Ilmenite as oxygen carrier in a Chemical Looping Combustion system with coal

Ana Cuadrat*, Alberto Abad, Francisco García-Labiano, Pilar Gayán, Luis F. de Diego, Juan Adánez

Department of Energy and Environment, Instituto de Carboquímica (CSIC), Miguel Luesma Castán 4, 50018 Zaragoza, Spain

Abstract

Chemical-Looping Combustion, CLC, is one of the most promising technologies to capture at a low cost the CO$_2$ in fossil-fuelled power plants. CLC is based on the transfer of the oxygen from air to the fuel by using a solid oxygen-carrier that circulates between two interconnected fluidized reactors: the fuel- and the air-reactor. The CO$_2$ capture is inherent to this process, as the air does not get mixed with the fuel. The selection of the oxygen-carrier is a key factor for the CLC technology development. Ilmenite is a low cost natural mineral which is promising for its large scale industrial use as oxygen-carrier with solid fuels. In this work, the CLC technology with a Colombian bituminous coal was investigated in a continuous rig using ilmenite as oxygen-carrier. The plant was basically composed of two interconnected fluidized-bed reactors joined by a loop seal, a riser for solids transport from the air- to the fuel-reactor, a cyclone and a solid valve to control the solids flow rate. This work is focused on the study of the fuel-reactor within the process. The behavior of the fuel-reactor has been evaluated. The effect of the temperature and coal particle size was investigated, on the combustion efficiency as well as on the extent of gasification in the fuel-reactor. The experiments were done at 820-950°C in the fuel-reactor. A particular feature of this plant is that the solids circulation rate can be also controlled. The fluidizing gas was water steam, which acts also as gasifying agent. Since this plant does not have carbon stripper, it is really focused on the study of the gasification and combustion of coal in the FR within the CLC process. The results obtained are valid for the scale-up of a CLC process fuelled with coal. The results show that ilmenite has a good behavior as oxygen-carrier and that optimization of the CLC technology with solid fuels can lead to energy production with high CO$_2$ capture efficiencies.

Keywords: Chemical-looping combustion (CLC); Oxygen carrier; Ilmenite; Carbon capture; Coal

1. Introduction

Chemical-Looping Combustion (CLC) is one of the most promising technologies to carry out CO$_2$ capture in power plants at a low cost [1-3]. A key factor for the CLC technology development is the selection of the oxygen-carrier. Suitable oxygen-carriers for a CLC process must have high reactivity during reduction and oxidation, high attrition resistance and agglomeration absence. Low cost of the material is rather desirable for its use with coal, as it
is predictable a partial loss together with the coal ashes when removing them from the reactor to avoid their accumulation in the system. Ilmenite is a low cost natural mineral which is promising for its large scale industrial use as oxygen-carrier with solid fuels. It is mainly composed of FeTiO₃ (FeO•TiO₂), where iron oxide is the active phase that behaves as the oxygen-carrier. Performance of ilmenite has been proven to be acceptable as oxygen carrier for CLC in recent studies made at different scales. On the whole, ilmenite shows adequate values of reactivity and oxygen transport capacity for its use in the CLC technology with solid fuels [4,5]. Ilmenite was also successfully tested in a 10 kWth chemical-looping combustor using South African coal and petroleum coke as solid fuels by Berguerand et al. [6,7].

In this work, the CLC technology with a Colombian bituminous coal was investigated in a continuous rig using ilmenite as oxygen-carrier. The aim of this work was to investigate the effect of the FR operation variables on the CLC performance. The effect of the temperature of the fuel reactor and coal particle size as operating conditions on the combustion efficiency as well as on the extent of gasification in the FR, O₂ polishing needs and volatile combustion were investigated. The experiments were carried out at 820-950ºC in the FR. The fluidizing gas was 100% steam, which also acts as gasifying agent. The results obtained are analyzed and discussed in order to be useful for the scale-up of a CLC process fuelled with coal.

2. Experimental section

2.1. Coal and ilmenite

The fuel used was a bituminous Colombian coal “El Cerrejón”. Bed agglomeration problems with pipes clogging were found when El Cerrejón coal was fed into the system because this coal showed a high swelling behavior. In order to reduce the swelling properties, the coal was subjected to a thermal pre-treatment for pre-oxidation. Coal was placed in trays in layers of about 3mm height and exposed to heating at 180ºC in air atmosphere for 28 hours. Elementary and proximate analyses of the pre-treated coal are shown in Table 1. The pre-oxidation causes an increase in O content and a decrease in the heating value. Despite this drop, the swelling properties of this coal were eliminated and no agglomeration problems occurred in any experiment. Three different coal particle sizes were used: +74-125, +125-200 and +200-300 μm.

<table>
<thead>
<tr>
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<td>65.8 %</td>
<td>2.3 %</td>
<td>33.0 %</td>
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<tr>
<td>H</td>
<td>3.3 %</td>
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<td>55.9 %</td>
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<tr>
<td>N</td>
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<td>S</td>
<td>0.6 %</td>
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<tr>
<td>O</td>
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Table 1  Properties of the used pre-treated Colombian coal.

The bed material was ilmenite with particle size of +150-300 μm. The oxidized species present in ilmenite are Fe₂TiO₅, Fe₂O₃ and TiO₂, and in its reduced state it is mainly composed of FeTiO₃, Fe₃O₄ and TiO₂. The initial oxygen transport capacity was measured to be 4%. More details about the composition and physical properties of ilmenite particles can be found elsewhere [4,8]. The ilmenite introduced was pre calcined at 950 ºC and has a relatively slow initial reactivity, but it activates during operation after various red-ox cycles undergone in the FR-AR system [4,8]. The activation of ilmenite was carried out and was in the continuous plant evaluated. However the experiments have been carried out when ilmenite was already activated for reduction, so that the reactivity of ilmenite was maximum and constant and did not affect the evaluation of other parameters.

2.2. CLC reactor

A schematic view of the experiment facility is shown in Figure 1. The CLC system was basically composed of two interconnected fluidized-bed reactors joined by a loop seal (2), a riser (4) for solids transport from the AR (3) to the FR (1), a cyclone (5) and a solids valve (7) to control the flow rate of solids fed to the FR. This design allowed
the variation and control of the solid circulation flow rate between both reactors by the use of the mentioned solids valve. The temperatures in the different sectors of the plan can be set, as it is heated up by various furnaces (10).

![Schematic diagram of the coal-fuelled CLC facility.](image)

Figure 1 Schematic diagram of the coal-fuelled CLC facility.

The FR consisted of a bubbling fluidized bed with 5 cm of inner diameter and 20 cm bed height. Coal (8) is fed by a screw feeder at the bottom of the bed above the FR distributor plate in order to maximize the time that volatile matter is in contact with the bed material. The screw feeder (9) has two steps: the first one with variable speed to control the coal flow rate, and the second has high rotating velocity to avoid coal pyrolysis inside the screw. A small N\textsubscript{2} flow is fed in the beginning of the screw to avoid possible volatile reverse flow or entrance of steam. In the FR the oxygen-carrier is reduced by the volatile matter and gasification products of coal. Reduced oxygen-carrier particles overflowed into the air-reactor through a U-shaped fluidized bed loop seal with an inner diameter of 50 mm, to avoid gas mixing between fuel and air. The absence of a carbon stripper facilitates the interpretation of the effect of these operational conditions on the results obtained. The oxidation of the carrier took place in the AR, consisting of a bubbling fluidized bed with 8 cm of inner diameter and 10 cm bed height, and followed by a riser. Secondary air was introduced at the top of the bubbling bed to help particle entrainment. N\textsubscript{2} and unreacted O\textsubscript{2} left the AR and went through a high-efficiency cyclone and a filter before the stack. The oxidized solid particles recovered by the cyclone were sent to a solids reservoir, setting the oxygen-carrier ready to start a new cycle. These particles act as a loop seal avoiding the leakage of gas between the FR and riser. The regenerated oxygen-carrier particles returned to the FR by gravity from the solids reservoir through a solids valve which controlled the flow rates of solids entering the FR. A diverting solids valve located below the cyclone allowed the measurement of the solids flow rates at any time. The total ilmenite inventory in the system was 3.5 kg and the solids inventory in the FR was 1.5 kg ilmenite.

In this facility coal is physically mixed with the oxygen-carrier in the FR and the carrier reacts with the gas product of the devolatilization and gasification of coal, where H\textsubscript{2} and CO are the main components. The FR is fluidized by steam, which acts also as a gasifying agent. The solid fuel gasification proceeds in the FR first and the resulting gases and volatiles are oxidized through reduction of the oxidized ilmenite, represented as Me\textsubscript{x}O\textsubscript{y}, by means of reaction (1). The ilmenite reduced in the FR, represented as Me\textsubscript{x}O\textsubscript{y-1}, is transferred to the AR where reaction (2) with oxygen from air takes place. Thus, ilmenite is regenerated to start a new cycle. Since this rig has no carbon stripper, unconverted char from the FR goes to the AR and is fully burnt there, releasing the CO\textsubscript{2} that is measured in the AR. The implementation of a carbon stripper would increase the carbon capture and therefore no CO\textsubscript{2} would be generated in the AR.

\[
\begin{align*}
\text{H}_2, \text{CO, Volatile matter (CH}_4) + n \text{ Me}_x\text{O}_y & \rightarrow \text{CO}_2 + \text{H}_2\text{O} + n \text{ Me}_{x-o} \text{O}_{y-1} \\
\text{Me}_{x-o} \text{O}_{y-1} + \frac{1}{2} \text{O}_2 & \rightarrow \text{Me}_x\text{O}_y
\end{align*}
\]
In the FR the steam introduced has a velocity of 11 cm/s at 900ºC and in the AR the stream of air has a velocity at 900 ºC of 60 cm/s. The circulation flow rate was measured and controlled to be about 42 g/h, corresponding to a thermal power of 255 Wth.

Different experiments were carried out varying the temperature and using different particle size of coal. The temperature was varied in the range 820 - 950 ºC. 30 different stable conditions were reached, at least 30 minutes each one. A total of 26 hours of continuous operation feeding coal and 35 hours of continuous fluidization were made.

During operation, temperatures in the bed and freeboard of the FR, AR bed and riser were monitored as well as the pressure drops in important locations of the system, such as the FR bed, the AR bed and the loop seal. The temperature in the AR was maintained at around 950ºC and the FR temperature was varied from 820 ºC to 950 ºC. Because of its small size, the system is not auto-thermal and is heated up with various ovens to get independent temperature control of the AR, FR, and FR freeboard. The FR freeboard is kept constant at about 900ºC in all the experiments.

CO, CO2, H2, CH4, and O2 were continuously analyzed in the exit streams from the FR and from the AR. In some selected experiments the tar amount present in FR product gases was determined following the tar protocol. In all the experiments the gas composition at the exit gases of FR and AR were analyzed and the oxygen demand of FR gases were calculated.

3. Results and discussion

3.1. Effect of the FR temperature

The influence of the FR temperature on the main parameters of the CLC process for different coal particle sizes was studied. As representative of the gas distributions obtained in this study, Figure 2 shows the evolution of the produced gas distributions in the FR as well as in the AR for the particle size of +125-200 µm for different FR temperatures when it was increased from 820 to 950 ºC. The concentrations in FR are in dry basis and N2 free. Stationary states after any parameter change were fast reached and all the points were therefore evaluated at stable conditions. The outlet of the FR is mainly composed of oxidized CO2, and H2 and CO as not fully oxidized products of char gasification. It is remarkable to say that there is little CH4 measured. As mentioned, the measured CO2 in the AR is due to the unreacted char flowing towards the AR due to the absence of a carbon stripper.

![Figure 2](image.png)

Figure 2  Gas distributions in FR and AR for increasing FR temperature. Solids rate: 3.5 kg/h. Coal particle size: +125-200 µm.

Besides, for two experiments at constant conditions for longer than one hour, tar measurements in the FR were done using tar protocol. The results showed that there were no tars in the FR outlet flow, that is, no hydrocarbons heavier than C5. Black solid particles, which were likely to be char particles, were recovered in the first liquid container of the tar condensation train. That indicated that some fraction of char was elutriated from the FR. Later GC measurements proved also that there were no C2-C4 volatiles in the FR. In this context, some assumptions were made. First was to calculate the elutriated char from the FR as the difference between the fed coal carbon and the measured carbon in the FR and AR gases. Second was to consider that all the hydrocarbons from the pyrolysis of the introduced coal are burned in the FR by ilmenite. Third, the product gas is only composed by CH4, CO2, CO, H2 and H2O. Nevertheless, CH4 concentration was always below 2% in dry and N2 free basis.
Figure 3 shows the variation with the temperature of the concentrations of CO$_2$, CO and H$_2$ at the outlet of the FR in dry and free N$_2$ basis for the different particle sizes. The gasification and combustion reactions are faster and promoted at higher temperatures, and therefore more CO$_2$ and H$_2$O as products of combustion are generated, with the corresponding decrease of CO and H$_2$.

The efficiencies that indicate the performance of the process are defined as follows. The gasification efficiency, $\eta_{\text{gasif}}$, is the fraction of the carbon of the effective coal that is converted to gas. The carbon content of the FR product gas is calculated as the sum of the flows of the carbon containing species in the FR, that is, $F_{\text{CO}_2,\text{FR}}$, $F_{\text{CO},\text{FR}}$, $F_{\text{CH}_4,\text{FR}}$.

$$\eta_{\text{gasif}} = \frac{F_{\text{CO}_2,\text{FR}} + F_{\text{CO},\text{FR}} + F_{\text{CH}_4,\text{FR}}}{C_{\text{coal eff, in}}} = \frac{F_{\text{CO}_2,\text{FR}} + F_{\text{CO},\text{FR}} + F_{\text{CH}_4,\text{FR}}}{(F_{\text{CO}_2,\text{FR}} + F_{\text{CO},\text{FR}} + F_{\text{CH}_4,\text{FR}} + C_{\text{char to AR}})} \quad (3)$$

The gas flows were calculated by multiplying the corresponding gas fraction and the outlet gas flow, $F_{\text{FR, outlet}}$, for the FR. The dry basis product gas flow, $F_{\text{FR}}$, was calculated by using the N$_2$ flow $F_{\text{N}_2,\text{FR}}$ that is introduced in the FR coming from the Loop Seal. Preliminary results showed that 30-40% of the N$_2$ introduced to fluidize the Loop Seal went to the FR and the rest to the AR. The outlet AR gas flow, $F_{\text{AR}}$, was calculated through the introduced N$_2$, $F_{\text{N}_2,\text{AR}}$, which is sum of the N$_2$ present in the air introduced in the AR and the N$_2$ coming from the Loop Seal.

$$F_{\text{FR}} = F_{\text{N}_2,\text{FR}}/[1-[\text{CH}_4]_{\text{FR}}+[\text{CO}_2]_{\text{FR}}+[\text{CO}]_{\text{FR}}+[\text{H}_2]_{\text{FR}}] \quad (4)$$

$$F_{\text{AR}} = F_{\text{N}_2,\text{AR}}/[1-[\text{CO}_2]_{\text{AR}}+[\text{O}_2]_{\text{AR}}] \quad (5)$$

The combustion efficiency in the FR is defined as the fraction of the oxygen demanded by the effective coal fed that is supplied by the oxygen carrier in the FR, O$_2$ from ilmenite. It is a measure of gas conversion and represents the extent of oxidation of volatiles and gasification products by the oxygen carrier.

$$\eta_{\text{comb}} = \frac{O_2 \text{ from ilmenite}}{O_2 \text{ demand coal eff in}} = \frac{0.5 (F_{\text{H}_2\text{O,FR}} - F_{\text{H}_2\text{O,in}}) + F_{\text{CO}_2,\text{FR}} + 0.5 F_{\text{CO,FR}} - 0.5 [O]_{\text{coal eff in}}}{O_2 \text{ demand coal eff in}} \quad (6)$$

$F_{\text{H}_2\text{O,gen,FR}}$, $F_{\text{CO}_2,\text{FR}}$ and $F_{\text{CO,FR}}$ being, respectively, the FR outlet flows of water, CO$_2$ and CO. $[O]_{\text{coal eff in}}$ is the amount of oxygen present in the used coal minus the oxygen contained in the elutriated char. O$_2$ demand coal eff in or the oxygen demand of coal is the number of moles of oxygen needed to burn the fuel completely and is calculated through the elemental analysis shown in Table 1. Figure 4 represents the efficiencies of gasification and combustion as a function of the reactor temperature for different coal particle sizes. As it was presumable, the results show that there is a continuous increase of efficiency with the temperature. For all coal particle sizes this raise is moderate until 910°C with a sharp increase in both efficiencies at higher temperatures. From 880 °C the gasification efficiency is higher than 38% and reaches a value of 86% at 950 °C. The combustion efficiency varies from 37% at 880°C to 92% at 950°C. At the lower temperature the gasification efficiency is close to the volatile matter content of the coal, see Table 1. This means that most of carbon in the gases comes from the volatiles, and few amounts of char are being gasified. The increase in the gasification efficiency with the temperature is due to more carbon in char is being gasified in the reactor. If the trends are extrapolated, it could be expected that all efficiencies would reach a value close to 100% at 1000°C, which means that most of carbon in the coal would exit with the FR flue gases.
Note that in this rig and most notably below 920 °C, char gasification presents low values. This is because the gasification rate at these temperatures is slow and thus a relatively high amount of char passes to the AR. In the big scale power plant this is not going to happen with the implementation of a carbon separation system. In this case, the char concentration in the FR should increase until the new steady state was reached.

An oxygen demand for the FR gases, $\Omega g$, was defined, as the fraction of oxygen lacking to achieve a complete combustion of the FR product gas in comparison to the oxygen demand of the effective introduced coal, $O_2$ demand coal$_{eff}$ in. This parameter is the most adequate to evaluate the performance of the combustion process in this facility.

$$\Omega g = \frac{O_2 \text{ demand gases FR}}{O_2 \text{ demand coal}_{eff} in} = \frac{[2 \cdot F_{CH_4,FR} + 0.5 \cdot F_{H_2,FR} + 0.5 \cdot F_{CO,FR}]}{O_2 \text{ demand coal}_{eff} in} \quad (7)$$

$F_{CH_4,FR}$, $F_{H_2,FR}$ and $F_{CO,FR}$ being the flows in the FR of CH$_4$, H$_2$ and CO. These flows were calculated from the CH$_4$, H$_2$ and CO concentration at the FR exit. The unburnt gases, CO, H$_2$ and CH$_4$, were on-line analyzed. The carbon measured in the gases coming from the FR and the AR is less than the carbon present in the introduced coal, because there is elutriation of char. Thus, the oxygen demand of the effective coal was calculated as the oxygen demand of coal minus the oxygen demand of the elutriated char, since this is actually the coal involved in the conversion process in the FR.

Figure 5 shows the oxygen demands due to the unburnt gases present in FR, $\Omega g$ at the different temperatures and the coal particle sizes used. Regarding the temperature, it can be seen that up to approximately 900°C, $\Omega g$ decreases because of the increase in ilmenite reaction rate with the temperature. However, from 910 °C a slight trend of increase with the temperature can be seen, which coincides with the sharp increase of gasification and combustion efficiencies, afterward shown. This can be explained through the raise of the devolatilization and gasification rate: there is more production of the gasification products, H$_2$ and CO, which is not fully compensated by the increase in the ilmenite reaction rate.

The calculated char concentration in the FR bed for all the experiments was about 0.35% and the oxygen demands and efficiencies here obtained are comparable and applicable to an optimized CLC power plant that would have the same char concentrations in the FR bed.
3.2. Effect of the coal particle size

The effect of the coal particle size on the process performance was investigated because it is a key parameter in the operation of fluidized-bed reactors. As it can be seen in Figure 5, the particle size does not seem to affect to the combustion efficiencies, and minor differences were found for the gasification efficiencies. However, the oxygen demand depended on the particle size. Figure 5 showed that the oxygen demand of gases, \( \Omega_g \), increased as the particle size was decreasing. The amount of elutriated char was calculated from the differences observed between the carbon fed with the coal and the measured in the FR and AR gas streams. As expected, smaller particles are more easily elutriated than bigger particles and for particle sizes of +74-125 \( \mu \text{m} \) about 40% of the introduced char has been elutriated, whereas about 10-20% of the char was entrained for bigger particles.

The higher values calculated for the gasification efficiency for the smaller particles in this unit are related to the relatively high amounts of CO and H\(_2\) in the gas product from FR, showed in Figure 3. Unconverted gaseous products, i.e., CO, H\(_2\) and CH\(_4\), can come from a partial oxidation of the gases in the FR, as well as char gasification in the FR freeboard. The high values obtained for elutriated char with the lower coal particle size cause a higher fraction of char in the freeboard for smaller particles. The result is therefore that gasified fuel particles in the freeboard have not been in contact with the ilmenite particles that remain at the bottom bed and thus production of CO and H\(_2\) is higher for the smaller coal size. Thus, the gasification efficiency increases for the particle size of +74-125 \( \mu \text{m} \) (see Figure 4) but the combustion efficiency was constant because the gasification products in the freeboard are not burnt. Indeed, the oxygen demand of gas, \( \Omega_g \), also increases when the particle size decreases, see Figure 5.

3.3. Activation of ilmenite in the continuous rig

Solids samples were extracted during the continuous experiments at several operating times. Since the initial carrier is calcined ilmenite, an activation process is expected to happen with the number of the red-ox cycles.

Reactivity of ilmenite samples taken after 1, 3, 4, 15, and 26 hours of operation was determined by TGA at 900°C with 5% H\(_2\)+40% H\(_2\)O mixtures, and afterwards to oxidation conditions in air. Figure 6 shows the conversion, \( X_{\text{ilm}} \), vs. time curves obtained for the reduction and oxidation for the different samples and for calcined ilmenite. The activation with the number of hours undergone by ilmenite can be seen. It can be considered that ilmenite is already active after 3 hours for the reduction reaction, although the reactivity increases further in minor extension after 15 h.

![Figure 6: Ilmenite conversion vs. time for a) reduction and b) oxidation for calcined ilmenite and for ilmenite samples after 1, 3, 4, 15 and 26 hours of continuous operation.](image)

Nevertheless, ilmenite is not completely activated for oxidation reaction after 26 hours operation yet. However, after 15 hours operation and up to an oxidation conversion about 50%, the reaction rate is the same as for the completely activated ilmenite. This is because the second part of the reaction mechanism for oxidation of ilmenite is controlled by diffusion and this reaction step disappears with the development of porosity in the particle [8]. The number of cycles to activate ilmenite increases as the variation of ilmenite conversion in successive redox cycles decreases [4]. The average change in ilmenite conversion in the experiments was low, \( \Delta X_{\text{ilm}} = 23\% \), and therefore a full activation could not happen yet.

The oxygen transport capability of ilmenite for samples taken at different operation times was determined by TGA. The oxygen transport capability was maintained roughly constant with the operation time. A slightly decrease from 4.0% for calcined ilmenite to 3.9% for particles used during 26 h in the unit was observed.
4. Conclusions

The effect on the FR performance of the temperature and the coal particle size were studied in a continuously operated CLC rig. Ilmenite was used as oxygen carrier and the bituminous Colombian coal “El Cerrejón” as fuel. The performance of the CLC system using a solid fuel was analyzed at several temperatures of the FR and various particle sizes of coal. At higher FR temperatures, gasification and combustion reactions are faster and promoted. Gasification and combustion efficiencies grow with the temperature, with a faster increase at T > 910 °C. Although the carbon capture efficiency in this rig is not representative for the CLC technology, since it does not have carbon stripper, high carbon capture efficiencies are expected to be obtained, as well as high combustion efficiency, specially at temperatures higher than 950 °C. Unburnt volatile matter was not present in the FR outlet, and low CH₄ concentration was found. Values for the oxygen demand of gases from 5% to 15% were found in all the experimental work, mainly due to unconverted CO and H₂. Thus, oxygen polishing was quite low.

Minor effect on the gasification efficiency can be expected when the coal particle size is varied. However, it is presumable that there was a continuation of gasification in the FR freeboard, this being more relevant for smaller particles. Thus, it is expected that the oxygen demand would increase as the coal particle size decreases.

Activation of ilmenite with the operating time in the unit was observed. For the reduction it can be considered that ilmenite is already active after 3 hours. On the other hand, the oxidation reaction is not fully activated after 43 hours operation yet. A decrease on the oxygen transport capability of ilmenite was not observed.

5. Nomenclature

C_{coal, eff} carbon present in the effective coal (mol/s)
C_{char to AR} Carbon in the char that goes to the AR (mol/s)
F_{AR outlet} AR gas flow (Lₙ/h)
F_{FR} dry basis product gas flow (Lₙ/h)
F_{H₂O,in} water steam inlet flow in the FR (Lₙ/h)
F_{CH₄,FR, CO₂,FR, CO,FR, H₂O,FR outlet} outlet flows in the FR of CH₄, CO₂, CO, H₂ and H₂O, respectively (mol/s)
MeₓOᵧ oxidized oxygen carrier
MeₓOᵧ₋₁ reduced oxygen carrier
X_{im} reaction conversion of ilmenite
[CH₄]_{FR}, [CO₂]_{FR}, [CO]_{FR}, [H₂]_{FR} dry basis fractions in the product gas of CH₄, CO₂, CO, and H₂
[O]_{coal, eff} amount of oxygen present in the effective coal fed
η_{comb} combustion efficiency
\ allele_c gasification efficiency
Ω g oxygen demand of FR product gas

6. References


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