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**The Effect of Alternative Fuel Combustion in the Cement Kiln Main Burner on
Production Capacity and Improvement with Oxygen Enrichment**

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The Effect of Alternative Fuel Combustion in the Cement Kiln Main Burner on Production Capacity and Improvement with Oxygen Enrichment

W.K. Hiromi Ariyaratne, Morten C. Melaaen, Lars-André Tokheim

Abstract— A mathematical model based on a mass and energy balance for the combustion in a cement rotary kiln was developed. The model was used to investigate the impact of replacing about 45 % of the primary coal energy by different alternative fuels. Refuse derived fuel, waste wood, solid hazardous waste and liquid hazardous waste were used in the modeling. The results showed that in order to keep the kiln temperature unchanged, and thereby maintain the required clinker quality, the production capacity had to be reduced by 1-15 %, depending on the fuel type. The reason for the reduction is increased exhaust gas flow rates caused by the fuel characteristics. The model, which has been successfully validated in a full-scale experiment, was also used to show that the negative impact on the production capacity can be avoided if a relatively small part of the combustion air is replaced by pure oxygen.

Keywords— Alternative fuels, Cement kiln main burner, Oxygen enrichment, Production capacity

I. INTRODUCTION

CEMENT production is a very energy demanding process ranging from 3.2 to 5 MJ/kg of clinker depending on the process conditions [1]. Fossil fuels, such as coal and petroleum coke, have traditionally been used as energy sources in the cement manufacturing industry. However, in recent decades, many countries around the world are implementing the process of applying waste and other alternatives to substitute fossil fuels in cement production, not only to reduce energy costs but also for saving natural resources, decreasing emissions, decreasing the need for waste disposal options and decreasing CO₂ emissions.

Before any alternative fuel is put into use in cement kilns, the net benefit of the use of alternative energy sources and several issues must be taken into concern. The technical feasibility, economic viability as well as impacts on product quality,

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production capacity, emissions and overall operation of the process should be investigated. Finding bottlenecks will clear the path for optimizing the process and the alternative fuel consumption.

The range of alternative fuels is very wide. Some researchers have focused on emissions and pollution due to waste derived fuel combustion in cement kilns [2]-[7]. Impacts on the product and the process have also been investigated by some authors. One study was carried out to investigate air demand and energy input by partly replacing primary fuel by meat and bone meal and sewage sludge [8]. The same authors also investigated the change in clinker composition due to change of fuels [9]. Another study [10] shows the effect of refuse derived fuel (RDF) combustion on clinker production, process and emissions. A detailed mathematical model for optimal selection of raw materials and fuels has been developed by other researchers [11]. In still another study, the flow of exhaust gases and the effect on operational parameters were found [12]. The effects of chemical characteristics of the fuel on key process parameters and the reduction of production capacity due to replacement of primary coal energy by meat and bone meal were investigated previously [13]. Furthermore, the effect of oxygen enriched combustion on the cement production process has been investigated by several authors [14]-[18].

In the present study, a mathematical model was implemented in Matlab, based on a mass and energy balance of a rotary cement kiln. There were two main objectives of the present study: 1) Quantification of the reduction of production capacity due to replacement of around 45% of the primary coal energy by different alternative fuels which have low heating values compared to coal. 2) Quantification of primary air replacement by pure oxygen required to compensate the negative impact on the clinker quality due to use of alternative fuels. The simulation results related to the first objective were verified by results from a full-scale test documented before [19]. A description of the kiln system is described in previous publications [20], [23].

II. FUELS

Analyses of different fuels are shown in Table I. RDF is

derived from municipal and industrial waste which has paper/cardboard, wood, textiles and plastic as the main constituents. SHW is solid hazardous waste and LHW is liquid hazardous waste. Both SHW and LHW contain organic

precalciner (secondary burner) was also carried out in order to determine the total exhaust gas flow rate at normal conditions. This figure is required to determine the reduction in clinker production rate when alternative fuels are used in the main

TABLE I
ULTIMATE ANALYSIS OF DIFFERENT FUELS (AS RECEIVED)

Parameter	Unit	Coal	RDF	Waste wood (China)	SHW	LHW
Net calorific value (NCV)	[MJ/kg]	28.30	18.20	14.90	15.90	14.60
C	[kg/kg]	0.729	0.431	0.403	0.359	0.437
H	[kg/kg]	0.039	0.062	0.053	0.053	0.080
O	[kg/kg]	0.056	0.304	0.390	0.285	0.253
S	[kg/kg]	0.014	0.004	0.000	0.012	0.016
N	[kg/kg]	0.017	0.007	0.041	0.006	0.018
Ash	[kg/kg]	0.136	0.121	0.071	0.167	0.000
Moisture	[kg/kg]	0.010	0.071	0.042	0.118	0.198

solvents, paint, glue and many other organic hazardous waste types, however in SHW, the waste is more pasty and is hence mixed with wood chips to improve handling properties. Among those fuels, only coal is fully fossil; wood is fully biogenic; whereas RDF, SHW and LHW are partly biogenic [21].

III. METHOD

A mass and energy balance for a cement rotary kiln was developed. Process parameters were taken from a full-scale cement plant in Norway which produces around 3400 tons of clinker per day. Only the rotary kiln (main burner) was included in the main model, however a mass balance for the

burner. Mass and energy transfer streams in the rotary kiln are shown in Fig. 1.

A. Case 1

This case corresponds to quantifying the reduction of production capacity due to replacement of primary coal energy by alternative fuels. In the first step of case 1, a mix of coal and each of the four alternative fuels described in Table I (one at a time) was supplied to the main burner. The mass flow rate of alternative fuel replaced around 45% of the primary coal energy, based on a simulation carried out previously for meat and bone meal [13], whereas the coal flow rate was calculated to give the same total fuel energy input as for pure coal

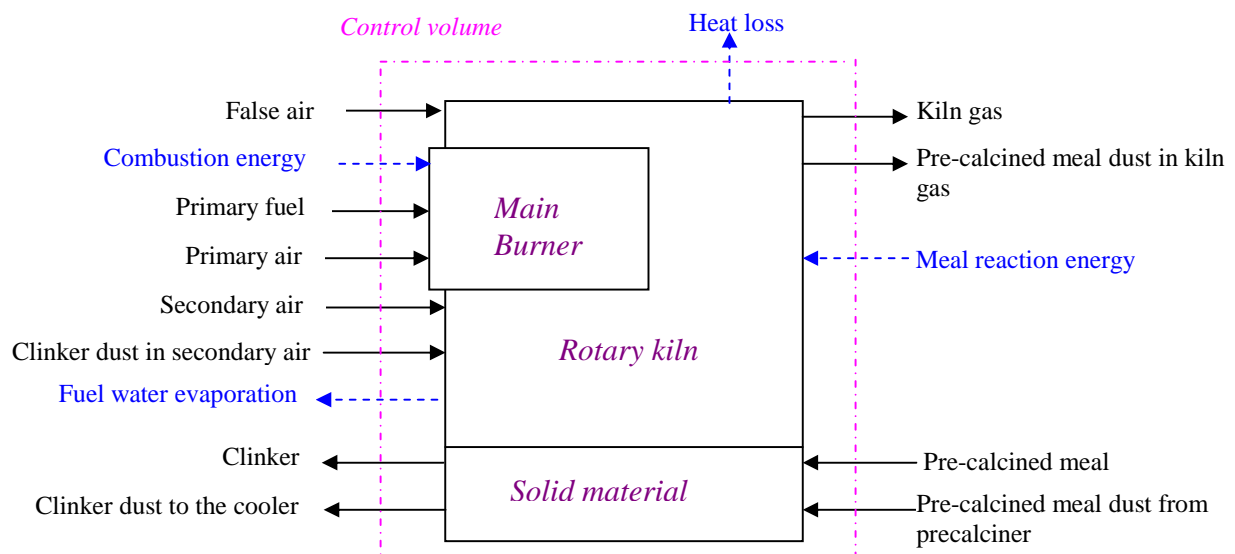


Fig. 1 Conceptual model of the rotary kiln (solid lines: mass streams and sensible energy flows; dotted lines: other energy streams)

combustion (reference case). In the second step, the energy input to the main burner was increased by increasing the coal flow rate of the mixture in order to obtain the same kiln gas temperature as in the coal reference case. Keeping the high temperature is crucial for maintaining the clinker mineral formation rate in the rotary kiln and hence maintaining the product quality. Increasing the fuel input however means that the clinker-specific fuel energy requirement in the main burner was increased. The third step was to tune the coal mass flow rate and clinker production rate in order to have not only kiln gas temperature similar but also the volumetric flow rate of total exhaust gas from the precalciner similar to that of the coal reference case. This was done in order to find the reduction in production capacity due to replacement of part of the coal by an alternative fuel. The reason for keeping the total exhaust gas flow rate at the reference case level is that the major gas flow resistance is in the preheater and precalciner, and this resistance typically represents a bottleneck for the production capacity. The excess O₂ for the combustion in the calciner was kept similar to the value used in the rotary kiln. The clinker-specific energy supply to the precalciner was kept constant in both cases; 90% of that energy was provided via RDF and the rest through coal (similar to the operation of the cement kiln system considered in this work).

B. Case 2

This corresponds to quantification of primary air replacement by pure oxygen. In the main burner, the air for the combustion is supplied through primary and secondary air, and some false air is also sucked into the system due to under pressure. Case 2 is similar to step 1 in case 1, but part of the primary air is replaced by pure oxygen in order to get the same kiln gas temperature as in the reference case, since this can be taken as an indirect measure of clinker quality. In this case, only a mix of RDF and coal was simulated.

IV. MODEL DESCRIPTION

Steady state was assumed for all variables in the system. Since the solid flow is fairly constant during normal operation of the kiln and the residence time for gas flow is in the order of seconds, this assumption is fairly justified. The model has five solid reactions inside the rotary kiln, including (rest) calcination, endothermic melting and exothermic sintering. 100% beta-dicalcium silicate (β -C₂S) formation in the rotary kiln was assumed. MgCO₃ of raw meal was set to zero as MgO in this case is bound in other minerals than carbonates.

The secondary fuel ash is transferred to the meal in the calciner while primary fuel ash is transferred to the meal in the kiln, i.e. both ashes ending up in the clinker. The false air inleakage in the rotary kiln was calculated as a function of secondary air flow. Dust which basically contains precalcined meal is entrained by the kiln gas and transferred to the precalcined meal in the precalciner, which means there is an energy transfer from the rotary kiln to the precalciner. Likewise, dust which basically contains clinker is entrained by the secondary air in the cooler and returned to the rotary kiln,

constituting an energy transfer from the kiln to the cooler. The rate of the combustion reactions were assumed to be fast compared to the residence time in the rotary kiln, hence the chemical kinetics were disregarded. Complete combustion of primary and secondary fuels was assumed. The elemental analysis was used for the evaluation of combustion reactions. A global mechanism was assumed for all elements, and the reactions were considered irreversible. The combustion products were hence taken as CO₂, H₂O, SO₂, N₂ and O₂. Moisture and ash were regarded as inert components. The total air or oxygen supply through the main burner inlets was kept constant independent of fuel type and fuel flow rate. The overall heat transfer coefficient for the evaluation of heat loss through the rotary kiln shell was kept constant in all cases due to lack of more accurate data. Work done by kiln rotation (very small compared to the thermal energy input) was ignored in the energy balance. The specific heat capacities of the kiln gas components were calculated as temperature functions [22].

V. MODEL DEVELOPMENT

The equations for the overall energy balance of the rotary kiln, the overall mass balance of the combined rotary kiln and precalciner, as well as other relevant equations, were developed. Not all equations are presented in this paper, but a description of the model development is given below. The same procedure was carried out for the reference case by using coal instead of a fuel mix.

A. Case 1

In step 1, the feed rates of alternative fuel and coal in the main burner were calculated in order to give the same total energy input in the main burner as in the coal reference case whereas 45% of the energy was supplied via alternative fuels. Then the mass-weighted averages were taken as fuel mixture characteristics. The flow rates of different elements were found by using ultimate analysis of the fuels (Table I). Then the air requirement for the combustion could be obtained; the excess air factor was specified to give a specified concentration of O₂ in kiln exhaust gas. This air is supplied as primary air (approximately 7-10 wt% of the total air supply), secondary air (ca 81-83 wt%) and false air (ca 9-10 wt%). Next, molar flow rates of combustion products were found balancing elements. As the clinker production rate was specified and the primary and secondary fuel ash contents were known (the ash ends up in clinker), the raw meal requirement could be obtained by additionally considering CO₂ release due to the decarbonation (calcination) of the calcium carbonate in the raw meal. The mass flow rate of the precalcined meal going to the rotary kiln includes the secondary fuel ash, some uncalcined calcium carbonate and calcium oxide, compounds other than calcium carbonate in raw meal and some recycled (fully calcined) dust. The recycling means that the real degree of calcination in the calciner is somewhat lower than the apparent degree of calcination. Summing the CO₂ release due to rest calcination inside the rotary kiln and flow of gaseous products due to

combustion and fuel water evaporation, the total kiln gas flow rates and the kiln exhaust gas composition can be obtained.

To find the net enthalpy of solid reactions, equations specified by the German Association of Cement Works (VDZ) were used [22]. Equation (1) and (3) correspond to heat absorption rate for CaCO_3 dissociation and tricalcium aluminate (C_3A) formation. Likewise, (2), (4) and (5) correspond to heat release rate due to formation of tetracalcium aluminoferrite (C_4AF), $\beta\text{-C}_2\text{S}$ and tricalcium silicate (C_3S), respectively. Then the net heat absorption rate due to solid reactions in the kiln could be calculated. To find the C_3S content of clinker, Bogue equations were used [22].

$$\Delta H_{\text{CaCO}_3} = V\dot{m}_{rm} w_{\text{CaCO}_3,rm} (1 - \eta_{cal}) \quad (1)$$

$$\Delta H_{\text{C}_4\text{AF}} = 3.043W\dot{m}_{cl} w_{\text{Fe}_2\text{O}_3,cl} \quad (2)$$

$$\Delta H_{\text{C}_3\text{A}} = X\dot{m}_{cl} (2.65w_{\text{Al}_2\text{O}_3,cl} - 1.692w_{\text{Fe}_2\text{O}_3,cl}) \quad (3)$$

$$\Delta H_{\beta\text{-C}_2\text{S}} = Y\dot{m}_{cl} (2.868w_{\text{SiO}_2,cl} - 0.754w_{\text{C}_3\text{S},cl}) \quad (4)$$

$$\Delta H_{\text{C}_3\text{S}} = Z\dot{m}_{cl} w_{\text{C}_3\text{S},cl} \quad (5)$$

$$H_R = \Delta H_{\text{CaCO}_3} - \Delta H_{\text{C}_4\text{AF}} + \Delta H_{\text{C}_3\text{A}} - \Delta H_{\beta\text{-C}_2\text{S}} - \Delta H_{\text{C}_3\text{S}} \quad (6)$$

Here, \dot{m}_n is the mass flow rate of component n , η_{cal} is the real degree of calcination and $w_{n,q}$ is the weight fraction of component n in stream q . Subscripts rm and cl stands for raw meal and clinker, respectively. The other constants are specific heat release values of different reactions, and these are defined in the section VI.

The overall energy balance is not presented here. However, in Fig. 1, all sensible energy flows along with mass streams and other energy streams are clearly shown.

In step 2, instead of keeping the main burner energy input constant, the coal feed rate was changed until the kiln temperature was as high as in the reference case.

In the third step, the clinker production rate and coal feed rate were tuned in order to keep the kiln gas temperature and the volumetric flow rate of the total exhaust gas from the precalciner similar to the coal reference case. For this step, the mass balance of the precalciner had to be included in the model in order to find the exhaust gas flow from the precalciner. Since the secondary (i.e. precalciner) fuel energy proportion was known ($\text{RDF}:\text{coal} = 9:1$), and the clinker-specific energy consumption for the precalciner was specified, the secondary fuel feed rate could be calculated. Then, using mass-weighted averages for the mixture parameters, the mass balance for the fuel combustion of the precalciner was carried out as for the main burner. The CO_2 release due to calcination inside the precalciner and the fuel moisture evaporation were also accounted for when finding the exhaust gas flow rate from the precalciner. By summing the exhaust gas flow rates from the main burner and from the precalciner, the total gas flow rate out of the precalciner was found.

B. Case 2

The simulation was carried out similar to step 1 of case 1, and RDF was used as the alternative fuel. A volumetric replacement factor was defined for the replacement of primary air by pure oxygen. Properties of the replaced portion of air were replaced by properties of O_2 , and the simulation was carried out by tuning the replacement factor till the kiln gas temperature was as in the coal reference case.

VI. MODEL INPUTS

The main model inputs are briefly described here. The clinker-specific heat consumption was set to 3.5 MJ/kg in the coal reference case, in case 2 and in the first step of case 1, but was changed in other steps in case 1, as described above. The apparent degree of calcination was kept at 88% in all cases; this is roughly corresponding to 46% of the total energy being supplied in the main burner, although slightly higher in steps 2 and 3 of case 1. The raw meal composition was taken as 66 wt% CaO , 21 wt% SiO_2 , 5 wt% Al_2O_3 , 3 wt% Fe_2O_3 and 5 wt% other compounds in clinker. The temperature of fuel, primary air, secondary air, false air, clinker and precalcined meal at the kiln inlet were 323, 1023, 298, 1673 and 1103 K, respectively. The O_2 level in the kiln gas and in the precalciner gas were kept at 3.5 vol% in all cases. The clinker-specific precalcined meal dust flow rate was set to 0.3, and the clinker-specific clinker dust flow rate was set to 0.2. The false air percentage was specified as 12 wt% of secondary air, and the overall heat transfer coefficient was taken as $6.6 \text{ W}\cdot\text{K}^{-1}\cdot\text{m}^{-2}$, based on actual measurements in the kiln system. Temperature functions were used to calculate specific heat capacities of dry coal, air, clinker, combustion products and precalcined meal, and the others were assumed constant [22]. The constants V , W , X , Y and Z in (1) to (5) are 1778 kJ/kg CaCO_3 in raw meal, 67 kJ/kg C_4AF in clinker, 74 kJ/kg C_3A in clinker, 700 kJ/kg C_2S in clinker and 495 kJ/kg C_3S in clinker, respectively [22].

VII. RESULTS AND DISCUSSION

Below, selected simulation results are presented and discussed.

A. Case 1

The impact of main burner combustion of different alternative fuels on the production capacity is discussed here.

Demanding that the fuel energy input to the main burner in step 1 is the same as in the coal reference case and the fraction of energy supply through alternative fuels is fixed, the fuel flow rates of mix of coal and alternative fuels are increased, mainly reflecting the heating value of the alternative fuels (Fig. 2 to Fig. 5). The mass flow rate for the mix of LHW and coal is highest since LHW has the lowest heating value among all alternative fuels (Table I). Furthermore, differences in ultimate analysis, for example ash content and water content, in different fuels will also give some impact on the clinker-specific fuel requirement. Given that the heating values are the

same, the fuels with high ash content have higher clinker-specific fuel mass flow requirement whereas it is similar for fuels with high water content as well (Fig. 3 and Fig. 5). The reason is that the water and ash from the fuel have to be heated from ambient temperature up to the clinker temperature, whereas the precalcined meal only has to be heated from the calcination temperature. The kiln gas temperatures are 24°C, 28°C, 22°C and 95°C lower than in the reference case for RDF, waste wood, SHW and LHW, respectively. This also clearly reflects the fuel heating values; the higher the heating value, the higher the kiln gas temperature. Furthermore, the kiln gas temperature is impacted by the ultimate analysis of the fuel. This effect is also described in a previous publication [13]. However, the total exhaust gas flow rates at normal conditions are now slightly higher for RDF (0.6%) and waste wood (1.2%) but significantly higher for LHW (8.8%).

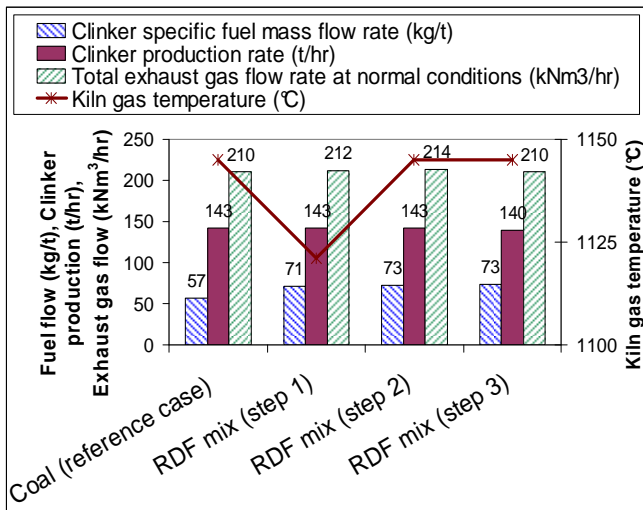


Fig. 2 Fuel flow, clinker production, exhaust gas flow rate and gas temperature for refuse derived fuel mix compared with the coal reference case

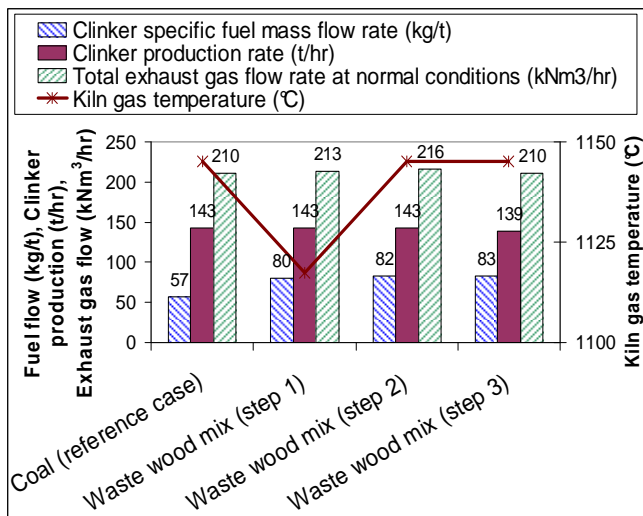


Fig. 3 Fuel flow, clinker production, exhaust gas flow rate and gas temperature for waste wood fuel mix compared with the coal reference case

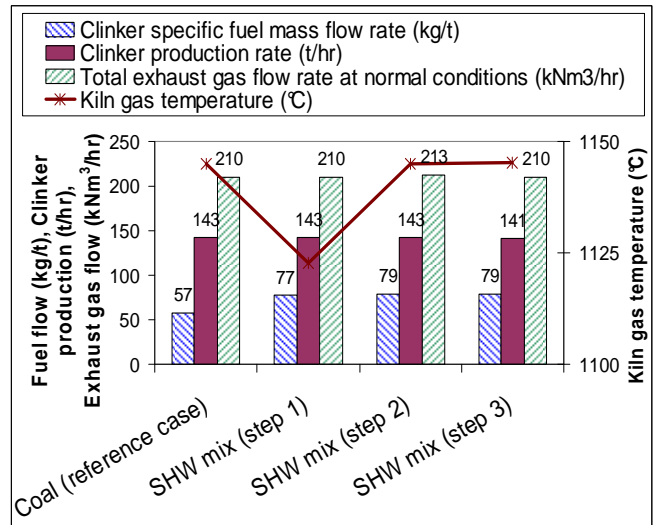


Fig. 4 Fuel flow, clinker production, exhaust gas flow rate and gas temperature for solid hazardous waste fuel mix compared with the coal reference case

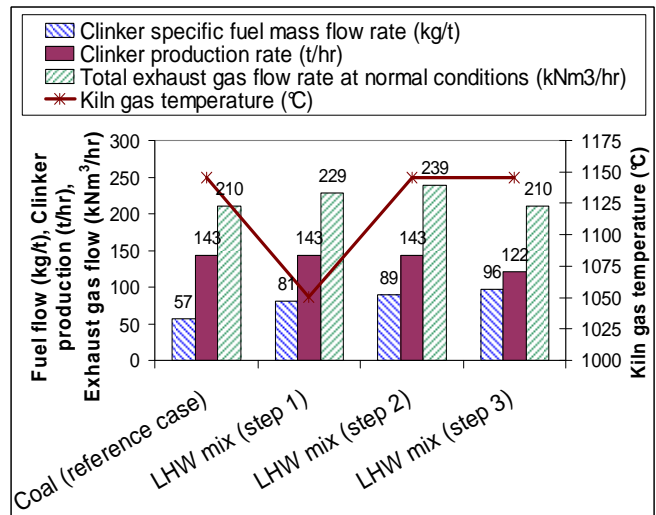


Fig. 5 Fuel flow, clinker production, exhaust gas flow rate and gas temperature for liquid hazardous waste fuel mix compared with the coal reference case

In step 2, as shown in Fig. 2 to Fig. 5, the specific fuel consumption is increased to obtain same kiln gas temperature as in the reference case (1145 °C). The clinker-specific energy consumption in the main burner is increased by 2.9%, 3.9%, 3.2% and 14.2% in this step for different fuel mixes. The air requirement for the combustion increases with the increase of fuel consumption, hence the kiln gas flow rate is further increased causing an increase in the total exhaust gas flow rate. The total exhaust gas flow rates are increased from 212 to 214 kNm³/hr for RDF fuel mix, from 213 to 216 kNm³/hr for waste wood fuel mix, from 210 to 213 kNm³/hr for SHW fuel mix and from 229 to 239 kNm³/hr for LHW fuel mix.

In step 3, the total exhaust gas flow rate is the same as in the reference case (210 kNm³/hr), however, the penalty is a reduced clinker production rate. Around 1.8%, 2.8%, 1.2% and 14.7% reductions in clinker production are calculated for

RDF, waste wood, SHW and LHW, respectively. In this case, the raw meal requirement is lower, hence the fuel energy requirement and CO₂ release due to calcination is lower in the precalciner, resulting in reduced exhaust gas from the precalciner, compensating for the increased exhaust gas from the kiln. (The sum of the two is equal to the total flow rate in the reference case.)

It appears that LHW is quite different from the other alternative fuels, giving a much lower capacity for production. Chemical characteristics of LHW can explain this. As described above, the kiln gas flow rate can be identified as one of the key parameters affecting the production capacity when alternative fuels are combusted in the main burner. Basically, the kiln gas flow rate follows the same trend as the air flow rate. The fact that the kiln gas flow rate is higher than the air flow rate is due to the addition of fuel (which is converted to combustion products) and the addition of CO₂ from the rest calcination. Therefore, the kiln gas flow may increase due to extra addition of fuel mass and hence combustion products when alternative fuels are used.

The other reason; air requirement for the combustion; depends on many factors. As presented in Table I, the very high water content in LHW (19.8%) compared to other fuels has to be heated from ambient temperature up to the clinker temperature which will reduce the kiln gas temperature requiring more fuels and more air for combustion in order to keep the required product quality. The higher the air requirement, the higher the kiln gas flow. Further, an increase in molar H/C ratio means a decrease in air requirement. For example, coal and LHW have H/C ratios of 0.64 and 2.19, respectively, meaning that the air requirement in mass basis is higher for coal than for LHW. However, this effect is inverted in energy basis (0.32 Nm³/MJ for coal vs. 0.51 Nm³/MJ for LHW), also giving a higher clinker-specific air requirement for LHW (0.60 Nm³/kg of clinker) than for coal (0.51 Nm³/kg of clinker). Therefore, it is evident that the LHW fuel mix gives the lowest production capacity, because it will produce highest kiln gas flow due to above mentioned characteristics. This finding corresponds well with actual experience at the plant.

B. Case 2

The replacement of part of the primary air by pure oxygen was investigated. The required replacement in order to get same kiln gas temperature as reference case was found for the RDF and coal mixture which was used in step 1 of case 1. Relevant results are presented and discussed here.

By demanding that the fuel energy input to the main burner is the same as in the reference case and replacing 45% of this energy by RDF, the kiln gas temperature is decreased reflecting the lower calorific value of RDF compared to coal (Fig. 7), when using only air as the oxidizer for the combustion. However, replacement of some primary air by oxygen may give a kiln gas temperature similar to that of the reference case also when keeping the clinker production rate similar to that of the reference case. This means that the clinker quality and the production capacity can be maintained

in spite of the fuel replacement. It was found that in order to satisfy the temperature requirement without decreasing the production, around 23.5 vol-% of the primary air would have to be replaced by pure oxygen (Fig. 6). This corresponds to an O₂ supply of 2.14 t/hr. An extra benefit of this oxygen enriched combustion is a slight decrease in total exhaust gas flow rate, from 212 to 205 kNm³/hr, i.e. even lower than in the coal reference case (210 kNm³/hr), which actually implies that a slight increase in production capacity could be obtained (Fig. 7).

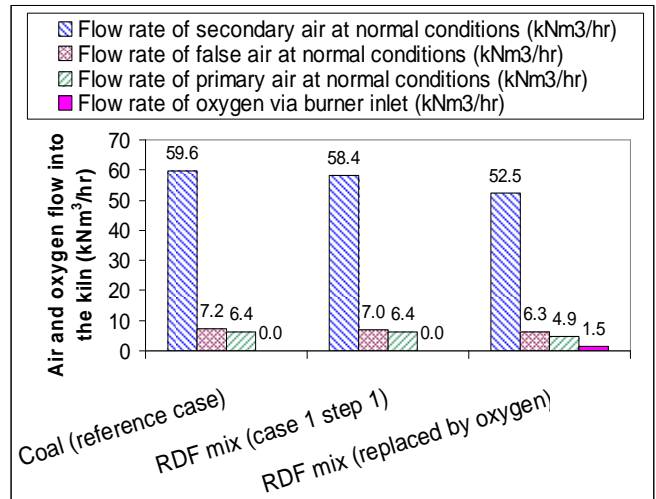


Fig. 6 Flow rates of air and oxygen streams for RDF fuel mix compared with the coal reference case

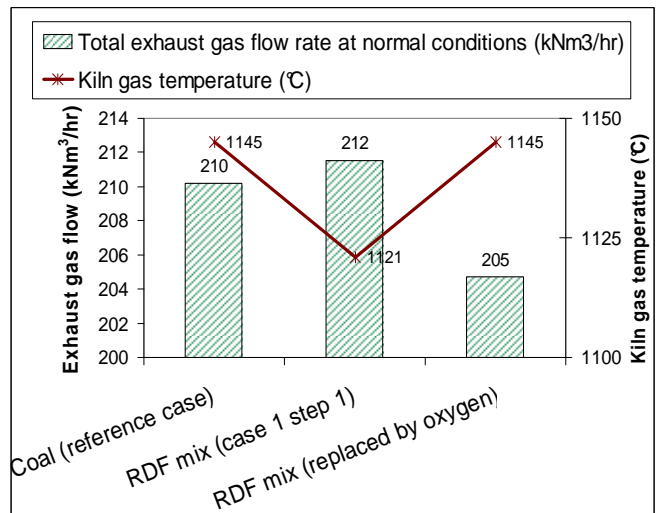


Fig. 7 Exhaust gas flow rate and gas temperature for RDF fuel mix compared with the coal reference case

Oxygen enriched combustion seems to be a favourable alternative to overcome the problems encountered with alternative fuel combustion in cement kilns in respect with product quality and production capacity. One obvious reason of the increased kiln gas temperature is that some of the N₂ which otherwise enters along with the combustion air, is now avoided, hence does not have to be heated. Further, some authors have observed that the maximum temperature inside

the kiln and the rate of consumption of volatiles was significantly higher when oxygen enriched air was used [15]. Other researchers [14] have also discussed the beneficial role of oxygen in the fuel combustion characteristics and conclude that the oxygen enrichment may increase the production rate and kiln efficiency because the reaction initiates faster and gives a higher maximum flame temperature.

However, one should of course also take into account the costs related to oxygen production in contrast to the benefits. Early-phase cost estimates indicate that replacing part of the air by oxygen may be economically favourable, considering the positive impact on the production capacity.

VIII. CONCLUSION

When replacing around 45% of the primary coal energy by different alternative fuels at the main burner, the production capacity must be reduced by 1.2%, 1.8%, 2.8% and 14.7% for solid hazardous waste, refuse derived fuel, waste wood and liquid hazardous waste, respectively, in order to maintain the kiln exhaust gas temperature and hence keep the clinker quality unchanged. The high value for liquid hazardous waste is mainly due to a very high water content in this fuel.

It is possible to avoid the reduction in production capacity by replacing a relatively small part of the combustion air by pure oxygen. When a mixture of refuse derived fuel and coal is used as the primary fuel, replacement of 23.5 vol-% of the primary air (or about 2 vol-% of the total combustion air) by pure oxygen is required. Early-phase cost estimates indicate that such a scheme may be economically favourable, considering the positive impact on the production capacity.

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