, e.g., with too low or high flame temperatures, so the data set was pruned by removing the infeasible solutions, leaving more than 1 250 feasible solutions. These were used to build the linearized BF model, with the volume flow rate of recycled top gas, the total volume flow rate of blast (pure oxygen with possible air additions) and the oxygen content, the specific oil rate, the temperature of the injected tuyere gases, and the specific pellet and limestone rates as inputs:

$$y_i = K_{i,0} + K_{i,1} \frac{\dot{V}_{rg}}{\text{km}^3/\text{n/h}} + K_{i,2} \frac{\dot{V}_{bl}}{\text{km}^3/\text{n/h}} + K_{i,3} \frac{Y_{O_2,bl}}{\%} + K_{i,4} \frac{m_{al}}{\text{kg/t/hm}} + K_{i,5} \frac{T_{bl}}{\text{C}} + K_{i,6} \frac{m_{pel}}{\text{kg/t/hm}} + K_{i,7} \frac{m_{line,BF}}{\text{kg/t/hm}} i = 1, \ldots, 14 \quad (1)$$

Table 1 lists the inputs (first seven rows) and the outputs (next fourteen rows) and their lower and upper limits, and finally two other constraints.

The CO$_2$ emissions from the system are calculated from a carbon balance equation, considering the outflows of carbon in the liquid steel, and the stripped and stored (subscript str) CO$_2$

$$\dot{m}_{CO_2} = \frac{44}{12} (\dot{m}_{coal} X_{C,coal} + \dot{m}_{lime} X_{C,lime} + \dot{m}_{oil} X_{C,oil}) + \dot{m}_{cokext} - \dot{m}_{CO_2,str} \quad (2)$$

where the total mass flow rate of limestone is the sum of the requirement from the sinter plant, blast furnace and BOF, the mass flow rate of external (bought) coke is the difference between the coke requirement in the BF and the supply from the coke plant, and the mass flow of stripped and stored carbon dioxide is

$$\dot{m}_{CO_2,str} = 0.95 J_{1} Y_{CO_2,BF} \frac{44}{22.4} \frac{\text{t}}{\text{m}^3/\text{n}} \quad (3)$$

2.2. Minimization of Steelmaking Costs

The system was optimized by minimizing the specific cost of production of liquid steel with respect to the seven input variables of the BF model (Eq. (1)), with the objective function

$$F = \frac{\dot{m}_{scr} \cdot c_{scr} + \dot{m}_{pel} \cdot c_{pel} + \dot{m}_{coal} \cdot c_{coal} + \dot{m}_{cokext} \cdot c_{cokext} + \dot{V}_{O_2} \cdot c_{O_2} + \dot{m}_{line,BF} \cdot c_{line,BF} + \dot{m}_{CO_2} \cdot c_{CO_2} \cdot \dot{Q}_{Oh} \cdot c_{heat}}{\dot{m}_{ls} \cdot c_{heat}} \quad (4)$$

where $c_i$ are specific or volumetric cost terms, and the total inflow of oxygen considers the pure oxygen flows to the BF and the BOF while the net outflows of power and district heat from the system appear in the last two credit terms. No credit was given for the production of slag, and the use of scrap in the BOF was simply fixed at $\dot{m}_{scrap} = 0.25 \dot{m}_{hm}$. The nonlinear programming problem, $\min (F)$, was solved numerically subject to the constraints of the variables and to upper limits on some of the raw material flows (cf. Table 1) with the cost factors presented in Table 2.

<table>
<thead>
<tr>
<th>Variable</th>
<th>Symbol</th>
<th>Unit</th>
<th>Range</th>
</tr>
</thead>
<tbody>
<tr>
<td>Recycled top gas</td>
<td>$\dot{V}_{rg}$</td>
<td>km$^3$/h</td>
<td>70-250</td>
</tr>
<tr>
<td>Blast volume</td>
<td>$\dot{V}_{bl}$</td>
<td>km$^3$/h</td>
<td>0-140</td>
</tr>
<tr>
<td>Blast oxygen</td>
<td>$X_{O_2,bl}$</td>
<td>vol.%</td>
<td>80-99</td>
</tr>
<tr>
<td>Specific oil rate</td>
<td>$m_{oil}$</td>
<td>kg/t/hm</td>
<td>0-200</td>
</tr>
<tr>
<td>Blast temperature</td>
<td>$T_{bl}$</td>
<td>°C</td>
<td>850-1100</td>
</tr>
<tr>
<td>Specific pellet rate</td>
<td>$m_{pel}$</td>
<td>kg/t</td>
<td>0-600</td>
</tr>
<tr>
<td>Specific limestone rate</td>
<td>$m_{ls,LF,LF}$</td>
<td>kg/t</td>
<td>0-100</td>
</tr>
<tr>
<td>Production rate</td>
<td>$m_{hm}$</td>
<td>t/h</td>
<td>150-196</td>
</tr>
<tr>
<td>Specific coke rate</td>
<td>$m_{cokext}$</td>
<td>kg/t</td>
<td>$\geq 0$</td>
</tr>
<tr>
<td>Flame temperature</td>
<td>$T_{F}$</td>
<td>°C</td>
<td>1900-2800</td>
</tr>
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<td>Top gas temperature</td>
<td>$T_{EP}$</td>
<td>°C</td>
<td>100-250</td>
</tr>
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<td>Bosch gas volume</td>
<td>$\dot{V}_{bg}$</td>
<td>km$^3$/h</td>
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</tr>
<tr>
<td>Solid residence time</td>
<td>$\tau$</td>
<td>h</td>
<td>6.0-9.0</td>
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<tr>
<td>Slag basicity</td>
<td>$B$</td>
<td>-</td>
<td>1.0-1.2</td>
</tr>
<tr>
<td>Slag rate</td>
<td>$m_{scrap}$</td>
<td>kg/t</td>
<td>$\geq 0$</td>
</tr>
<tr>
<td>Top gas volume</td>
<td>$\dot{V}_{bg}$</td>
<td>km$^3$/h</td>
<td>$\geq 0$</td>
</tr>
<tr>
<td>Top gas CO</td>
<td>$Y_{O_2,bl}$</td>
<td>vol.%</td>
<td>$\geq 0$</td>
</tr>
<tr>
<td>Top gas CO$_2$</td>
<td>$Y_{CO_2,bl}$</td>
<td>vol.%</td>
<td>$\geq 0$</td>
</tr>
<tr>
<td>Top gas H$_2$</td>
<td>$Y_{H_2,bl}$</td>
<td>vol.%</td>
<td>$\geq 0$</td>
</tr>
<tr>
<td>Top gas N$_2$</td>
<td>$Y_{N_2,bl}$</td>
<td>vol.%</td>
<td>$\geq 0$</td>
</tr>
<tr>
<td>Top gas heating value</td>
<td>$H_{EP}$</td>
<td>MJ/m$^3$</td>
<td>$\geq 0$</td>
</tr>
<tr>
<td>Sinter</td>
<td>$m_{scrap}$</td>
<td>t/h</td>
<td>0-160</td>
</tr>
<tr>
<td>Own coke</td>
<td>$m_{cr,bf}$</td>
<td>t/h</td>
<td>0-55</td>
</tr>
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</table>
Table 3. Costs of CO₂ emission and stripping/capture, as well as
as credit for heat.

<table>
<thead>
<tr>
<th>Case</th>
<th>c_CO₂ (€/t)</th>
<th>c_em (€/t)</th>
<th>c_mstr (€/MWh)</th>
</tr>
</thead>
<tbody>
<tr>
<td>A</td>
<td>0</td>
<td>0</td>
<td>10</td>
</tr>
<tr>
<td>B</td>
<td>20</td>
<td>0</td>
<td>10</td>
</tr>
<tr>
<td>C</td>
<td>0</td>
<td>20</td>
<td>10</td>
</tr>
<tr>
<td>D</td>
<td>20</td>
<td>20</td>
<td>10</td>
</tr>
<tr>
<td>E</td>
<td>40</td>
<td>20</td>
<td>10</td>
</tr>
<tr>
<td>F</td>
<td>40</td>
<td>20</td>
<td>30</td>
</tr>
</tbody>
</table>

3. Results and Discussion

The system was optimized for production rates in the range 150–195 tls/h, since 196 tls/h was the maximum production under the present process constraints; this is 6% higher than the maximum production rate estimated for conventional operation. A deeper analysis of the results reveals that, e.g., the drop in the optimum recycling degree at the production rate 175–180 tls/h is due to the fact that a more economical solution, which for production rates \( m_{\text{ls}} < 180 \) tls/h is infeasible because of a too low top gas temperature \( T_{\text{BF}} \), gets feasible at \( m_{\text{ls}} = 180 \) tls/h with \( T_{\text{BF}} \) at its lower boundary (100°C). The fact that the former global minimum (with \( \beta = 73\% \)) becomes a local one at higher production rates is seen in the asterisks depicting the local minima. The production costs at this state are still about 1.6€/t higher at \( m_{\text{ls}} = 180 \) tls/h.

Finally, it is interesting to note that the optimization problem does not always give a unique solution: For some of the cases studied, one or several other local minima (i.e., points that are better than all points in their neighborhood) occur, which correspond to higher costs than in the global minimum, represented by the line in Fig. 2. These have been depicted by asterisks and are seen to appear mainly at higher production rates (\( m_{\text{ls}} \geq 180 \) tls/h) and correspond to 1–3€/t is higher costs than in the global minimum. The global minimum was located by starting the iterative search in the optimization from a large number of initial values of the unknowns in Eq. (1). A deeper analysis of the results reveals that, e.g., the drop in the optimum recycling degree at the production rate 175–180 tls/h is due to the fact that a more economical solution, which for production rates \( m_{\text{ls}} < 180 \) tls/h is infeasible because of a too low top gas temperature \( T_{\text{BF}} \), gets feasible at \( m_{\text{ls}} = 180 \) tls/h with \( T_{\text{BF}} \) at its lower boundary (100°C). The fact that the former global minimum (with \( \beta = 73\% \)) becomes a local one at higher production rates is seen in the asterisks depicting the local minima. The production costs at this state are still about 1.6€/t higher at \( m_{\text{ls}} = 180 \) tls/h.

The fact that multiple optima exist illustrates the complexity of the optimization task at hand, where the nonlinearity introduced by the recycling loop is the origin of the ambiguity. However, in the following cases only the solutions corresponding to the global optima will be illustrated for the sake of brevity.

Case B

Figure 3 depicts the results for a price of emissions roughly corresponding to the present one in Europe, i.e., \( c_{\text{CO₂}} = 20 \)€/t, but with zero stripping/storage costs, \( c_{\text{mstr}} = 0 \)€/t. The increased emission price is seen to raise the steel price by 12–15€/t and results in maximum top gas recycling (about 98.6%, where the remaining 1.4% is sufficient to purge the nitrogen that enters the blast furnace) except at 195 tls/h, combined with low blast volume and maximum oxygen enrichment. The reason for this is that the higher emission price raises the costs, so the optimization leads to a point where the emissions are considerably lower, which is achieved at maximum top gas recycling (=maximum stripping). The blast volume increases monotonically with the production rate. The oil rate is relatively low, and varies
between 40 and 60 kg/t hm, exhibiting a maximum at a production rate of 165 t s/h, while the RAR is very low: It decreases from about 320 to 310 kg/t hm.

Case C

The optimum states for no emission costs but with costs of CO₂ stripping/storage of \( c_{str}/H11005 \), illustrated in Fig. 4, is very similar with those of Case A, but follow a path through the first local minimum (first asterisk of Fig. 2) with higher blast volume, oil rate and RAR, and lower gas recycling degree up to 180 t ls/h. After this point identical values with Case A are obtained. The costs are initially about 10 €/t higher than in Case A, but decrease to be only 5 €/t higher at the highest production rate.

Case D

The case with equal, but higher, costs of emissions and stripping/storage, \( c_{CO2}/H11005 = c_{str}/H11005 \), yields results with more complex transitions between states of operation (Fig. 5). The steel price is rather independent of the production rate, due to the fact that the optimization method has found ingenious transitions between the operation states along with changes in the production rate. The blast volume is initially low but increases with the production rate, and in larger steps at 165–170 t ls/h and 185–190 t ls/h. These changes are connected with larger transitions in the other variables. At low production rates, the optimal states apply full top gas recycling, and maximum oxygen enrichment of the blast, which at 170 t ls/h are decreased to 80% and 75%, respectively. For production rates of 190 t ls/h and higher, the recycling degree drops further to about 55%. The oil injection rate increases throughout the interval with major transitions at the same points as the blast volume and recycling degree, and the RAR follows the same pattern but on a low level; it grows from 320 to 380 kg/t hm.

Case E

Further increase of the emission costs to 40 €/t, keeping the price of stripping/storage at 20 €/t yields practically
identical results as for Case B, which is expected, since the price difference between the emission and stripping/storage is the same. The steel price, however, grows to 244–246 €/t ls, with the lowest price at a production rate of 170–175 t ls/h.

3.2. Optimal States and Active Constraints

A deeper analysis of the results of the previous subsection reveals that the six cases (A–F) studied correspond to only a few distinct states of operation of the blast furnace. The optimal states for a moderate steel production rate (175 t ls/h) have been collected in Table 4, where three different states occur: Cases A, D and F result in identical process variables, likewise Cases B and E, while Case C somewhat differs from the first category. Common for all cases are the sinter and pellet rates and slag basicity, and the flame temperature is at, or close to, its lower constraint. A clear difference in states between the ones corresponding to maximum or lower top gas recycling degrees can be seen. The former applies full oxygen blast, maximum preheating of the injected gases and yields a RAR of only 312 kg/t hm, but also no net credit for power and heat production (Table 5). The observed reduction in the RAR exceeding 100 kg/t hm is in good agreement with results reported by other investigators and also with the findings from the experimental blast furnace (EBF) in Luleå, Sweden.13)

3.3. Emissions

As noted in the previous subsections, cost structures with high price of emissions compared to the price of CO₂ stripping/storage give rise to low emission rates since this forces the optimization to avoid emission costs. Figure 6 depicts the specific emissions for Cases A, B and C. The clearly lowest emissions correspond to Case B where full top gas recycling is applied. Compared to the specific emissions of the plant at conventional operation where the emissions are about 1.6 t CO₂/t ls,5) the reduction is 1.0–1.1 t CO₂/t ls, in good agreement with the results observed in the campaign of the EBF,13) where also about 0.7 t CO₂/t ls was ascribed to the CO₂ stripping/storage. The dramatic increase in the specific emissions at 175–180 t ls/h for Cases A and C is due to the simultaneous decrease in the top gas recycling degree. Table 5 reports the emissions for all cases at a production rate of 175 t ls/h.

4. Conclusions

This paper has applied mathematical simulation and optimization to analyze the concept top gas recycling in the blast furnace for suppressing the coke rate and emissions in the steel plant. Using models of the unit processes in an integrated steel plant, the costs of liquid steel production were minimized subject to process constraints. The study demonstrated that the optimal recycling degree of top gas varies considerably with the cost structure of CO₂ emissions and stripping/storage. In some cases it is beneficial to recycle as much top gas as possible, both from an economic and environmental point of view, but for other prices of CO₂ emissions and stripping/storage a lower degree is economically justified. Furthermore, the energy requirements of the whole plant, as well as possible deliveries (of district heat) to nearby society affect the optimal states. The findings also give indications on states of operation that stay (close to) optimal despite changes in the production rate. For instance, maintaining minimum steel production costs under the price structure of Case D (Fig. 5) is difficult due to the required major transitions of the BF operation state.

Table 4. Process variables for a steel production rate of mₚ=175 t/h. The flow of sinter is maximum (160 t/h), the slag basicity is minimum (1.00), and the specific pellet rate is 495 kg/t hm for all cases. Variables at their constraints are marked in bold.

<table>
<thead>
<tr>
<th>Variable</th>
<th>Cases A, D, F</th>
<th>Cases B, E</th>
<th>Case C</th>
</tr>
</thead>
<tbody>
<tr>
<td>Blast volume (km³/h)</td>
<td>37.6</td>
<td>22.8</td>
<td>38.8</td>
</tr>
<tr>
<td>Blast oxygen content (%)</td>
<td>80.0</td>
<td>99.0</td>
<td>80.0</td>
</tr>
<tr>
<td>Top gas recycling degree (%)</td>
<td>72.9</td>
<td>90.6</td>
<td>71.0</td>
</tr>
<tr>
<td>Top gas volume (km³/h)</td>
<td>185.1</td>
<td>186.2</td>
<td>184.7</td>
</tr>
<tr>
<td>Coal rate (t/h)</td>
<td>60.5</td>
<td>67.8</td>
<td>60.4</td>
</tr>
<tr>
<td>Specific coke rate (kg/t hm)</td>
<td>227</td>
<td>281</td>
<td>227</td>
</tr>
<tr>
<td>Specific oil rate (kg/t hm)</td>
<td>125</td>
<td>51</td>
<td>131</td>
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<tr>
<td>RAR (kg/t hm)</td>
<td>352</td>
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<td>358</td>
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<tr>
<td>Blast temperature (°C)</td>
<td>1066</td>
<td>1100</td>
<td>1015</td>
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<tr>
<td>Specific limestone rate (kg/t hm)</td>
<td>3.4</td>
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<tr>
<td>Flame temperature (°C)</td>
<td>1908</td>
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<td>Bush gas volume (km³/h)</td>
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<td>Solid residence time (h)</td>
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<tr>
<td>Top gas CO (%)</td>
<td>37.6</td>
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<td>37.6</td>
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<tr>
<td>Top gas CO₂ (%)</td>
<td>31.7</td>
<td>35.2</td>
<td>31.7</td>
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<tr>
<td>Top gas H₂ (%)</td>
<td>15.7</td>
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<td>16.2</td>
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<td>Top gas N₂ (%)</td>
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<tr>
<td>Top gas heating value (MJ/m³/h)</td>
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<td>7.0</td>
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<td>Slag rate (kg/t hm)</td>
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<td>201</td>
<td>196</td>
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<td>Compressor power (MW)</td>
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<td>7.2</td>
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<td>Net power production (MW)</td>
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<td>District heat (MW)</td>
<td>82.9</td>
<td>20.5</td>
<td>89.9</td>
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</table>

Table 5. Price of liquid steel, specific emissions and credit for power and heat for a steel production rate of mₚ=175 t/h.

<table>
<thead>
<tr>
<th>Variable</th>
<th>Case A</th>
<th>Case B</th>
<th>Case C</th>
</tr>
</thead>
<tbody>
<tr>
<td>Price of liquid steel (€/t)</td>
<td>200.0</td>
<td>219.1</td>
<td>215.0</td>
</tr>
<tr>
<td>Specific CO₂ emissions (€/t)</td>
<td>0.83</td>
<td>0.51</td>
<td>0.86</td>
</tr>
<tr>
<td>Credit power and heat (€/t)</td>
<td>7.9</td>
<td>0.0</td>
<td>8.8</td>
</tr>
</tbody>
</table>

Fig. 6. Specific emissions vs. steel production rates for Cases A, B and C (cf. Table 3).
Future work will extend the analysis to include all states between conventional BF operation and full top gas recycling by relaxing the lower constraint of the blast oxygen content. This will require a detailed treatment of the local optima encountered for different recycling degrees. Another future prospect is to analyze the distribution of the hot reducing gases between tuyere and shaft injection.

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Nomenclature
\[ c : \text{ Cost factor (€/t, €/km}^3\text{m or C/MWh)} \]
\[ B : \text{ Slag basicity (}=X_{\text{CaO}}/X_{\text{SiO}_2}) \]
\[ \dot{E} : \text{ Energy flow (MW)} \]
\[ F : \text{ Economic objective function (€/t ls)} \]
\[ G : \text{ Heat conductivity (MW/°C)} \]
\[ H : \text{ Effective heating value (MJ/kg or MJ/nm}^3\text{)} \]
\[ K : \text{ Coefficient in linear BF model, Eq. (1)} \]
\[ \dot{m} : \text{ Specific mass flow (kg/t)} \]
\[ \dot{m} : \text{ Mass flow rate (t/h)} \]
\[ P : \text{ Electric power (MW)} \]
\[ \dot{Q} : \text{ Heat flow (MW)} \]
\[ T : \text{ Temperature (°C)} \]
\[ \dot{V} : \text{ Volume flow rate (km}^3\text{h)} \]
\[ X : \text{ Mass ratio} \]
\[ Y : \text{ Molar ratio} \]

Greek
\[ \alpha : \text{ Share of coke oven gas burned in hot stoves} \]
\[ \beta : \text{ Share of BF top gas recycled} \]
\[ \chi : \text{ Parameters in Eq. (A2)} \]
\[ \Delta : \text{ Difference operator} \]
\[ \gamma : \text{ Overall power production efficiency, Eq. (A6)} \]
\[ \eta : \text{ Distribution factor of heat and power} \]
\[ \varphi : \text{ Parameters in Eq. (A5)} \]
\[ \kappa : \text{ Share of BOF gas burned in power plant} \]
\[ \xi : \text{ Parameters in Eq. (A1)} \]
\[ \tau : \text{ Residence time of burden (h)} \]

Subscripts
\[ \text{BF} : \text{ Blast furnace} \]
\[ \text{bg} : \text{ Bosh gas} \]
\[ \text{bl} : \text{ Blast} \]
\[ \text{BOF} : \text{ Basic oxygen furnace} \]
\[ \text{C} : \text{ Carbon} \]
\[ \text{coal} : \text{ Coking coal} \]
\[ \text{CO} : \text{ Carbon monoxide} \]
\[ \text{COG} : \text{ Coke oven gas} \]
\[ \text{CO}_2 : \text{ Carbon dioxide} \]
\[ \text{cok} : \text{ Coke} \]
\[ \text{dh} : \text{ District heat} \]
\[ \text{el} : \text{ Electricity} \]
\[ \text{ext} : \text{ External (coke)} \]
\[ \text{fl} : \text{ (Raceway adiabatic) flame} \]
\[ \text{H}_2 : \text{ Hydrogen} \]
\[ \text{heat} : \text{ Heat} \]
\[ \text{hm} : \text{ Hot metal} \]
\[ i : \text{ General index} \]
\[ \text{int} : \text{ Internal (cokem)} \]
\[ j : \text{ General index} \]
\[ \text{lime} : \text{ Limestone} \]
\[ \text{ls} : \text{ liquid steel} \]
\[ \text{O}_2 : \text{ Oxygen} \]
\[ \text{oil} : \text{ (Heavy) oil} \]
\[ \text{ore} : \text{ Ore} \]
\[ \text{pel} : \text{ Pellet} \]
\[ \text{pp} : \text{ Power plant} \]
\[ \text{rg} : \text{ Recycled gas} \]
\[ \text{scrap} : \text{ Scrap} \]
\[ \text{sint} : \text{ Sinter} \]
\[ \text{slag} : \text{ Slag} \]
\[ \text{st} : \text{ Hot stoves} \]
\[ \text{str} : \text{ Stripped and stored} \]

A dot above a symbol denotes a flow rate.

REFERENCES

Appendix
This appendix briefly outlines the simplified models of the coke plant, sinter plant, hot stoves, basic oxygen furnace and power plant used in the study.

Coke Plant
Linear relations between the mass flow rate of feed coal and the mass flow rate of coke and volume flow rate of (purified) coke oven gas (COG) are assumed
\[ \dot{m}_{\text{cokem}} = \xi \dot{m}_{\text{coal}}, \quad \dot{V}_{\text{COG}} = \xi \dot{m}_{\text{cokem}} \quad \text{......(A1a, 1b)} \]
Sinter Plant

Only the raw materials iron ore, coke and limestone are considered, and in addition to them, the recovered heat is also taken into account, i.e.,

\[ \dot{m}_{\text{sint}} = \chi_1 \dot{m}_{\text{ore}}, \quad \dot{m}_{\text{coke,sint}} = \chi_2 \dot{m}_{\text{sint}}, \quad \dot{m}_{\text{limestone,sint}} = \chi_3 \dot{m}_{\text{sint}}, \]

\[ \dot{Q}_{\text{sint}} = \chi_4 \dot{m}_{\text{sint}} \]

which gives the (internal) flow rate of coke available for the blast furnace:

\[ \dot{m}_{\text{coke,int}} = \dot{m}_{\text{coke}} - \dot{m}_{\text{coke,sint}} \] (A3)

Hot Stoves

The strongly oxygen-enriched blast and the recycled and CO₂-stripped top gas are compressed and then heated in the hot stoves, which are assumed to operate as a single continuous counter-current heat exchanger in steady state with the heat transferred from coke oven gas burned on the heat releasing side. The heat flow rate is given by

\[ \dot{Q}_{\text{st}} = G_{\text{st}} (\Delta T_1 - \Delta T_2) / \ln(\Delta T_1 - \Delta T_2) \] (A4)

where \( G_{\text{st}} \) is the heat conductance of the stoves and subscripts 1 and 2 denote the conditions at the two ends of the stove system. This equation is combined with two energy balance equations, for the heat releasing and heat accepting sides, yielding the demand of COG for reaching the required “blast” temperature.

Basic Oxygen Furnace

The mass flow of liquid steel, as well as the volume flow rates of oxygen consumed and off-gases arising in the BOF are given as function of the mass flow of hot metal (hm)

\[ m_{\text{ls}} = \varphi_1 \dot{m}_{\text{hm}} + \dot{m}_{\text{scrap}}, \quad \dot{V}_{\text{O}_2,\text{BOF}} = \varphi_2 \dot{m}_{\text{hm}}, \quad \dot{V}_{\text{BOF}} = \varphi_3 \dot{m}_{\text{hm}} \]

...........(A5a–5c)

Power Plant

Power and heat are produced in the power plant, where a mixture of the blast furnace top gas and the coke oven gas (excluding the part consumed by the hot stoves), and part (\( k \)) of the BOF gas are burned. Given the state of the high pressure steam and the turbine efficiency, as well as the water temperature for the district heat production, we have

\[ P_{\text{pp}} = \eta \eta' \dot{E}_{\text{pp}}, \quad \dot{Q}_{\text{pp}} = (1 - \eta) \gamma \dot{E}_{\text{pp}}, \]

\[ \dot{E}_{\text{pp}} = (1 - \alpha) \dot{V}_{\text{COG}} H_{\text{COG}} + (1 - \beta) \dot{V}_{\text{BF}} H_{\text{BF}} + \kappa \dot{V}_{\text{BOF}} H_{\text{BOF}} \]

...........(A6a–6c)

where \( \eta \) is the distribution factor of energy between power and heat, \( \gamma \) is the overall efficiency of the power plant, and \( H_i (i = \text{BF, COG or BOF}) \) is the effective heating value.

The model parameters used were \( \zeta_1 = 0.695, \ zeta_2 = 319.7 \)
m³n/t, \( \chi_1 = 1.042, \ \chi_2 = 0.0460, \ \chi_3 = 0.0714, \ \chi_4 = 85.12 \)
MJ/t, \( \varphi_1 = 0.8953, \ \varphi_2 = 45.62 \)
m³n/t, \( \varphi_3 = 41.48 \)
m³n/t, \( \kappa = 0.5, \ \gamma = 0.83 \) and \( \eta = 0.2 \), and the effective heating values of the off-gases were \( H_{\text{COG}} = 16.8 \)
MJ/m³ and \( H_{\text{BOF}} = 11.4 \)
MJ/m³.