Numerical simulation of the blast-furnace process

G. A. Flierman and H. Oderkerk

A description is given of a dynamic blast-furnace model developed at Hoogovens. The model has been used to survey the possibilities for replacement of a substantial part of the metallurgical coke by heavy fuel oil. It appears that oil injection affects the pig-iron temperature via the flame temperature. Compensation of this effect either by an increased blast temperature or by oxygen enrichment of the blast is discussed. Transient behaviour of the blast furnace following a sudden variation of an input factor has been studied. Coke equivalents of the variations applied have been calculated. Some examples are given of a simulation of blast-furnace incidents and their consequences.

LIST OF SYMBOLS

\( a_{Si} \) silicon activity in liquid iron
\( a_{SiO_2} \) silica activity in slag
C, Fe, etc. mean quantities of solid or liquid material in a cell, kmol
CO, H_2O, etc. mean quantities of gaseous compounds passing the cell, kmol/min
\( E_i \) activation energy for reaction \( i \), kcal/kmol
\( F \) surface-area factor in silica reduction
\( f \) fraction of a fusible compound that is in the liquid state
\( g_i \) rate constant for reaction \( i \)
\( GAS \) mean quantity of gas passing through a cell, kmol/min
\( h \) height above tuyere plane, m
\( H_2O(l) \) liquid water in coke, kmol
\( K_i \) equilibrium constant for reaction \( i \)
pCO, pH_2, etc. partial pressures of gaseous compounds, atm
p*H_2O saturation pressure of water vapour, atm
\( P_{TOP} \) top pressure, atm
\( R \) gas constant, 1.985 kcal/kmol, °C
\( r \) furnace radius, m
\( r' \) radius of deadman, m

\( t \) time step, min
\( T_C \) mean coke temperature, K
\( T_G \) mean gas temperature, K
\( T_O \) mean ore temperature, K
\( u \) relative velocity of liquids with respect to the coke
\( V_i \) rate of reaction \( i \), kmol per cell per min
\( x \) mean degree of oxidation of an ore, atoms O/atom Fe

The blast-furnace process is a continuous countercurrent process involving heat and mass transfer. It is also one of the most complicated processes in metallurgy.

First, the number of tasks that have to be performed inside a single reaction vessel is unusually large. Second, the raw materials from which the iron is to be produced contain many impurities which, owing to the high temperatures prevailing in the furnace, often take part in the chemical reactions. Moreover, the composition of the raw materials is not always the same. Third, variations in the gas-distribution pattern cannot be fully prevented even if a movable throat armour is used to improve the burden distribution. Finally, mechanical failures may cause major process disturbances. Consequently, the process is seldom, if ever, in a steady state, which is illustrated by random fluctuations in pig-iron
The inherent imperfections of the process depicted above can be very disturbing when it is attempted to derive relationships between variables from blast-furnace data. Full-scale trials in which one of the input parameters undergoes a sufficiently large variation to allow calculation of the desired relationship with reasonable accuracy are quite costly, and usually interfere with the interests of the production department. A suitable mathematical model of the process may be of considerable help in determining whether an industrial experiment is justified. For mathematical models based on reaction kinetics, see Refs. 1–8. However, only the last two models are dynamic in character, i.e., they permit the calculation of transient states as well as the final steady state. Results of the last model are yet to be published.7,9

Fielden et al.7 have calculated the responses in silicon content and pig-iron temperature as well as the changes in the top gas when blast temperature, coke rate, or oil consumption were varied stepwise. Flierman et al.2 have calculated the coke consumption as a function of rating, reducibility, coke reactivity, degree of prereduction, hot-blast temperature, and fuel-oil injection rate. The use of the model, however, was limited to the calculation of steady states. Besides, it permitted the use of only one ore at a time. The dynamic model to be presented in this paper suffers from neither of these shortcomings.

OUTLINE OF CALCULATION METHOD

Description of the process in terms suitable for a mathematical model

The blast furnace is described as a reaction vessel with even distribution of both gaseous and solid/liquid flow over the furnace cross-section. The deadman is considered to be impermeable for both liquids and gases. At any time, the actual state of a blast-furnace process can be defined by

(i) the input variables, i.e., the blast conditions, and the amount and compositions of the materials charged at the top
(ii) the characteristics of the furnace itself, such as dimensions and specific heat losses
(iii) the temperatures and concentrations in the furnace as a function of height.

If the variables mentioned above under (i) and (ii) are kept at constant values, the process will in due time arrive at a stationary state, that is to say, the temperatures and concentrations in the furnace at a given level will no longer vary with time. The time whereafter the stationary state is reached will depend upon the magnitude of the difference between the original temperatures and concentrations and those of the steady state to be attained. This state itself obviously depends only on the input variables and the furnace characteristics.

For the purpose of a numerical calculation, the contents of the furnace above the tuyere level (with the exclusion of the deadman) are divided into a large number of layers having the following properties:

(i) all characteristics, such as temperatures and quantities of the various materials, are uniformly distributed within a layer
(ii) each individual layer retains its identity throughout its descent from the stockline to the tuyere level.

A layer being a very small amount of material (i.e., the materials needed to produce 10 kilograms of ‘standard’ pig iron) it is convenient to combine a number of layers in a cell. The properties of a cell are calculated from those of the individual layers. Once a cell has been formed all properties are again evenly distributed within that cell. The combination of layers to cells is variable. The number of layers in a cell is chosen such that the expected change in gas temperature is about 30°C and that the increment in oxygen content of the gas does not exceed 0.1 g-atom oxygen per mole of gas. The result of this procedure is that the working volume of the furnace above the tuyere plane is divided into about 70 cells, which, nevertheless, only small quantities of material are being considered where the gradients are large. This is illustrated in Fig. 1.

At the start of the calculation the temperatures and quantities in every cell must be known. A standard set of data representing a reference period of the furnace under consideration is used for this purpose. From the blast conditions, the amount of oxygen per minute available for coke combustion is determined. Subsequently, a calculated number of layers taken from one or more of the undermost cells is split off and the coke content is burned with the hot blast. At a given blast rate, the number of layers, or rather the amount of carbon they contain, determines the length of the time step t, which at present is 10 minutes. Quantities and compositions of pig iron and slag are calculated, and these products are added to the contents of the hearth. The charge column is moved downward until the tuyere level has been reached again. Quantity, composition, and enthalpy of the gas phase are calculated, assuming that the resulting gas contains no H₂O or CO₂. Heat loss at the tuyere level is subtracted from the gas enthalpy, whereafter the gas temperature is calculated.

Subsequently, a new cell l is formed and contacted with the gas. It will be clear that, although the properties of the gas entering the cell are constant, those of the gas leaving it will gradually change. Besides, temperature and concentration gradients will develop inside the cell. For the next cell things will be even more complicated as the properties of the entering gas will no longer be constant.

To avoid the inherent mathematical difficulties the situation pictured above was replaced by a simpler one that is physically equivalent, but very much easier to calculate. In other words, mathematical transformation of the partial differential equations describing the continuous process into finite difference equations is avoided by replacing the process itself by a discrete one. This
Expected values for the mean temperatures and concentrations needed to start the iteration process are derived from the results calculated for the preceding cells. In the calculations, volume is used as a geometric parameter rather than height. Matter taking volume disappears from the process in the following ways:

Coke

(i) by combustion with blast oxygen
(ii) by solution in liquid iron
(iii) by reaction with the slag constituents FeO, MnO and SiO₂
(iv) by reaction with CO₂ and H₂O

Ores and solid iron

(i) by melting

At the end of each time interval, an appropriate number of layers is charged to compensate for the change in total volume. The fraction of voids is assumed to be 0.35.

Heat transfer and thermal balances

Three discriminate temperatures are calculated, i.e. the mean temperature of the ores (or temperature of liquid iron and slag, as the case may be), the mean temperature of the coke, and the gas temperature. The solids are assumed to be homogeneous in temperature, the influence of particle size and thermal conductivity of the material being taken into account via the heat-transfer coefficient. Heat transfer by radiation is considered to be unimportant.

The Reynolds–Nusselt relationships obtained by Beer and Kainer may be used to calculate the heat-transfer coefficients to sinter and to coke. The values found in this way are 233 and 51 W/m² K (200 and 44 kcal/m² °C h), respectively. It is felt, however, that these values are valid only in the case of a strictly homogeneous gas distribution which, in practice, does not exist. Moreover, it is not believed that all of the ore surface nor all of the interstitial space is accessible to the gas flow. The results presented in this paper are obtained by using coefficients of 47 W/m² K (40 kcal/m² °C h) for sinter, and 33 W/m² K (28 kcal/m² °C h) for coke.

It is very difficult to describe the heat transfer to the molten materials in the lower part of the furnace because film thickness, path of descent etc. are unknown. There are three possible ways of heat transfer: coke–liquid, gas–coke, and gas–liquid. The heat-transfer coefficient from gas to coke is the only one which also occurs in equations pertaining to other parts of the furnace, the other two being completely free chosen factors. As only a very small fraction of the total surface area of the coke is wetted by the liquids, heat transfer between coke and liquids will hardly influence the coke temperature and is therefore not taken into account.

The heat-transfer coefficient from gas to liquid has been chosen such that in the reference case a pig-iron temperature of 1570°C is obtained at tuyere level. Owing to heat losses in hearth and runners this corresponds approximately to the observed 1420°C in the
torpedo car. Heat of reaction is subtracted from, or added to, the enthalpy of the component, the temperature of which is rate-determining.

The heats of melting are introduced via the enthalpies of the components. Melting is assumed to take place between 1227 and 1327°C. A factor $f$ is defined to indicate the fraction of a fusible component that is in the fluid state. Heat losses to cooling water have been measured at various heights in the furnace for several of the furnaces at Hoogovens. In the model these heat losses are assumed to be proportional to the local gas temperature. The proportionality coefficient, which is a function of height, has been chosen such that the heat losses calculated for the reference case are in accordance with the actual observations. The heat losses are subtracted from the gas enthalpy.

**Mass transfer**

The following reactions are being taken into account:

$\text{H}_2\text{O}(l) \rightarrow \text{H}_2\text{O}$ \hspace{1cm} (1)

$\text{FeO}_x + \text{CO} \rightleftharpoons \text{FeO}_{x-1} + \text{CO}_2$ \hspace{1cm} (2)

$\text{FeO}_x + \text{H}_2 \rightleftharpoons \text{FeO}_{x-1} + \text{H}_2\text{O}$ \hspace{1cm} (3)

$\text{CO}_2 + \text{C} \rightarrow 2\text{CO}$ \hspace{1cm} (4)

$\text{H}_2\text{O} + \text{C} \rightarrow \text{CO} + \text{H}_2$ \hspace{1cm} (5)

$(\text{FeO}_{0.945}\text{O}) + \text{C} \rightarrow 0.947\text{Fe} + \text{CO}$ \hspace{1cm} (6)

$(\text{MnO}) + \text{C} \rightarrow [\text{Mn}] + \text{CO}$ \hspace{1cm} (7)

$(\text{SiO}_2) + 2\text{C} \rightarrow [\text{Si}] + 2\text{CO}$ \hspace{1cm} (8)

It will be observed that the watergas shift reaction is not included. It is assumed that this reaction has to be catalyzed either by carbon or by Fe/FeO. In the first case, the rate equations for the reactions (4) and (5) will implicitly describe the rate of the shift reaction. In the second case the reaction is described by rate equations (2) and (3). The rate equations for the chemical reactions will be found in the appendix.

**DESCRIPTION OF REFERENCE STATE**

Since work on the dynamic model was started several years ago, the reference state that is now to be described may seem slightly old-fashioned. The geometry of the model (Fig. 2) corresponds to that of Hoogovens BF5 as it was in 1967. Production was about 100 tons per hour. The burden consisted mainly of acid and basic sinter, but in addition small amounts of lump ores, steel slag, and bauxite were charged. Slag weight was 321 kg, and wet coke consumption 450 kg, both per ton of pig iron. The carbon content of the wet coke was 89.6%. Taking the losses into account, 2163 Nm$^3$/min of hot blast containing 9 g/Nm$^3$ H$_2$O and having a temperature of 982°C was blown into the furnace. Injection of heavy fuel oil amounted to 5.36 t/h. Top gas pressure was 0.2 atmospheres. Calculations based on these data led to the results given in Table 1.

**TABLE 1 Calculations based on experiments on BF5 model**

<table>
<thead>
<tr>
<th></th>
<th>BF 5</th>
<th>Model</th>
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<tr>
<td>CO, %</td>
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<td>22.93</td>
</tr>
<tr>
<td>CO$_2$, %</td>
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<td>H$_2$, %</td>
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<td>Temperature, °C</td>
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<td>Pig iron</td>
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<td>C, %</td>
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<td>4.62</td>
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<tr>
<td>Mn, %</td>
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<td>0.84</td>
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<tr>
<td>Si, %</td>
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</table>

A closer approximation of the measured gas analysis could be realized, but in view of the inaccuracy of the blast-furnace input data this was thought to be unimportant.

Temperature and concentration profiles for the reference state are shown in Figs. 1 and 3. Temperatures for the gas-ore pinchpoint, at 9.5 m above the tuyere plane, are 919 and 911°C, respectively. For the gas-coke
TABLE 2 Characteristics of the steady states

<table>
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<tr>
<th>Process characteristics</th>
<th>Experiments*</th>
<th>1</th>
<th>2</th>
<th>3</th>
<th>4</th>
<th>5</th>
<th>6</th>
<th>7</th>
<th>8</th>
<th>9</th>
<th>10</th>
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<td>0.60</td>
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<td>0.61</td>
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<td>Coke consumption</td>
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<td>1.893</td>
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<td>upper limit</td>
<td>m</td>
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<td>307</td>
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<td>CO</td>
<td>%</td>
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<td>24.37</td>
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<td>20.44</td>
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<td>CO₂ dry basis</td>
<td>%</td>
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<td>18.48</td>
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<td>H₂</td>
<td>%</td>
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<td>0.75</td>
<td>4.21</td>
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<td>H₂ utilization</td>
<td>%</td>
<td>41.1</td>
<td>39.8</td>
<td>41.9</td>
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<td>Indirect reduction</td>
<td>%</td>
<td>68.4</td>
<td>63.8</td>
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<td>78.7</td>
<td>77.6</td>
<td>80.5</td>
<td>80.9</td>
</tr>
</tbody>
</table>

* Variations applied are underlined
† At tuyere level
‡ Including extra oxygen and humidity

pinpointed, at 15.7 m above the tuyere plane, temperatures are 776 and 766°C. The hydrogen increase at about 900°C is caused by liberation of the hydrogen content of the coke. It will be observed that there is only the slightest indication of a chemical reserve zone. This is partly due to the fact that, in accordance with observations made in the laboratory, the sinter is supposed not to react with relatively CO₂-rich gas before a temperature of 600 to 700°C has been reached, which leaves a rather short indirect reduction zone. It will be shown later on (Fig. 14) that the chemical reserve zone becomes more pronounced if the furnace stack is hotter.

On the other hand, the burden components have different reducibilities. The position of the FeO level for acid sinter is 1.5 m closer to the tuyere plane than that for basic sinter. Figure 14 demonstrates that an increase in reducibility of the acid sinter has an effect on the CO₂ profile.

**INJECTION OF FUEL OIL**

Partial replacement of metallurgical coke by auxiliary fuels is now a well-established practice. Therefore, some model experiments were conducted involving the use of various amounts of fuel oil, eventually combined with oxygen enrichment of the blast or increased blast temperature. In these cases, the dynamics were thought to be of minor importance compared to the steady-state characteristics. To save computer time, the paths along
which these steady states had to be reached were made as short as possible, using means that are not at the disposal of the blast-furnace operator. As a consequence, only the characteristics of the steady states are presented (Table 2).

Table 3 shows the coke savings obtained by injection of fuel oil without the use of extra oxygen or increased blast temperature. The pig-iron silicon content is 0.6%.

One of the consequences of oil injection is a lower pig-iron temperature. This temperature, as well as the silicon content, depends on the time available for heat and mass transfer from gas to liquid, and thus on the distance between tuyeres and melting zone. A relatively small part of the nitrogen has been removed from it.

If the original silicon content is to be maintained some extra fuel is required (Table 2, column 10). It appears that this brings the indirect reduction back to its original level, so that the pig-iron temperature increases even if the flame temperature decreases. Contrary to the expectations a slight fuel saving was found when oxygen enrichment was used in combination with very high amounts of fuel oil (Table 2, columns 8, 9, 10). It appears that in these cases the indirect reduction actually increases. The data suggest an optimum in fuel economy somewhere around 5% of extra oxygen. More work is needed to confirm this.

### Table 3: Coke savings obtained by injection of fuel oil

<table>
<thead>
<tr>
<th>Oil, kg/THM</th>
<th>Coke, kg/THM</th>
<th>Integral replacement ratio</th>
</tr>
</thead>
<tbody>
<tr>
<td>50</td>
<td>518</td>
<td>-</td>
</tr>
<tr>
<td>100</td>
<td>448</td>
<td>1.40</td>
</tr>
<tr>
<td>200</td>
<td>388</td>
<td>1.30</td>
</tr>
<tr>
<td>200</td>
<td>281</td>
<td>1.18</td>
</tr>
</tbody>
</table>

A general impression of the relationships between pig-iron temperature, silicon content, flame temperature, and level of melting zone is given in Fig. 4, which was constructed using all available data. Even so, no claims for accuracy are being made. It appears that the silicon content depends mainly on the melting-zone level. Conversely, the flame temperature is the predominant factor in determining the pig-iron temperature. The flame temperature may be raised by way of a higher hot-blast temperature. A comparison of the columns 6, 7 and 8 in Table 2 shows that, although a nice coke saving is obtained, the effect on the pig-iron temperature is rather small. For the rest, most plants will not have much room left for a substantial increase of the hot-blast temperature.

Oxygen enrichment of the blast is another way to increase the flame temperature. Columns 3 and 4 in Table 2 show that oxygen enrichment from 21 to 24% results in a slightly cooler furnace hearth. The pig-iron temperature decreases in spite of a higher flame temperature. This is brought about by a change in the temperature profile in the stack. As a consequence, the indirect reduction zone is shortened. More carbon is consumed for endothermic FeO reduction and less is available for combustion. Besides, less heat is imported with the hot blast because a substantial part of the nitrogen has been removed from it.

### Dynamic Responses of the Furnace

The dynamic behaviour of the blast-furnace process was studied by applying stepwise increases or decreases in one of the independent variables. To avoid confusion it is pointed out that, for instance, the amount of oil injected per hour is an independent variable, whereas the oil consumption per ton of pig iron is not. Likewise, the amount of coke per ton of iron in ore can be varied independently, whereas the coke consumption per ton of hot metal depends on the pig-iron composition. This means that the end effect of, for instance, a step in blast enthalpy has to be corrected for variations in fuel consumption.

### Variations in Common Blast-furnace Practice

The results of calculations in which respectively coke consumption, oil rate, hot-blast temperature, blast moisture, and blast rate were changed stepwise are summarized in Table 4, columns 11 to 15 and Figs. 5 to 9. In these figures a logarithmic time scale has been used. The short horizontal lines at left indicate the levels of the dependent variables at time zero. It appears that, especially in the case of the fuel steps, it takes a very long time before the new steady state is attained. In our opinion, actual blast-furnace trials might easily lead to different conclusions because, owing to the ever-present scatter in silicon, it is very hard to decide at what moment the experiment may be considered as finished.

In the case of the oil step the dynamic behaviour of the furnace is characterized by an initial chilling due to the decomposition of the oil. It takes some hours before the effect of the increased indirect reduction becomes noticeable in the hearth. The calculated increase in pig-iron silicon due to an increase in coke consumption may be somewhat larger than commonly found in practice.
5 Responses to a step in coke charged of +30 kg per ton of pig iron

7 Responses to a step in blast temperature of +100°C

6 Responses to a step in oil rate of 50%; oil consumption increased from 54 to 91 kg/THM

8 Responses to a step in blast moisture of +10 g/m³
literature. We are under the impression, however, that the experimental evidence concerning the Si–coke relationship (60 to 70 kg coke per % Si) has not been renewed for several years. Since then, the value of several items on the furnace heat and mass balances may have changed considerably. For instance, the amount of reducing gas available per ton of pig iron is much smaller, and the CO/CO₂ ratio in top gas decreases steadily. Accordingly, if pig iron with a high silicon content is to be produced, a larger fraction of the additional CO generated by combustion of the extra coke will participate in the ore reduction, thereby lowering the incremental heat requirements.

In the calculations, pig iron with a 0.77% higher silicon content was obtained at the cost of an increment in total fuel consumption of 31.2 kg, expressed as coke with 89.6% C. This means that 40.5 kg of coke was needed per % of silicon. An additional 2 kg of coke would have been required if it had been attempted to maintain a constant slag composition. To obtain the coke equivalent of each step use was made of the coke-silicon relationship calculated above. Coke-oil replacement ratios were derived from Table 3.

The effect of the variation in blast temperature may be compared with that found in the experiments 12 and 13: 16.8 resp. 13.1 kg coke/100°C. Figure 9 shows that the well-known rapid heating of a furnace following a

TABLE 4 Results of variations in common blast-furnace practice

<table>
<thead>
<tr>
<th>Process characteristics</th>
<th>Experiments*</th>
<th></th>
<th></th>
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<td>12</td>
<td>13</td>
<td>14</td>
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<td>5.36</td>
<td>5.09</td>
<td>5.36</td>
<td>5.36</td>
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<td>9</td>
<td>19</td>
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<td>oxygen %</td>
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<td>21.0</td>
<td>21.0</td>
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<td>2199</td>
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<td>Indirect reduction %</td>
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<td>67.9</td>
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<td>Coke equivalent of step kg/t pig iron</td>
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<td>+12.1</td>
<td>-50</td>
<td>67.9</td>
<td>69.2</td>
<td>70.1</td>
</tr>
</tbody>
</table>

* Variations applied are underlined
† At tuyere level
‡ Including extra oxygen and humidity
Influence of change in flame temperature on model response to extra heat

decrease in blast rate cannot be reproduced by the model. It is concluded that, in addition to the assumptions made in the model, other hypotheses would be necessary to explain this phenomenon. So far, no attempts in this direction have been made.

It remains to be mentioned that the dynamic behaviour of the furnace depends to some degree on the change in flame temperature. In Fig. 10 the relations between pig-iron temperature and silicon content are shown for the experiments 11 and 13. Points on the curves are spaced half an hour apart. In experiment 11 there is virtually no change in flame temperature. As soon as the extra coke arrives at the tuyeres pig-iron temperature and silicon start to increase simultaneously.

Experiment 13, on the other hand, is characterized by a sudden increase of the flame temperature which, as is shown in Fig. 4, has a predominant influence on the pig-iron temperature. Consequently, the rise in pig-iron temperature during the first hour is more important than the increase in silicon content.

Effect of process disturbances

The independent variables mentioned thus far can be measured and controlled. There are others, however, which are much harder to control. One of these is the gas distribution over the furnace cross-section. In the model, which is one-dimensional, simulation of variations in the gas-distribution pattern is rather difficult. It is possible, however, to simulate the resultant changes in the average values of the heat and mass transfer coefficients. It is felt that there must be some similarity between the effects of a change in gas-distribution pattern and a change in burden reducibility, in itself one of the factors which is difficult to control.

Figures 11 to 13 show the effect of a 20% increase in reducibility starting from the steady states 1, 6 and 8 indicated in Table 2. In these cases one of the old ores was replaced by a new ore having a 56% higher reducibility. It appears that the responses are very slow, especially when the fuel oil rate is high.

The gas curves show an initial increase in indirect reduction. When the new ore approaches the melting zone gas production by direct reduction is diminished. A declining gas volume causes the top-gas temperature to decrease until the old ore has disappeared from the direct reduction zone. Gas production increases again when the coke saved in the direct reduction zone reaches the tuyeres. At the same time, indirect reduction starts to slowly decrease because the rising melting zone
diminishes the space available for this reaction. The magnitude of this effect, of course, depends upon the initial level of the melting zone.

CO₂ profiles for the steady states 1, 6, and 8, and for the experiments with increased reducibilities, are shown in Fig. 14. In the experiments 6 and 8 the stack is hotter than in experiment 1, due to the increased amount of reducing gas. Consequently, reduction starts earlier and more room is left for a chemical reserve zone. This zone, of course, is more distinct when the ore has a better reducibility. Figure 15 shows the consequences of an interruption of the fuel oil injection. Nothing seems to happen at first, but after about three hours the furnace starts to chill rapidly. Oil injection was gradually resumed after six hours, and the initial injection rate of 9-4 tons per hour (96 kg per ton of pig iron) was attained after eight hours. The reheating of the furnace is a very slow process. This simulation was performed following an incident at one of the IJMuiden blast furnaces caused by interruption of the oil flow. This led to a chilled furnace in about the same time as indicated by the calculation.

Finally, some calculations were performed concerning cooling water leakage into the lower part of the furnace. The steady states 1, 6 and 8 served again as starting points. The amount of water was 3-25 tons per hour, which led to more than one percent of extra hydrogen in top-gas. As a leakage of this size will certainly not be tolerated for more than a few hours, it did not seem to make much sense to prolong the calculations too far. Consequently, the silicon curves in Fig. 16 are plotted against a linear time scale.

It appears that there is not much difference between the silicon response curves. However, pig-iron temperatures at tuyere level after 12 h were 1 501°, 1 443°, and 1 475°C, respectively. It is concluded that, as far as water leakage is concerned, low flame temperatures do necessarily lead to earlier freezing of the furnace.

**Limitations of the model**

As follows from the basic assumptions, the use of the model is limited to experiments in which modifications in gas, liquid and solid flow distribution over the furnace cross-section play a minor role. It should be remarked that fundamental knowledge pertaining to gas and liquid dynamics in the lower part of the furnace, including the deadman, is still far from complete. In our opinion, substitution of gaps in knowledge by hypotheses based on intuition should not be carried too far.

**Execution of the calculations**

Use is made of a Univac 1108 computer. It takes about five minutes of calculation time to simulate one day of blast-furnace practice. The program occupies about 16 K core space (1 K = 1 024 words of 36 bits).

**ACKNOWLEDGEMENT**

The authors are grateful to Mr H. Veldman for the important part he took in the development of the Fortran program.
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APPENDIX
Evaporation of coke moisture
The saturation pressure of water vapour can be approximated as follows:
\[ p^*H_2O = \exp(-5098/TC + 13.67) \text{ atmospheres} \]
The partial water vapour pressure being
\[ pH_2O = P_{TOP} \times H_2O/GAS \]
the following simple rate equation will suffice:
\[ V_1 = g_1 \cdot H_2O(l) \cdot (p^*H_2O - pH_2O) \]
The rate constant \( g_1 \) has been chosen such that all of the water evaporates in the top cells. Its value is 0·1.

Gaseous reduction of iron oxides
\[ V_2 = g_2 \cdot Fe(pCO - K_2 \cdot pCO_2) \cdot \exp(-E_2/R \cdot TO) \times (1 - f) \]
\[ V_3 = g_3 \cdot Fe(pH_2 - K_3 \cdot pH_2) \cdot \exp(-E_3/R \cdot TO) \times (1 - f) \]
The rate constants \( g_2 \) and \( g_3 \) are functions of the actual value of \( x = O/Fe \). The equilibrium constants \( K_2 \) and \( K_3 \) are functions of both temperature and \( x \). The numerical values of \( g, K \) and \( E \) for different ores have to be derived from laboratory experiments.

Gaseous oxidation of coke carbon
\[ V_4 = g_4 \cdot C(pCO_2 - K_4 \cdot p^2CO) \cdot \exp(-E_4/R \cdot TO) \]
\[ V_5 = g_5 \cdot C(pH_2O - K_5 \cdot pCO \cdot pH_2) \cdot \exp(-E_5/R \cdot TO) \]
The values of the rate constants \( g_4 \) and \( g_5 \) are determined by the coke reactivity which, in its turn, depends upon the coke size, and on catalytic effects. The values used for the rate constants \( g_4 \) and \( g_5 \) are \( 2.8 \times 10^5 \) and \( 4.2 \times 10^4 \). The activation energies are 50 and 40 kcal per mole, respectively.

Reduction of liquid FeO
It is assumed that liquid FeO is reduced by solid carbon only, according to the following simple rate equation:
\[ V_6 = g_6 \cdot O \cdot \exp(-E_6/R \cdot TO) \cdot f/u \]
A value of 30 kcal/mol has been assumed for \( E_6 \). The value of \( g_6 \) is chosen such that all FeO is reduced before tuyere level is reached. This requirement is satisfied by \( g_6 = 10 \).

Reduction of MnO
MnO is supposed to be reduced from the slag phase by solid carbon.
\[ V_7 = g_7 \cdot MnO \cdot \exp(-E_7/R \cdot TO) \cdot f/u \]
in which \( g_7 \) equals \( 1.43 \times 10^5 \). A value of 50 kcal/mol was assigned to \( E_7 \).

Reduction of silica
It is assumed that silica reduction proceeds to completion above the tuyere zone. Oxygen transport in the metal phase is considered to be the rate-determining step. The following rate equation, which may look somewhat ambitious in view of the simplicity of the preceding ones, was derived:
\[ V_8 = g_8 \cdot F[(aSiO_2/\alpha Si)^{0.4} - K_8 \cdot pCO] \cdot \exp(-E_8/R \cdot TO)/u \]
where \( F = (\text{kmol of slag times kmol of iron}^{0.5}) \)
\( aSiO_2 = 0.1(\text{SiO}_2/\text{kmol of slag}) \)
\( \log aSi = \log(\text{Si}/\text{Fe}) + 4.45 \times 10^{-4} \cdot TO - 2.67 \)
\( K_8 = \exp(42850/TO - 22.1) \)
\( g_8 = 3.5 \times 10^8 \)
\( E_8 = 80 \text{ kcal/mol} \)