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Utilization of oxygen as partial replacement for air in cement kiln combustion processes

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Summary:

This thesis is started with introduction, continued with model building and its simulation results, then ended with economic analysis.

The idea of this study is to reduce the nitrogen content of the atmospheric air which is used in conventional combustion. Nitrogen, which is the biggest part of the air, consumes much energy released from fuel combustion as well as occupies the volume of the equipment. Theoretically, by decreasing the portion of the nitrogen in the combustion air, heat efficiency as well as equipment capacity can be increased.

The study is done by building mathematical model of heat and energy balance around the kiln and calciner of the existing cement plant then simulated at different oxygen level to get changes in fuel rate, exit gas flow rate, pure oxygen flow rate, and increase in capacity.

The study shows that there is heat efficiency at the kiln but reversely at the calciner. Increase in capacity is identified on both coal and waste fuel case when combustion air at kiln is partially replaced with oxygen. Economic analysis shows that implementation of utilization of oxygen as partial replacement for air in cement kiln combustion accompanied with capacity increase is profitable.

Preface

Praise to the God almighty for his bless and willingness so I can complete this thesis. May we always be in His grace and protection.

I would like to express my thanks to the University of South Eastern Norway (USN) and Government of Norway for the opportunity given to me to study master program in process technology at USN.

Special thanks to my supervisor Lars-André Tokheim who has helped and guided me to walk a winding road in the work of this thesis. I also thank to my co-supervisor Christoffer Moen for his valuable comment and suggestion during this thesis development.

My thanks to my beloved family for their endless prayer, sacrifice, and support for me so I can reach this point:

- My wife Ovi shofianur, who has encouraged me to study master program and always be there for me in this journey.
- My kids, Faustine, Ritzaleigh and Audi, for whom this all I did – I wish you all can do better than me.
- My father Syamsir and my mom Sulasni.
- My Brother Wendri, and my sisters Linda, Retno and Yusi.

Finally, I want to thank to my all my friends and all parties in USN who has supported me during my study period.

I realize that this thesis is far from perfection, improvement and correction is required. But at least this can be a reference for a similar works at USN. Hopefully, similar work in the future can do much better.

Jakarta, Friday, May 14, 2021.

Syaiful Bahri

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Nomenclature

AFR	Air to Fuel Ratio
ASU	Air Separation Unit
CAPEX	Capital Expenditure
DCF	Discounted Cash Flow
ELSE	ELektrifisert SEmentproduksjon (Norwegian)
HHV	Higher Heating Value (kJ/kg)
HYSYS	Hyprotech Systems
LHV	Lower Heating Value (kJ/kg)
NOK	Norwegian Kroner
NPV	Net Present Value
OPEX	Operational Expenditure
STP	Standard Temperature and Pressure
USA	United States of America
USD	United States Dollar
USN	University of South Eastern Norway
XRD	X-Ray Diffraction

1 Introduction

This part describes about background of the topic, objective, problem statement, outline of the thesis, and task description.

1.1 Background

Combustion is the oldest chemical process developed by man in history inspired by nature phenomenon. The cave man has used fire to grill their food and as light source in the dark by burning the wood. Following the development of human civilization, utilization of fire also evolute from a simple cooking or lighting to a more complex thing in many aspects of human activity. Modern man also developed combustion process to the higher extent with different fuel and complicated technic of combustion. Science behind the combustion has been developed and compiled from time to time by researcher. Nevertheless, there are still many new subjects of combustion are waiting to be explored and interested to be researched.

Sophisticated and complex combustion process has been become the heart of many chemical processes in industry nowadays where some of them involve massive usage of fuel and huge of combustion equipment in size. One of the chemical process where combustion play a vital role in the process is cement industry where raw material mainly composed of limestone from quarry is burned in cement kiln at high temperature to form clinker where calcium oxide is the main substance of the compound.

One of current research's topic on combustion is partially replacement of air with oxygen. The idea of combustion with pure or partially pure oxygen combustion on cement plant has been discussed since 1976[1] in order to increase the combustion efficiency and production capacity. There has been application of partially replacement of air with oxygen on cement plant in USA reported, but the information reagarding the performance of the cement plant is not widely opened [2], [3].

In this thesis, application of partially replacement of air with oxygen on combustion in cement industry is selected as the topic. Existing Norcem cement located at Brevik is used as the basis of the study. Therefore, the application of partially replacement application of air with oxygen on combustion in this thesis is envisaged for existing cement plant.

This thesis uses theoretical and analytical approach to find the impact of application of partially replacement of air with oxygen on combustion in existing cement plant. Common engineering practice and simple economic analysis application are also as complimentary to get a comprehensive result.

There is no specific study yet done on this topic at USN, therefore carrying out thesis on this topic is challenging and need to be done.

1.2 Problem Statement

Utilization waste as fuel in industrial combustion process has become a breakthrough in solving waste problem, particularly the municipal solid waste from urban area from environmental and economic point of view. But thermodynamically, waste fuel has a lower heating value compared to other fuel such as oil, fuel gas or coal. This constraint results in a lower cement production capacity of using the waste fuel compared to coal or fuel gas. To compensate the lower capacity, kiln using the solid waste fuel must be larger in size that in turn increase investment cost.

In combustion process, heat released by the fuel burning is absorbed by the material fed into the combustion chamber as well by flue gas coming out from the combustion process. Since air is supplied into the kiln as oxygen source for combustion process, Nitrogen as 79% of air also included in the air. This Nitrogen supply absorbs large amount of heat resulted from combustion without positive contribution to the production process. Beside of the existence of Nitrogen in combustion process causes environmental problem due to NOx formation which harm the environment due to acid rain and greenhouse effect caused by NOx.

Theoretically, by the partially replacing of the air with oxygen in combustion process, there will be less Nitrogen exist in the combustion process. The absence / less of Nitrogen means that there will be heat conserved and less exhasut gas that can be utilized for production increase. By the other word, we can increase the capacity of the existing kiln by partially replacing of the air with Oxygen in combustion process.

It is needed to study to what extent of temperature can be attained by partially replacing of the air with oxygen in combustion process beside of the appropriate flow rate of oxygen required as well as waste fuel flow rate to get an optimum combustion process. By having the information of temperature and flow rate, required changes in mechanical design of the kiln can be recommended. Including to be recommended the general design of additional facility required regarding the pure oxygen provision.

1.3 Objective

Objective of the study is to evaluate the technical and comercial impact of utilization of oxygen as partial replacement for air in cement kiln combustion processes. Evaluation will be focused on several representative parameters technically and comercially, that are:

- Maximum theoritical temperature that can be attained by the combustion process.
- Maximum waste fuel flow rate can be increased.
- Maximum production capacity of kiln can be increased.
- Required constructional changes of the kiln.
- Cost related to the process changes.
- Economic prices of oxygen for the process change.

1.4 Thesis Outline

This thesis is arranged in a sequential order so that is easy to be understood and followed, the outline is as following:

- Started with the introduction of the thesis and background of why the idea of partially replacing of the air with oxygen in combustion process is selected as the topic in this thesis.
- A more detail about the focus of the topic and objective of the study is described in the next chapter.
- Then in the next chapter literature survey related to the topic is reported to give a theoretical basis to the idea.
- Evaluation of operating variable changes due to the idea implementation is done by analytical approach using mathematical model that will be developed for base case in the next chapter.
- In the subsequent chapter, the mathematical developed then is applied to any variation of appropriate operating variable or parameter.
- Results of simulation is then presented in the subsequent chapter.
- Recommendation of required changes of the existing kiln and additional facility required is presented in the next chapter.
- Finally, economic analysis is to be done in the next chapter to evaluate the feasibility of the idea implementation.

1.5 Task description

This thesis basically is a sort of report of the following tasks as agreed in the early stage of this thesis development with the supervisor, that are:

- Describe the relevant chemical and physical processes
  This task will be covered in chapter 2 – literature survey.
- Evaluate the impact of using oxygen as a partial replacement for air in waste fuel combustion; the impact on fuel feed rate and production capacity is of particularly high importance
  This task will be covered in chapter 4 - evaluation of the impact of using oxygen as a partial replacement for air in waste fuel combustion.
- As part of the evaluation, make a model based on mass and energy balances for (part of) the system
  This task will be covered in chapter 3 – mass and energy balance and appendix C - Mass and Energy balance model
- Select an appropriate simulation tool based on an assessment of different available options, and calculate relevant mass flow rates, temperatures, duties, etc. using the selected tool
  Microsoft Excel will be used as a tool for simulation.
- Make relevant process flow diagrams with process values for selected cases based on relevant design basis values
  This task will be covered in chapter 2 - literature survey and chapter 3- mass and energy balance.

- Simulate different cases with the selected simulation tool, varying key parameters in the system
  This task will be carried out and the result will be presented in appendix E – coal case simulation result and appendix F – waste fuel case simulation result.
- Describe required constructional changes to the kiln system
  This task will be covered in chapter 5 – constructional changes to the kiln.
- Assess local handling and intermediate storage of oxygen
  This task will be covered in chapter 6 – local handling and intermediate storage of oxygen.
- Determine the required size of relevant equipment units
  This task will be covered in chapter 6 – local handling and intermediate storage of oxygen.
- Make estimates of investment costs (CAPEX) and operational costs (OPEX) of the suggested process changes.
  This task will be covered in chapter 7 – economic analysis and appendix G - Relative NPV calculation result
- Determine what oxygen purchasing prices that would be required for oxygen utilization to be economically viable in the cement plant
  This task will be covered in chapter 7 – economic analysis.

2 Literature Survey

This chapter describes about theoretical basis related to the topic i.e combustion, cement kiln process description, and combustion of solid fuels in oxygen-enriched air.

2.1 Combustion

Within this chapter some relevant combustion's topics will be described.

2.1.1 Combustion Chemistry

According to Cambridge dictionary, combustion is defined as "The chemical process in which substances mix with oxygen in the air to produce heat and light. The actual chemical kinetics of reaction between fuel and oxygen does not start with H_2 and O_2 directly. In fact, H_2 and O_2 do not directly react with each other at all; breaking both H–H and O–O bonds simultaneously during a single molecular collision is less probable than other chemical pathways. Oxidation reaction between fuel and oxygen involves many reactions and steps– up to thousand, depend on the fuel type, the reaction is called elementary reaction [4]. The collection of elementary reactions that describe the overall global reaction is referred to as a reaction or combustion mechanism.

But for simplification, combustion chemistry is usually expressed in single global stoichiometric reaction. A stoichiometric mixture contains the exact amount of fuel and oxidizer such that after combustion is completed, all the fuel and oxidizer are consumed to form products. This ideal mixture approximately yields the maximum flame temperature, as all the energy released from combustion is used to heat the products [4].

Combustion stoichiometry for a general hydrocarbon fuel with air can be expressed as

The stoichiometric combustion reaction for $C_\alpha H_\beta O_\gamma$ in air

$$C_\alpha H_\beta O_\gamma + \left(\alpha + \frac{\beta}{4} - \frac{\gamma}{2}\right)(O_2 + 3.77N_2) \rightarrow \alpha CO_2 + \frac{\beta}{2} H_2O + 3.77\left(\alpha + \frac{\beta}{4} - \frac{\gamma}{2}\right) N_2 \tag{2.1}$$

The stoichiometric combustion reaction for $C_\alpha H_\beta O_\gamma S_\delta$ in air:

$$C_\alpha H_\beta O_\gamma S_\delta + \left(\alpha + \frac{\beta}{4} - \frac{\gamma}{2} + \delta\right)(O_2 + 3.77N_2) \rightarrow \alpha CO_2 + \frac{\beta}{2} H_2O + \delta SO_2 + 3.77\left(\alpha + \frac{\beta}{4} - \frac{\gamma}{2} + \delta\right) N_2 \tag{2.2}$$

The stoichiometric combustion reaction for $C_\alpha H_\beta O_\gamma N_\delta S_\epsilon$ in air:

$$C_\alpha H_\beta O_\gamma N_\delta S_\epsilon + \left(\alpha + \frac{\beta}{4} - \frac{\gamma}{2} + \epsilon\right)(O_2 + 3.77N_2) \rightarrow \alpha CO_2 + \frac{\beta}{2} H_2O + \epsilon SO_2 + \left(3.77\left(\alpha + \frac{\beta}{4} - \frac{\gamma}{2} + \epsilon\right) + \frac{\delta}{2}\right) N_2 \tag{2.3}$$

The stoichiometric combustion reaction for $C_\alpha H_\beta O_\gamma F_\delta$ in air:

$$C_\alpha H_\beta O_\gamma F_\delta + \left(\alpha + \frac{\beta - \delta}{4} - \frac{\gamma}{2}\right)(O_2 + 3.77N_2) \rightarrow \alpha CO_2 + \frac{\beta - \delta}{2} H_2O + \delta HF + 3.77\left(\alpha + \frac{\beta - \delta}{4} - \frac{\gamma}{2}\right) N_2 \tag{2.4}$$

2.1.2 Heating Value

Heating values of a fuel (units of kJ/kg or MJ/kg) are traditionally used to quantify the maximum amount of heat that can be generated by combustion with air at standard conditions (STP) (25ºC and 101.3 kPa). The amount of heat release from combustion of the fuel will depend on the phase of water in the products. If water is in the gas phase in the products, the value of total heat release is denoted as the lower heating value (LHV). When the water vapor is condensed to liquid, additional energy (equal to the latent heat of vaporization) can be extracted and the total energy release is called the higher heating value (HHV). The value of the LHV can be calculated from the HHV by subtracting the amount of energy released during the phase change of water from vapor to liquid as [5]

$$LHV = HHV - \frac{N_{H2O,P} M_{H2O} h_{fg}}{N_{fuel} M_{fuel}} \quad (MJ/kg) \tag{2.5}$$

HHV for Combustion Processes from a Constant-Pressure Reactor

$$HHV = \frac{-Q^{o}_{rxn,p}}{N_{fuel} M_{fuel}} \quad (MJ/kg) \tag{2.6}$$

HHV for Combustion Processes from a Constant-Volume Reactor

$$HHV = \frac{-Q^{o}_{rxn,v} - \left(\sum_i N_{i,p} - \sum_i N_{i,R}\right) \hat{R}_u T_o}{N_{fuel} M_{fuel}} \quad (MJ/kg) \tag{2.7}$$

2.1.3 Adiabatic Flame Temperature

Adiabatic flame temperature is highest temperature of combustion that can be achieved when there is no heat loss to surrounding environment and all the energy released from combustion is used to heat the combustion products.

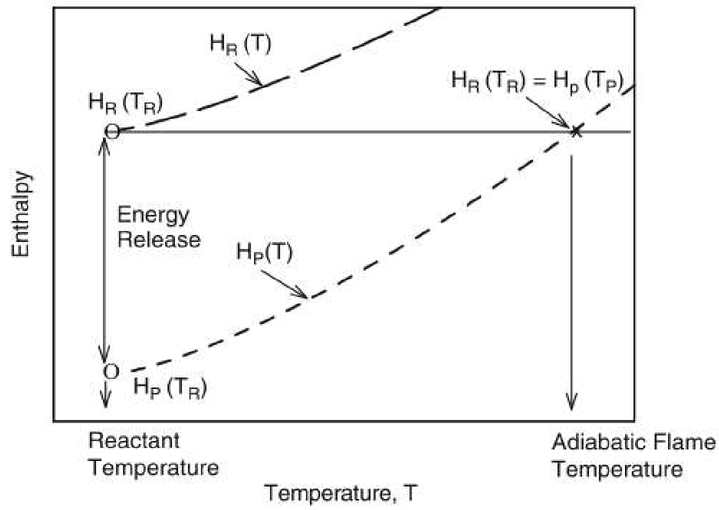

Fig. 2.1 Graphical interpretation of adiabatic flame temperature [4]

Figure 2.1 is a graphic explanation of how the adiabatic flame temperature is determined. At the initial reactant temperature, the enthalpy of the product mixture is lower than that of the reactant mixture. The energy released from combustion is used to heat up the products such that the condition H_P (TP) = H_R (TR) is met [4].

Adiabatic constant-pressure analysis is used here to calculate the adiabatic flame temperature. Under this idealized condition, conservation of energy is [4]

$$H_p(T_p) = H_R(T_R) \tag{2.8}$$

Where

$$H_p(T_p) = \sum_i N_{i,p}\hat{h}_{i,P} = \sum_i N_{i,p}\left[\Delta\hat{h}^o_{i,P} + \Delta\hat{h}_{si,P}(T_p)\right] \tag{2.9}$$

And

$$H_R(T_R) = \sum_i N_{i,R}\hat{h}_{i,R} = \sum_i N_{i,R}\left[\Delta\hat{h}^o_{i,R} + \Delta\hat{h}_{si,R}(T_R)\right] \tag{2.10}$$

Where,

$\Delta\hat{h}^o_i = Enthalphy\ of\ formation$ (MJ/kg or MJ/kmol)

$\Delta\hat{h}_{si} = Sensible\ entahlphy\ of\ reaction$ (MJ/kg or MJ/kmol)

And,

$$\Delta\hat{h}_{si} = \int_{T_0}^{T} \hat{C}_p(T)\, dT \tag{2.11}$$

And,

$$\Delta\hat{h}_i = \Delta\hat{h}_i^o + \Delta\hat{h}_{si} \tag{2.12}$$

The standard enthalpy of formation, $\Delta\hat{h}_i^o$, quantifies the chemical bond energy of a chemical species at standard conditions. The enthalpy of formation of a substance is the energy needed for the formation of that substance from its constituent elements at STP conditions (25C and 1 atm). The molar base enthalpy of formation, $\Delta\hat{h}_i^o$, has units of MJ/kmol, and the mass base enthalpy of formation, $\Delta\hat{h}_i^o$, has units of MJ/kg. Elements in their most stable forms, such as C (graphite) , H2, O2 , and N2 , have enthalpies of formation of zero[4].

Three different methods can be used to obtain product temperature T_P [5]:

1. Using an average cp value,

$$\Delta\hat{h}_{si} = \int_{T_0}^{T} \hat{C}_p(T)\, dT \tag{2.13}$$

$$\Delta\hat{h}_i = \Delta\hat{h}_i^o + \Delta\hat{h}_{si} \tag{2.14}$$

$$T_p = T_R + \frac{LHV.N_{fuel}.M_{fuel}}{\sum_i N_{i,P}\hat{c}_{pi}} \tag{2.15}$$

2. An iterative enthalpy balance,

With an initial guess of flame temperature, $T_{p,1}$, one evaluates $H_p(T_{p,1})$. If $H_p(T_{p,1}) < H_R(T_R)$, we guess a higher flame temperature, $T_{p,2}$. One repeats this process until the two closest temperatures are found such that $H_p(T_{f,1}) < H_R(T_R) < H_p(T_{f,2})$. The product temperature can be estimated by linear interpolation. This method, although more accurate, still assumes complete combustion to the major products.

3. Finding the equilibrium state using computer software (such as Cantera, STANJAN).

2.2 Cement kiln

This chapter will describe in brief the introduction of the cement plant process description and introduction on oxygen enriched air combustion.

2.2.1 Cement kiln process description

Cement is a substance that mainly consists of lime (CaO), silica (SiO_2), alumina (Al_2O_3) and iron oxide (Fe_2O_3) that used in construction as a binder of a mixture of sand, gravel and water to form concrete. Cement materials can be classified into two distinct categories: non-hydraulic cements and hydraulic cements according to their respective setting and hardening mechanisms.

There are many type of cement, but the most commonly used type of modern cement is Portland cement, a form of hydraulic cement, is by far the most common type of cement in general used around the world. This cement is made by heating limestone (calcium carbonate as the source of CaO) with other materials (such as clay as the source of SiO_2, Al_2O_3, Fe_2O_3) to 1,450 °C (2,640 °F).

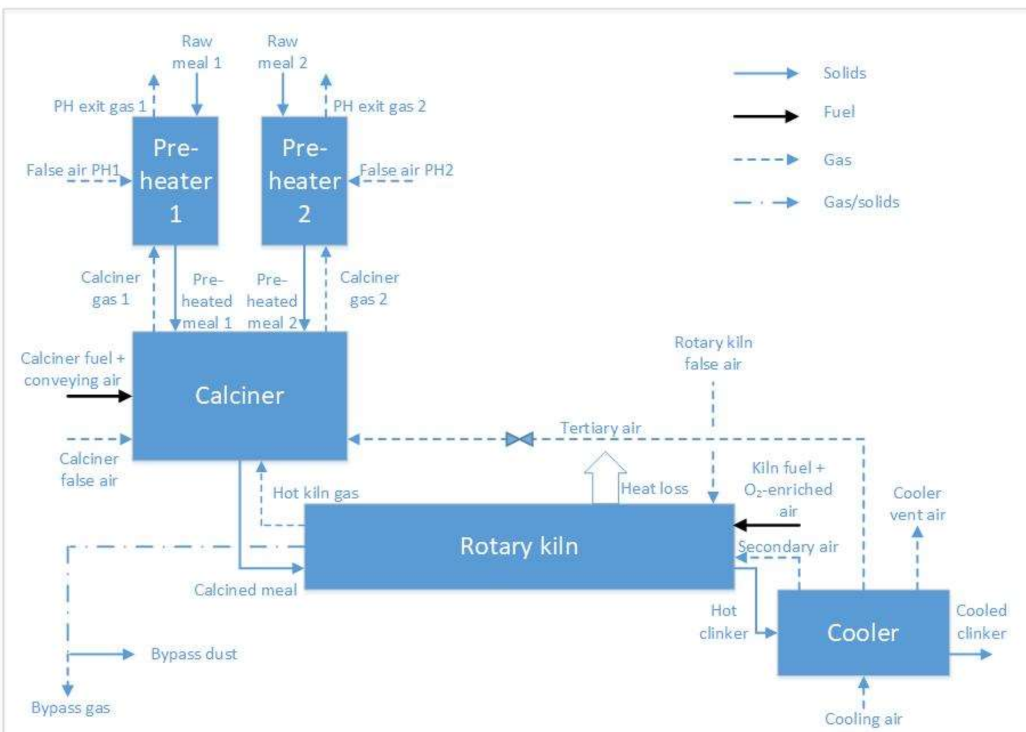

Fig. 2.2 Cement plant block diagram [5]

Figure 2.2 ilsutrates the process flow diagram of cement plant. The heart of cement plant is what as known as kiln, the equipment where transformation process of the raw material to cement occurs via pyroprocessing process. The main chemical reaction in pyroprocessing is a process known as calcination that liberates a molecule of carbon dioxide from the calcium carbonate to form calcium oxide, or quicklime, which then chemically combines with the other materials in the mix to form calcium silicates and other cementitious compounds. The resulting hard substance, called 'clinker', is then ground with a small amount of gypsum into a powder to make ordinary Portland cement.

There are several types of kiln and the most common used in cement industry is rotary kiln. Typical of rotary kiln is as depicted by figure 2.3 below.

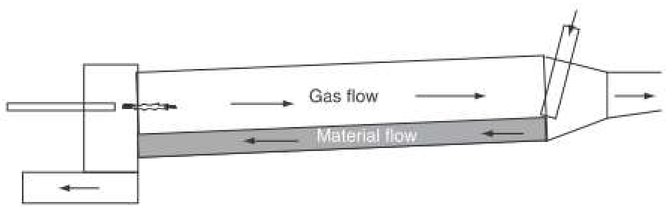

Figure 2.3 Schematic diagram of countercurrent flow rotary kiln configuration [6].

2.2.2 Reaction zone and chemical reaction in rotary Kiln

Based on chemical and physical process taking place to the feed material inside the kiln, there are five distinct zones within the kiln, their location and length being different for each type of kiln system used. These zones are as shown in the table 2.1.

Table 2.1 Zone in rotary cement kiln [6]

Chemical & physical process	Temperature range of material (°C)
Drying and preheating zone	15 - 805
Calcining zone	805 - 1200
Upper transition zone	1200 - 1400
Sintering zone	1400 - 1510
Cooling (lower transition) zone	1590 - 1290

For more detail of reaction taking place inside the kiln are as shown in table 2.2 below.

Table 2.2 Reactions and reaction enthalpies [7].

Reaction	Reaction equation	Standard enthalpy of reaction [kJ/kg]
I. Formation of oxides and decomposing reactions		
Evaporation of water	$H_2O(l) \rightarrow H_2O(g)$	2453
Decomposition of kaolinite	$Al_2O_3.2SiO_2.2H_2O \rightarrow Al_2O_3+2SiO_2+2H_2O$	780
Oxidation of carbon	$C + O_2 \rightarrow CO_2$	-33913
Dissociation of MgCO 3	$MgCO_3 \rightarrow MgO+CO_2$	1395
Dissociation of CaCO3	$CaCO_3 \rightarrow CaO+CO_2$	1780
II. Formation of intermediates		
Formation of CA	$CaO+Al_2O_3 \rightarrow CaO.Al_2O_3$	-100
Formation of C2F	$2CaO+ Fe_2O_3 \rightarrow 2CaO.Fe_2O_3$	-114
Formation of β-C2S	$2CaO + SiO_2 \rightarrow 2CaO.SiO_2$	-732
III. Sintering reactions		
Formation of C4AF	$CA + C_2F + CaO \rightarrow C_4AF$	25
Formation of C3A	$CA + 2CaO \rightarrow C_3A$	25
Formation of C3S	$\beta\text{-}C_2S + CaO \rightarrow C_3S$	59

2.2.3 Rotary cement kiln energy usage

It is needed energy about 800 kcal (3.4 MJ) to form 1 kg clinker which is defined as the difference of input and heat output of the process, and almost the same amount heat loses during the process to environment [6]. Table 2.3 and 2.4 shown the theoretical heat requirement and heat loses of a cement kiln.

Table 2.3 Typical cement kiln heat balance [6]

Heat Loss/Transfer	Kcal/kg Clinker	Percent of Total (800 kcal/kg Basis)
Raw meal to clinker	417	52.1
Preheater exhaust	183	22.9
Cooler exhaust	78	9.8
Clinker discharge	17	2.1
Dust loss	17	2.1
Shell loss	88	14.75
Total	800	100

Table 2.4 Theoretical minimum process heat of formation of cement clinker [6]

Event/Process	Temperature Range (°C)	Energy (kcal/ kg Clinker)
Heat in		
Sensible heat to raw material at temperature	20–450	170
Dehydration of clay at temperature	450	40
Sensible heat into raw material at	450–900	195
Dissociation of $CaCO_3$ at	900	475
Sensible heat into material at	900–1400	125
Net heat of melting	1400	25
Subtotal	20–1400	1030
Heat out		
Exothermic crystallization of dehydrated clay	–	10
Exothermic formation of cement compounds	–	100
Cooling of clinker	1400–1420	360
Cooling of CO_2	900–920	120
Cooling and condensing of steam	450–470	20
Subtotal	1400–1420	610
Theoretical minimum process heat required to form 1 kg of clinker		420

2.3 Combustion of solid fuels in oxygen-enriched air in cement kilns

Combustion in oxygen enriched air as illustrated by figure 2.4 is a modification form of conventional combustion of fuel with air where air is partially replaced with oxygen at any certain level hence the oxygen fraction will be higher in the combustion air compared to atmospheric air. The lower fraction of nitrogen in combustion air theoretically can increase the fuel efficiency or production capacity.

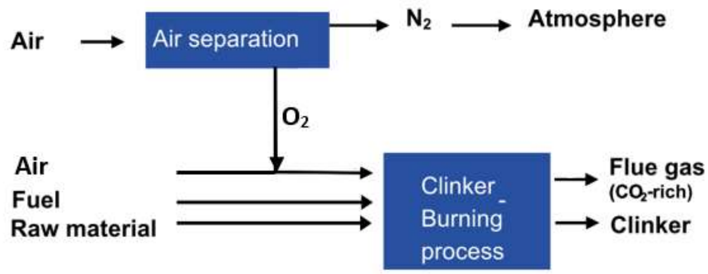

Figure 2.4 Schematic diagram of oxygen-enriched air combustion with ASU unit.

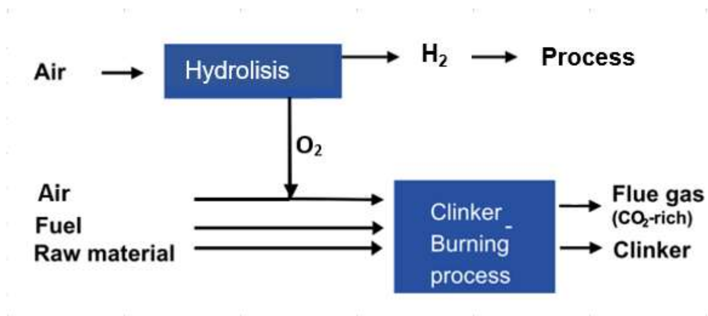

Figure 2.5 Schematic diagram of oxygen-enriched air combustion with hydrolisis unit.

Another terminology related to modified combustion air is oxy-fuel combustion technology that based on the concept of oxygen combustion and replaces the air currently used in cement kilns with a mixture of pure oxygen and exhaust CO2 recycled back to the kiln. This oxy-fuel combustion is particularly for CO2 capture purpose. Figure 2.5 below depicts the concept of oxy-fuel combustion.

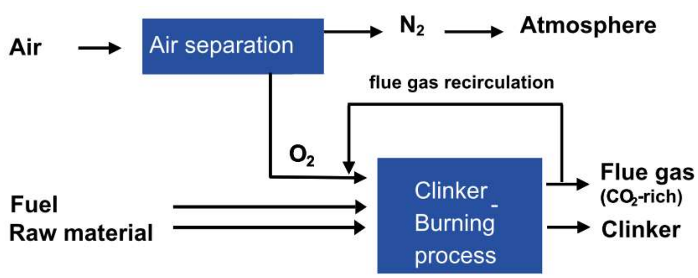

Figure 2.5 Schematic diagram of oxy-fuel combustion [8].

There have been many research and study of oxygen-enriched air combustion of cement kiln conducted, mostly driven by CO2 capture issue since the exhaust gas from oxygen-enriched air combustion cement kiln is dominated by CO2 that readily to be stored without further complex treatment. From economic point of view, research on oxygen-enriched air combustion of cement kiln is mainly purposed to increase the production capacity and fuel efficiency. The study by [9], describes in general the various aspects of oxygen-enriched air combustion of cement kiln. Also, several study on flame profile and heat radiation profile of the burner of the kiln has been conducted by [10], [11] and [12].

Application of oxygen-enriched air combustion of cement kiln has been reported at California Portland Cement Company's – Mojave Plant [2] and at TXI Midlothian Cement Plant – Texas USA[3]. The California Portland Cement Company's plant reported the increase of clinker production in the ratio around 4 ton per day clinker / ton per day oxygen as depicted by figure 2.6 below.

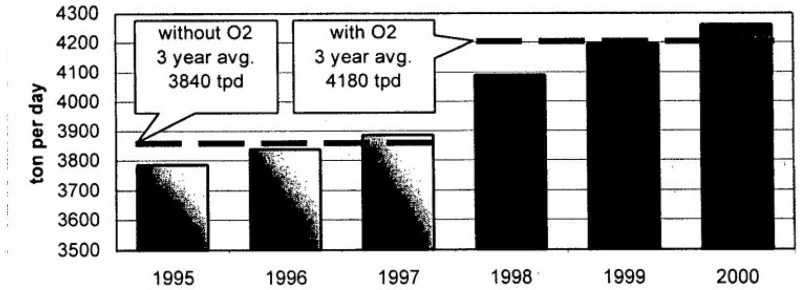

Figure 2.6 Average daily clinker production of California Portland Cement Company [2].

Meanwhile the TXI Midlothian Cement Plant reported a production increase of over 6% was achieved using oxygen enrichment in the kilns as shown figure 2.7.

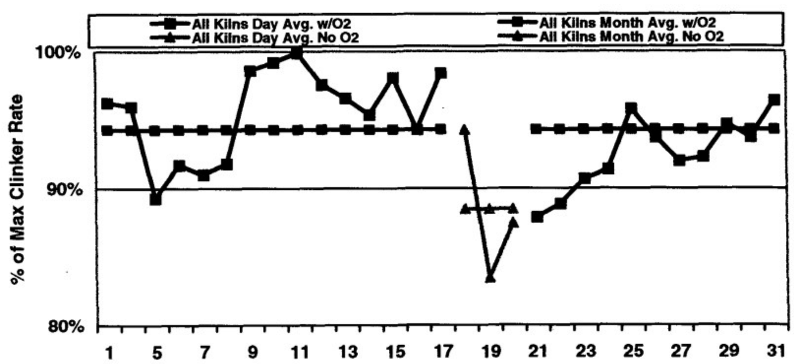

Figure 2.7 Average clinker production as a percentage of maximum of TXI Midlothian Cement Company [3].

In addition to the benefits derived from oxygen-enriched air combustion technology, some challenges in associated with oxygen-enriched air combustion also arises and need to be resolved. More detail regarding challenges in oxygen-enriched air combustion will be describe later in the next chapter.

3 Mass and energy balance

This chapter describes about calculation basis and mathematical correlation to be used in mass and energy balance calculation around rotary kiln.

Spreadsheets is to be developed based on the series and sequence of mass energy balance formula in the chapter below for simulating the varied cases as attached in appendix C, meanwhile the simulation result is attached in appendix E and F.

3.1 Calculation basis

3.1.1 Fuel characteristic

Coal is used in calciner and rotary kiln in simulation of coal case. Meanwhile for waste fuel case simulation, coal is used in calciner and fuel mix referred to Norcem Brevik's specification is used in rotary kiln. Specification of fuel is as stipulated in the table 3.1 below.

Table 3.1. Ultimate analysis (C, H, O, S, N), moisture content, ash content, lower heating value (LHV) and fossil fraction and split between rotary kiln and calciner for different fuels [13].

Parameter	Unit	Coal	RDF	SHW	AM	LHW	Fuel mix rotary kiln	Fuel mix calciner
Mass fraction of C	kg/kg	0.722	0.348	0.359	0.463	0.437	0.601	0.395
Mass fraction of H	kg/kg	0.040	0.050	0.053	0.065	0.080	0.054	0.049
Mass fraction of O	kg/kg	0.057	0.245	0.285	0.149	0.253	0.121	0.232
Mass fraction of S	kg/kg	0.012	0.003	0.012	0.004	0.016	0.011	0.006
Mass fraction of N	kg/kg	0.016	0.006	0.006	0.097	0.018	0.034	0.007
Mass fraction of moisture	kg/kg	0.018	0.250	0.118	0.029	0.198	0.060	0.192
Mass fraction of ash	kg/kg	0.135	0.098	0.167	0.192	0.000	0.118	0.119
Lower heating value	MJ/kg	28.0	14.2	15.9	19.4	14.6	23.1	16.2
Fossil fraction	-	100 %	30 %	70 %	0 %	100 %	78 %	48 %
Mass fraction used in the rotary kiln	-	56 %	0 %	0 %	22 %	22 %	100 %	0 %
Mass fraction used in the calciner	-	12 %	65 %	24 %	0 %	0 %	0 %	100 %
Energy fraction used in the rotary kiln	-	67 %	0 %	0 %	19 %	14 %	100 %	0 %
Energy fraction used in the calciner	-	20 %	57 %	23 %	0 %	0 %	0 %	100 %

3.1.2 Design basis

Several parameters needed to be defined as a basis for mass and energy balance calculation which are referred to Norcem Brevik's operating conditions and parameters as stipulated in table 3.2 and 3.3 below.

Table 3.2. Design basis values for the mass balance [13].

Parameter	Unit	Coal reference case
Clinker production	t/y	1000000
Operation time	h/y	7315

Specific thermal energy consumption	MJ/kg_clinker	3.4
Mass fraction of $CaCO_3$ in raw meal	kg/kg	0.77
Thermal energy fraction in the rotary kiln	-	38 %
Degree of calcination in the calciner	-	94 %
Primary air supply (rel. to stoich.) in the main burner	-	8 %
False air in the rotary kiln	kg/kg_clinker	0.03
O_2 in the rotary kiln exit gas	%	3.0
Kiln operating pressure	Pa	101325
Calciner operating pressure	Pa	101325
Calciner exit gas temperature	°C	900

Table 3.3. Design basis values for energy balance [13].

Parameter	Unit	Coal reference case
Reference temperature	°C	25
Hot clinker temperature	°C	1400
Calcination temperature	°C	900
Kiln exit gas temperature	°C	1150
Specific calcination enthalpy (at reference)	MJ/kg CO_2	-3.6
Rotary kiln fuel inlet temperature	°C	30
Rotary kiln primary air inlet temperature	°C	30
Specific rotary kiln heat loss	MW	6
Thermal energy fraction in rotary kiln	%	38

3.2 Mass balance

Mass conservation law or as well-known as Lomonosov-Lavoiser law stated that the mass of enclosed system is remain constants overtime regardless the process happened in that system.

3.2.1 Capacity calculation

Raw meal $\dot{m}_{RM}$ (kg/h) is the raw material fed to the system. For simplification, $CaCO_3$ is assumed the only source of CO_2 in raw meal. Therefore, the CO_2 fraction in the raw meal $w_{CO2,RM}$ [kg/kg] can be calculated as

$$w_{CO2,RM} = w_{CaCO3,RM} \cdot \frac{M_{CO2}}{M_{CaCO3}} \tag{3.1}$$

Where,

$w_{CaCO3,RM}$ = CaCO3 content in the raw meal [kg/kg], (design basis value)

M_{CO2} = molecular masses of CO_2 [kg/mol]

M_{CaCO3} = molecular masses of $CaCO_3$ [kg/mol]

The pre calciner-generated CO_2, $\dot{m}_{CO2,precal}$ [kg/h], is calculated as

$$\dot{m}_{CO2,precal} = \dot{m}_{CO2,RM} \cdot f_{precal} \tag{3.2}$$

Where,

$f_{precal} =$ calcination degree of calciner [%], (design basis)

The pre-calcined meal, $\dot{m}_{precal}$ [kg/h], is then found by:

$$\dot{m}_{precal} = \dot{m}_{RM} - \dot{m}_{CO2,precal} \tag{3.3}$$

The $\dot{m}_{precal}$ needs to be adjusted with ash and dust to balance the mass flow rate around the calciner referred to the design basis.

The post calciner-generated CO_2, $\dot{m}_{CO2,postcal}$ [kg/h], is calculated as

$$\dot{m}_{CO2,postcal} = \dot{m}_{CO2,RM} \cdot f_{postcal} \tag{3.4}$$

Where,

$f_{postcal} =$ calcination degree of rotary kiln [%], (design basis)

The pre-calcined meal, $\dot{m}_{clinker}$ [kg/h], is then found by:

$$\dot{m}_{clinker} = \dot{m}_{precal} - \dot{m}_{CO2,postcal} \tag{3.5}$$

The $\dot{m}_{clinker}$ needs to be adjusted with ash and dust to balance the mass flow rate around the calciner referred to the design basis.

Since the design basis is referred to $\dot{m}_{clinker}$, backward calculation should be done by goal seek.

3.2.2 Mass balance around kiln

According to that above law, mass balance around the kiln is calculated to determine the clinker production rate, fuel, air, and flue gas flow rate as illustrated by figure 3.1 below.

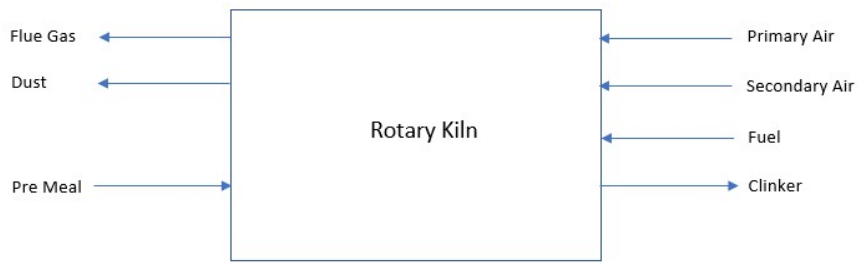

Figure 3.1 Mass balance flow diagram

The clinker production rate is calculated as; $\dot{m}_{cli}$ [kg/h] is calculated as

$$\dot{m}_{cli}\ [kg/h] = \frac{\dot{m}_{cli}\ [t/y]}{t_{op}\ [\mathrm{h}/y]}\ .1000 \tag{3.6}$$

Where,

$\dot{m}_{cli}$ = clinker production rate$[t/y]$ (design basis value)

t_{op} = annual operation time [h/y] (design basis value)

The required thermal energy input to the process, E_{th} [MJ/h], is defined as

$$E_{th} = \dot{m}_{cli}\ .\hat{E}_{th} \tag{3.7}$$

Where,

$\hat{E}_{th}$ = Specific thermal energy consumption [MJ/kg clinker] (design basis value)

$\dot{m}_{cli}$ = clinker production rate$[kg/h]$

The thermal energy input in the rotary kiln, $E_{th,rk}$ [MJ/h], then found as

$$E_{th,rk} = f_{th,rk} \, . E_{th} \tag{3.8}$$

Where,

$f_{th,rk}$ = rotary kiln thermal energy fraction [%], (design basis value)

The fuel requirement in the rotary kiln, $\dot{m}_{fuel,rk}$ [kg/h], is calculated as

$$\dot{m}_{fuel,rk} = \frac{E_{th,rk}}{LHV_{fuel,rk}} \tag{3.9}$$

Where,

$LHV_{fuel,rk}$ = rotary kiln fuel lower heating value [MJ/kg], (design basis value)

The ash input in the rotary kiln, $\dot{m}_{ash,rk}$ [t/h], is defined as

$$\dot{m}_{ash,rk} = \dot{m}_{fuel,rk} \, . w_{ash,fuel,rk} \tag{3.10}$$

Where,

$w_{ash,fuel,rk}$ = ash content in the rotary kiln fuel [kg/kg]

The dust input to rotary kiln is calculated as a balance of the precalcined meal flow rate and the ash from rotary kiln fuel referred to design basis value as follow

$$\dot{m}_{Dust} = \dot{m}_{precal} + \dot{m}_{ash,rk} - \dot{m}_{cli} \tag{3.11}$$

The stoichiometric combustion air flow rate in the rotary kiln, $\dot{m}_{air,rk,st}$ [kg/h], is determined as

$$\dot{m}_{air,rk,st} = \dot{m}_{fuel,rk} \, . AFR_{rk,st} \tag{3.12}$$

Where,

$AFR_{rk,st} =$ stoichiometric air/fuel ratio of the rotary kiln fuel [kg/kg], is calculated based on the fuel mix composition given in Table 3.1.

The excess air in the rotary kiln is set to a value that gives the correct kiln inlet O_2 concentration (which is a design basis value), by adjusting excess air factor, λ_{rk} [kg/kg]:

$$\dot{m}_{air,rk,exc} = \dot{m}_{air,rk,st} \, . (\lambda_{rk} - 1) \tag{3.13}$$

The actual air supply to the rotary kiln, $\dot{m}_{air,rk}$ [k/h], is then found by addition:

$$\dot{m}_{air,rk} = \dot{m}_{air,rk,st} + \dot{m}_{air,rk,exc} \tag{3.14}$$

The primary air flow rate in the rotary kiln, $\dot{m}_{air,prim,rk}$ [kg/h], is determined as

$$\dot{m}_{air,prim,rk} = \dot{m}_{air,rk,st} \cdot f_{air,prim,rk} \tag{3.15}$$

Where,

$f_{air,prim,rk} =$ the primary air factor [kg/kg], (design basis value)

The false air flow rate in the rotary kiln, $\dot{m}_{air,false,rk}$ [t/h], is found as

$$\dot{m}_{air,false,rk} = \dot{m}_{cli} \cdot f_{air,false,rk} \tag{3.16}$$

Where,

$f_{air,false,rk} =$ false air factor [kg/kg], (design basis value)

And the secondary air, $\dot{m}_{air,sec,rk}$ [t/h], is then found by

$$\dot{m}_{air,sec,rk} = \dot{m}_{air,rk} - \dot{m}_{air,prim,rk} - \dot{m}_{air,false,rk} \tag{3.17}$$

The excess O_2 in the rotary kiln, $\dot{m}_{O2,rk,exc}$ [kg/h], is then determined as

$$\dot{m}_{O2,rk,exc} = \dot{m}_{air,rk,exc} \cdot \dot{w}_{O2,air} \tag{3.18}$$

Where,

$\dot{w}_{O2,air} = O_2$ content in air [kg/kg]

The nitrogen added through rotary kiln fuel, $\dot{m}_{N2,fuel,rk}$ [kg/h], is calculated as

$$\dot{m}_{N2,fuel,rk} = \dot{m}_{fuel,rk} \cdot w_{N,fuel,rk} \tag{3.19}$$

Where,

$w_{N,fuel,rk} =$ nitrogen content in the rotary kiln fuel [kg/kg], calculated from the fuel composition

It is assumed that all fuel-N is converted to N2 (fuel NOx formation is neglected). The total N2 added to the rotary kiln, $\dot{m}_{N2,rk}$ [t/h], is the found as

$$\dot{m}_{N2,rk} = \dot{m}_{N2,fuel,rk} + m_{air,rk} \cdot w_{N2,air} \qquad (3.20)$$

Where,

$\dot{w}_{N2,air}$ = N_2 content in air [kg/kg]

It is assumed that all fuel-sulphur is oxidized to SO_2. The total SO_2 added to the rotary kiln, $\dot{m}_{SO2,rk}$ [kg/h], is then calculated as

$$\dot{m}_{SO2,rk} = \dot{m}_{fuel,rk} \cdot w_{S,fuel,rk} \cdot \frac{M_{SO2}}{M_S} \qquad (3.21)$$

Where,

$w_{S,fuel,rk}$ = sulphur content in the rotary kiln fuel [kg/kg]

M_{SO2} = molecular mass of SO_2 [kg/mol]

M_S = molecular mass of S [kg/mol]

The H_2O added to the rotary kiln, $\dot{m}_{H2O,rk}$ [kg/h], is the sum of moisture content of fuel and moisture formed from combustion reaction as follow

$$\dot{m}_{H2O,rk} = \dot{m}_{fuel,rk} \cdot (w_{moist,fuel,rk} + w_{H,fuel,rk} \cdot \frac{M_{H2O}}{M_{H2}}) \qquad (3.22)$$

Where,

$w_{moist,fuel,rk}$ = fuel moisture content [kg/kg]

M_{H2O} = molecular mass of H2O [kg/mol]

M_{H2} = molecular mass of H2 [kg/mol]

The CO_2 released through combustion in the rotary kiln, $\dot{m}_{CO2,comb,rk}$ [kg/h], is calculated as

$$\dot{m}_{CO2,comb,rk} = \dot{m}_{fuel,rk} \cdot w_{C,fuel,rk} \cdot \frac{M_{CO2}}{M_C} \qquad (3.23)$$

Where,

$w_{C,fuel,rk}$ = fuel moisture content [kg/kg]

M_{CO2} = molecular mass of CO2 [kg/mol]

M_C = molecular mass of C [kg/mol]

The total CO_2 generation in the rotary kiln, $\dot{m}_{CO2,rk}$ [kg/h], is then found by addition:

$$\dot{m}_{CO2,rk} = \dot{m}_{CO2,comb,rk} + \dot{m}_{CO2,postcal} \tag{3.24}$$

The total gas mass flow rate at the kiln gas outlet (the solids inlet), $\dot{m}_{gas,rk}$ [kg/h], is the sum of the five different gas components described above

$$\dot{m}_{gas,rk} = \dot{m}_{O2,rk,exc} + \dot{m}_{N2,rk} + \dot{m}_{SO2,rk} + \dot{m}_{H2O,rk} + \dot{m}_{CO2,rk} \tag{3.25}$$

The mass fraction of O_2, $w_{O2,rk,exc}$ in the kiln exit gas is

$$w_{O2,rk,exc} = \frac{\dot{m}_{O2,rk,exc}}{\dot{m}_{gas,rk}} \tag{3.26}$$

The mass fraction of N_2, $w_{N2,rk}$ in the kiln exit gas is

$$w_{N2,rk} = \frac{\dot{m}_{N2,rk}}{\dot{m}_{gas,rk}} \tag{3.27}$$

The mass fraction of SO_2, $w_{SO2,rk}$ in the kiln exit gas is

$$w_{SO2,rk} = \frac{\dot{m}_{SO2,rk}}{\dot{m}_{gas,rk}} \tag{3.28}$$

The mass fraction of H_2O, $w_{H2O,rk}$ in the kiln exit gas is

$$w_{H2O,rk} = \frac{\dot{m}_{H2O,rk}}{\dot{m}_{gas,rk}} \tag{3.29}$$

The mass fraction of CO_2, $w_{CO2,rk}$ in the kiln exit gas is

$$w_{CO2,rk} = \frac{\dot{m}_{CO2,rk}}{\dot{m}_{gas,rk}} \tag{3.30}$$

The total gas volume flow rate at the kiln exit is

$$\dot{V}_{gas,rk} = \dot{V}_{O2,rk,exc} + \dot{V}_{N2,rk} + \dot{V}_{SO2,rk} + \dot{V}_{H2O,rk} + \dot{V}_{CO2,rk} \tag{3.31}$$

And,

$$\dot{V}_{O2,rk,exc} = \dot{m}_{O2,rk,exc} \cdot \rho_{O2} \tag{3.32}$$

$$\dot{V}_{N2,rk} = \dot{m}_{N2,rk} \cdot \rho_{N2} \tag{3.33}$$

$$\dot{V}_{SO2,rk} = \dot{m}_{SO2,rk} \cdot \rho_{SO2} \tag{3.34}$$

$$\dot{V}_{H2O,rk} = \dot{m}_{H2O,rk} \cdot \rho_{H2O} \tag{3.35}$$

$$\dot{V}_{CO2,rk} = \dot{m}_{CO2,rk} \cdot \rho_{CO2} \tag{3.36}$$

And,

ρ_{O2} is density of O_2, calculated as

$$\rho_{O2} = \frac{P.M_{O2}}{R.T} \tag{3.37}$$

ρ_{N2} is density of N_2, calculated as

$$\rho_{N2} = \frac{P.M_{N2}}{R.T} \tag{3.38}$$

ρ_{SO2} is density of SO_2, calculated as

$$\rho_{SO2} = \frac{P.M_{SO2}}{R.T} \tag{3.39}$$

ρ_{H2O} is density of H_2O, calculated as

$$\rho_{H2O} = \frac{P.M_{H2O}}{R.T} \tag{3.40}$$

ρ_{H2O} is density of CO_2, calculated as

$$\rho_{CO2} = \frac{P.M_{CO2}}{R.T} \tag{3.41}$$

Where,

$P = kiln\ pressure\ (design\ basis\ value)$

$P = kiln\ exit\ gas\ temperature\ (design\ basis\ value)$

$R =$ gas constant 8.314

The volume fraction of O_2 in the rotary kiln gas, $y_{O2,rk}$ is calculated by

$$v_{O2,rk} = \frac{\dot{V}_{O2,rk}}{\dot{V}_{gas,rk}} \tag{3.42}$$

The O_2 volume fraction at dry conditions in the kiln inlet, $y_{O2,rk,dry}$ which should be equal to the design basis value, is calculated by:

$$v_{O2,rk,dry} = \frac{\dot{V}_{O2,rk}}{\dot{V}_{gas,rk} - \dot{V}_{H2O}} \tag{3.43}$$

Pure oxygen required for any oxygen level of combustion air is calculated by

$$\dot{m}_{O2,pure} = \dot{m}_{air,rk,exc} \cdot \frac{(O_{2,level} - 20.95)}{100} \tag{3.44}$$

Where,

$O_{2,level} = oxygen\ level\ in\ kiln\ combustion\ air$

3.2.3 Mass balance around calciner

Mass balance around calciner is limited to gas flow calculation only since it is needed to determine the overall gas flow that will be bottleneck of capacity increase.

The thermal energy input in the calciner, $E_{th,cal}$ [MJ/h], then found as

$$E_{th,cal} = f_{th,cal} \cdot E_{th} \tag{3.45}$$

Where,

$f_{th,cal}$ = Calciner thermal energy fraction [%], (design basis value)

The fuel requirement in the rotary kiln, $\dot{m}_{fuel,cal}$ [kg/h], is calculated as

$$\dot{m}_{fuel,cal} = \frac{E_{th,cal}}{LHV_{fuel,cal}} \tag{3.46}$$

Where,

$LHV_{fuel,cal}$ = Calciner fuel lower heating value [MJ/kg], (design basis value)

For oxygen enriched air combustion case, since the clinker exit gas is less than the reference case, less energy input to calciner from the rotary kiln exit gas needs to be compensated by more fuel demand at calciner. Additional fuel $\dot{m}_{add,fuel,ca}$ (kg/hr) required in calciner is calculated as:

$$\dot{m}_{add,fuel,,cal} = \frac{\dot{E}_{spec,gas,rk,ref\ case} - \dot{E}_{spec,gas,rk,O2\ case}}{LHV_{fuel,cal}} \tag{3.47}$$

Where,

$\dot{E}_{spec,gas,rk,ref\ case}$ = Specific energy of rotary kiln exit gas of reference case [MJ/hr]

$\dot{E}_{spec,gas,rk,O2\ case}$ = Specific energy of rotary kiln exit gas of O_2 enriched case [MJ/h]

$\dot{E}_{spec,gas,rk,ref\ case}$ is calculated as,

$$\dot{E}_{spec,gas,rk,ref\ case} = \frac{\dot{E}_{gas,rk,ref\ case}}{\dot{m}_{gas,rk,ref\ case}} \tag{3.48}$$

And $\dot{E}_{spec,gas,rk,O2\ case}$ is calculated as,

$$\dot{E}_{spec,gas,rk,O2\ case} = \frac{\dot{E}_{gas,rk,O2\ case}}{\dot{m}_{gas,rk,O2\ case}} \tag{3.49}$$

$\dot{E}_{gas,rk,ref\ case}$ and $\dot{E}_{gas,rk,O2\ case}$ is calculated following formula [X] below.

Total fuel required $\dot{m}_{tot,fuel,cal}$ [kg/hr] in calciner is,

$$\dot{m}_{tot,fuel,cal} = \dot{m}_{fuel,cal} + \dot{m}_{add,fuel,cal} \tag{3.50}$$

The stoichiometric combustion air flow rate in the calciner, $\dot{m}_{air,cal,st}$ [kg/h], is determined as

$$\dot{m}_{air,cal,st} = \dot{m}_{tot,fuel,cal} \cdot AFR_{cal,st} \tag{3.51}$$

Where,

$AFR_{cal,st} =$ stoichiometric air/fuel ratio of the calciner fuel [kg/kg], is calculated based on the fuel mix composition given in Table 3.1.

The excess air in the calciner is set to a value that gives the correct kiln inlet O_2 concentration (which is a design basis value), by adjusting excess air factor, λ_{rk} [kg/kg]

$$\dot{m}_{air,cal,exc} = \dot{m}_{air,cal,st} \cdot (\lambda_{rk} - 1) \tag{3.52}$$

The actual air supply to the calciner, $\dot{m}_{air,cal}$ [t/h], is then found by addition

$$\dot{m}_{air,cal} = \dot{m}_{air,cal,st} + \dot{m}_{air,cal,exc} \tag{3.53}$$

The excess O_2 in the calciner, $\dot{m}_{O2,cal,exc}$ [kg/h], is then determined as

$$\dot{m}_{O2,cal,exc} = \dot{m}_{air,cal,exc} \cdot \dot{w}_{O2,air} \tag{3.54}$$

Where,

$\dot{w}_{O2,air} = O_2$ content in air [kg/kg]

The nitrogen added through calciner fuel, $\dot{m}_{N2,fuel,cal}$ [t/h], is calculated as

$$\dot{m}_{N2,fuel,cal} = \dot{m}_{fuel,cal} \cdot w_{N,fuel,cal} \tag{3.55}$$

Where,

$w_{N,fuel,cal}$ = nitrogen content in the calciner fuel [kg/kg], calculated from the fuel composition

It is assumed that all fuel-N is converted to N2 (fuel NOx formation is neglected).

The total N2 added to the calciner, $\dot{m}_{N2,cal}$ [t/h], is the found as,

$$\dot{m}_{N2,cal} = \dot{m}_{N2,fuel,cal} + m_{air,cal} \cdot w_{N2,air} \tag{3.56}$$

Where,

$\dot{w}_{N2,air}$ = N_2 content in air [kg/kg]

It is assumed that all fuel-sulphur is oxidized to SO_2. The total SO_2 added to the calciner, $\dot{m}_{SO2,rk}$ [kg/h], is then calculated as

$$\dot{m}_{SO2,cal} = \dot{m}_{fuel,cal} \cdot w_{S,fuel,cal} \cdot \frac{M_{SO2}}{M_S} \tag{3.57}$$

Where,

$w_{S,fuel,cal}$ = sulphur content in the calciner fuel [kg/kg]

M_{SO2} = molecular mass of SO_2 [kg/mol]

M_S = molecular mass of S [kg/mol]

The H_2O added to the calciner, $\dot{m}_{H2O,cal}$ [kg/h], is the sum of moisture content of fuel and moisture formed from combustion reaction as follow

$$\dot{m}_{H2O,cal} = \dot{m}_{fuel,cal} \cdot (w_{moist,fuel,cal} + w_{H,fuel,cal} \cdot \frac{M_{H2O}}{M_{H2}}) \tag{3.58}$$

Where,

$w_{moist,fuel,cal}$ = fuel moisture content [kg/kg]

M_{H2O} = molecular mass of H2O [kg/mol]

M_{H2} = molecular mass of H2 [kg/mol]

The CO_2 released through combustion in the calciner, $\dot{m}_{CO\ ,comb,cal}$ [kg/h], is calculated as

$$\dot{m}_{CO2,comb,cal} = \dot{m}_{fuel,cal} \cdot w_{C,fuel,cal} \cdot \frac{M_{CO2}}{M_C} \tag{3.59}$$

Where,

$w_{C,fuel,cal}$ = fuel moisture content [kg/kg]

$M_{CO2} = \text{ molecular mass of CO2}$ [kg/mol]

$M_C = \text{ molecular mass of C}$ [kg/mol]

The total CO_2 generation in the calciner, $\dot{m}_{CO2,cal}$ [kg/h], is then found by addition

$$\dot{m}_{CO2,cal} = \dot{m}_{CO2,comb,cal} + \dot{m}_{CO2,precal} \tag{3.60}$$

The pre calciner-generated CO_2, $\dot{m}_{CO2,precal}$ [t/h], is calculated as

$$\dot{m}_{CO2,precal} = \dot{m}_{CO2,RM} \cdot f_{precal} \tag{3.61}$$

Where,

$f_{precal} = \text{ calcination degree of calciner}$ [%], (design basis)

The total gas mass flow rate at the calciner gas outlet (the solids inlet), $\dot{m}_{gas,cal}$ [kg/h], is the sum of the five different gas components described above

$$\dot{m}_{gas,cal} = \dot{m}_{O2,cal,exc} + \dot{m}_{N2,cal} + \dot{m}_{SO2,cal} + \dot{m}_{H2O,cal} + \dot{m}_{CO2,cal} \tag{3.62}$$

The mass fraction of O_2, $w_{O2,cal,exc}$ in the calciner exit gas is

$$w_{O2,cal,exc} = \frac{\dot{m}_{O2,cal,exc}}{\dot{m}_{gas,cal}} \tag{3.63}$$

The mass fraction of N_2, $w_{N2,cal}$ in the calciner exit gas is

$$w_{N2,cal} = \frac{\dot{m}_{N2,cal}}{\dot{m}_{gas,cal}} \tag{3.64}$$

The mass fraction of SO_2, $w_{SO2,cal}$ in the calciner exit gas is

$$w_{SO2,cal} = \frac{\dot{m}_{SO2,cal}}{\dot{m}_{gas,cal}} \tag{3.65}$$

The mass fraction of H_2O, $w_{H2O,cal}$ in the calciner exit gas is

$$w_{H2O,cal} = \frac{\dot{m}_{H2O,cal}}{\dot{m}_{gas,cal}} \tag{3.66}$$

The mass fraction of CO_2, $w_{CO2,cal}$ in the calciner exit gas is

$$w_{CO2,cal} = \frac{\dot{m}_{CO2,cal}}{\dot{m}_{gas,cal}} \tag{3.67}$$

The total gas volume flow rate at the calciner outlet is

$$\dot{V}_{gas,cal} = \dot{V}_{O2,cal,exc} + \dot{V}_{N2,cal} + \dot{V}_{SO2,cal} + \dot{V}_{H2O,cal} + \dot{V}_{CO2,cal} \tag{3.68}$$

And,

$$\dot{V}_{O2,cal,exc} = \dot{m}_{O2,cal,exc} \cdot \rho_{O2} \tag{3.69}$$

$$\dot{V}_{N2,cal} = \dot{m}_{N2,cal} \cdot \rho_{N2} \tag{3.70}$$

$$\dot{V}_{SO2,cal} = \dot{m}_{SO2,cal} \cdot \rho_{SO2} \tag{3.71}$$

$$\dot{V}_{H2O,cal} = \dot{m}_{H2O,cal} \cdot \rho_{H2O} \tag{3.72}$$

$$\dot{V}_{CO2,cal} = \dot{m}_{CO2,cal} \cdot \rho_{CO2} \tag{3.73}$$

And,

ρ_{O2} is density of O_2, calculated as

$$\rho_{O2} = \frac{P.M_{O2}}{R.T} \tag{3.74}$$

ρ_{N2} is density of N_2, calculated as

$$\rho_{N2} = \frac{P.M_{N2}}{R.T} \tag{3.75}$$

ρ_{SO2} is density of SO_2, calculated as

$$\rho_{SO2} = \frac{P.M_{SO2}}{R.T} \tag{3.76}$$

ρ_{H2O} is density of H_2O, calculated as

$$\rho_{H2O} = \frac{P.M_{H2O}}{R.T} \tag{3.77}$$

ρ_{H2O} is density of CO_2, calculated as

$$\rho_{CO2} = \frac{P.M_{CO2}}{R.T} \tag{3.78}$$

Where,

$P = Calciner\ pressure\ (design\ basis\ value)$

$P = Calciner\ exit\ gas\ temperature\ (design\ basis\ value)$

$R =$ gas constant 8.314

The volume fraction of O_2 in the calciner gas, $v_{O2,rk}$ is calculated by

$$v_{O2,rk} = \frac{\dot{V}_{O2,rk}}{\dot{V}_{gas,rk}} \tag{3.79}$$

The O_2 volume fraction at dry conditions in the kiln inlet, $v_{O2,rk,dry}$ which should be equal to the design basis value, is calculated by:

$$v_{O2,rk,dry} = \frac{\dot{V}_{O2,rk}}{\dot{V}_{gas,rk} - \dot{V}_{H2O}} \tag{3.80}$$

The total gas out from calciner $\dot{V}_{gas,cal,tot}$ is the sum of rotary kiln exit gas $\dot{V}_{gas,rk}$ plus the calciner exit gas $\dot{V}_{gas,cal}$ as below

$$\dot{V}_{gas,cal,tot} = \dot{V}_{gas,cal} + \dot{V}_{gas,rk} \tag{3.81}$$

3.3 Energy balance

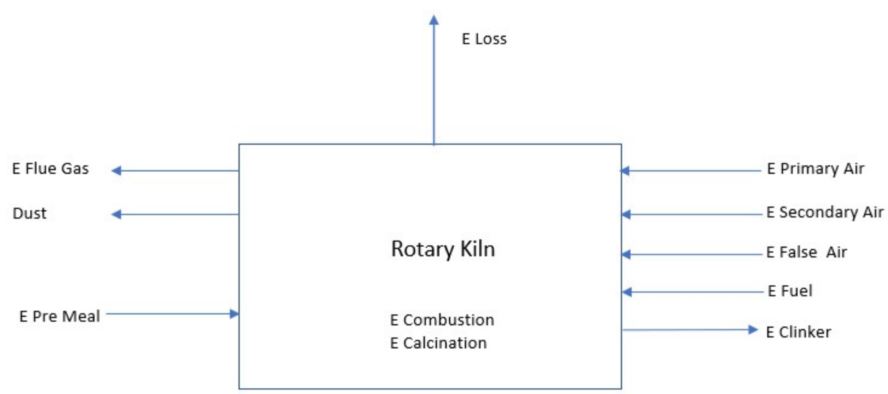

Figure 3.2 Energy balance flow diagram

Figure 3.2 illustrates the energy balance around the rotary kiln. According to the 1st thermodynamic law about energy conservation, the sum of energy in to the rotary kiln, $E_{in,rk}$, plus the generated energy inside the rotary kiln, $E_{gen,rk}$, is equal with energy out from rotary kiln, $E_{out,rk}$, or can be expressed as

$$E_{in,rk} + E_{gen,rk} = E_{out,rk} \tag{3.82}$$

Refer to figure 3.2, The energy in $E_{in,rk}$ is the sum of energy in, as expressed as

$$E_{in,rk} = E_{meal,cal} + E_{prim} + E_{sec} + E_{false} + \; + E_{fuel} \tag{3.83}$$

Where,

$E_{meal,cal}$ = Energy of precalcined meal (kJ/h)

E_{prim} = Energy of primary air (kJ/h)

E_{sec} = Energy of secondary air (kJ/h)

E_{false} = Energy of false air (kJ/h)

E_{fuel} = Energy of fuel (kJ/h)

Meanwhile the energy generated $E_{gen,rk}$is the expressed as

$$E_{gen,rk} = E_{comb} + E_{post\ cal} \tag{3.84}$$

Where,

E_{comb} = Energy generation due to combustion (kJ/h)

$E_{post\ cal}$ = Energy generation due to post calcination (kJ/h)

$E_{other,rk}$ = Energy generation due to other clinker related reaction (kJ/h)

Energy out is the sum of energy below

$$E_{out,rk} = E_{clinker} + E_{gas,rk} + E_{Dust} + E_{loss} \tag{3.85}$$

Where,

$E_{clinker}$ = Energy of clinker (kJ/h)

$E_{gas,rk}$ = Energy of flue gas out from rotary kiln (kJ/h)

E_{Dust} = Energy of dust (kJ/h)

E_{loss} = Energy loss from rotary kiln shell to surrounding (kJ/h)

All the energy is referred to reference condition, therefore the energies are calculated as:

$$E_{meal,cal} = \dot{m}_{meal,cal} \,.\, C_{p\ meal,cal} \,.\, (T_{cal} - T_{ref}) \tag{3.86}$$

$$E_{prim} = \dot{m}_{prim} \,.\, C_{p\ prim} \,.\, (T_{prim} - T_{ref}) \tag{3.87}$$

$$E_{sec} = \dot{m}_{sec} \, . \, C_{p\,sec} \, . \, (T_{sec} - T_{ref}) \tag{3.88}$$

$$E_{false} = \dot{m}_{false} \, . \, C_{p\,false} \, . \, (T_{false} - T_{ref}) \tag{3.89}$$

$$E_{clinker} = \dot{m}_{clinker} \, . \, C_{p\,clinker} \, . \, (T_{clinker} - T_{ref}) \tag{3.90}$$

$$E_{gas,rk} = \dot{m}_{gas,rk} \, . \, C_{p\,gas,rk} \, . \, (T_{gas,rk} - T_{ref}) \tag{3.91}$$

$$E_{Dust} = \dot{m}_{Dust} \, . \, C_{p\,Dust} \, . \, (T_{Dust} - T_{ref}) \tag{3.92}$$

For energy generation term, are expressed as

$$E_{comb} = \dot{m}_{fuel,rk} \, . \, LHV_{fuel,rk} \tag{3.93}$$

$$E_{post\,cal} = \dot{m}_{CO2\,post\,cal} \, . \, H_{cal} \tag{3.94}$$

For simplification of calculation, $C_{p\,Dust}$ and $C_{p\,clinker}$ is estimated as 1 kJ/kg of dust and clinker respectively. $C_{p\,meal,cal}$ is defined effective C_p where the value is adjusted to balance the energy in the rotary kiln system. Meanwhile energy loss based on experienced is assumed to be 6 MW for all cases.

4 Impact of using oxygen as a partial replacement for air in rotary kiln

This chapter is about evaluation of the impact of using oxygen as a partial replacement for air combustion especially on flame temperature, fuel rate, pure oxygen flow rate, and production capacity of cement kiln.

4.1 Impact on adiabatic flame temperature

Impact of using oxygen as a partial replacement for air combustion on adiabatic flame temperature is calculated using average heat capacity method follows the formula 2.15 in chapter 2. It is found that the adiabatic temperature increases with the increase of oxygen fraction in combustion air as shown in the table 3.1 below.

Table 3.1 Theoretical adiabatic flame temperature of oxygen enriched air combustion

Oxygen %	Adiabatic Flame Temperature (C)	
	Coal	Waste fuel
21	2649	2465
25	2878	2668
30	3137	2892
33	3288	3022
40	3557	3253
43	3668	3348

That above adiabatic flame temperature are based on combustion air inlet temperature at 881^{o}C. However adibatic flame temperature is maximum theoritical temperature can be attained during the combustion that only considering the flue gas of the combustion. In fact, the energy released from fuel combustion is also absorbed by solid material, gas in surrounding other than flue gas, and combustion chamber wall – that will restrain the temperature of the system to reach the adiabatic flame temperature.

Spreadsheet for calculation of adibatic flame temperature is attached in appendix B.

4.2 Impact on fuel rate

The fuel consumption rate of kiln as depicted by figure 4.1 and 4.3 – fuel consumption rate of constant capacity case below is in line with the basic idea of oxygen enriched air combustion theory where the decrease of the flue gas flow rate would increase the fuel efficiency. On coal fuel case, kiln fuel consumption rate decreases from 6.3 t/h to 5.8 t/h. Meanwhile, on waste fuel case, kiln fuel consumption rate decrease from 7.1 t/h to 6.4 t/h.

Whereas on capacity increase cases as illustrated by figure 4.2 and 4.4 below the fuel consumption rate increases is so small or almost constant on both coal and waste fuel cases. On coal fuel case, the fuel consumption is almost constant on 6.3 t/h meanwhile on waste fuel case, the fuel consumption rate is almost constant on 7.9 t/h except below O_2 level 23%. Regarding the sharp changes below O_2 level 23% as shown on waste fuel graph figure 4.4 is related to the capacity changes that will be elaborated more in the next part.

One of the draw-back of the oxygen enriched combustion on calciner is less energy brought by the kiln exit gas to calciner compared to the conventional combustion since the kiln exit gas flow rate of oxygen enriched combustion is lower than conventional combustion. Therefore, some additional fuel needs to be added to calciner to compensate the energy shortage. This effect is shown by figure 4.1 to figure 4.4 below

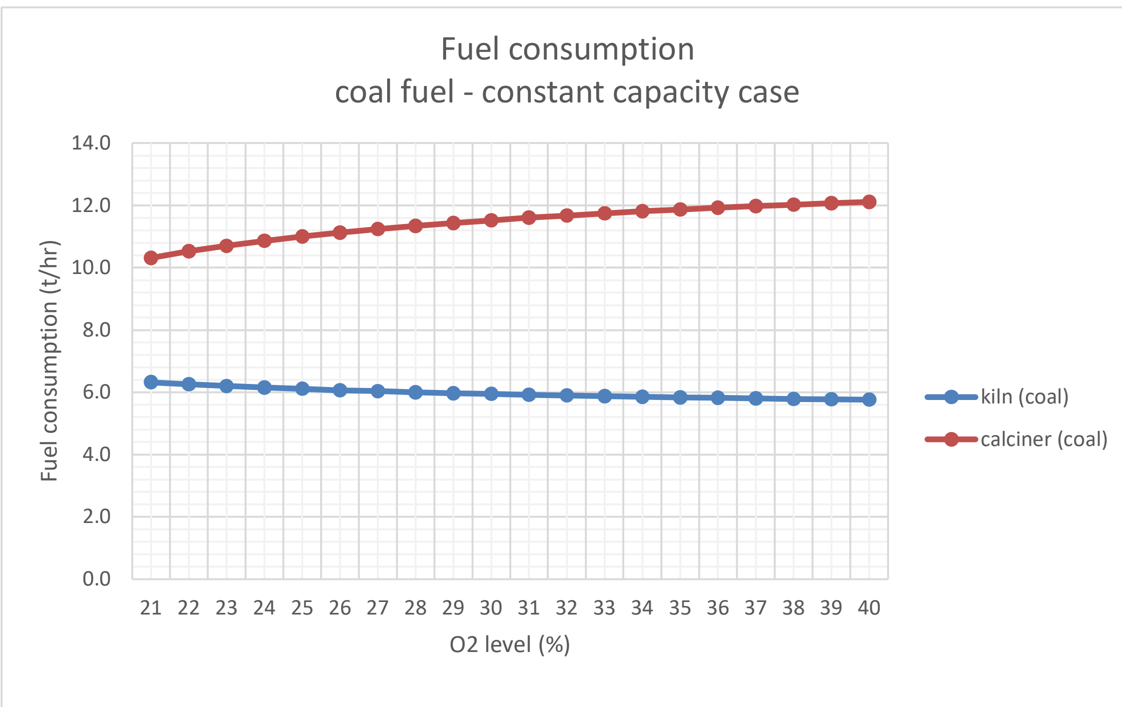

Figure 4.1 fuel consumption rate on oxygen enriched air combustion of coal fuel – constant capacity case

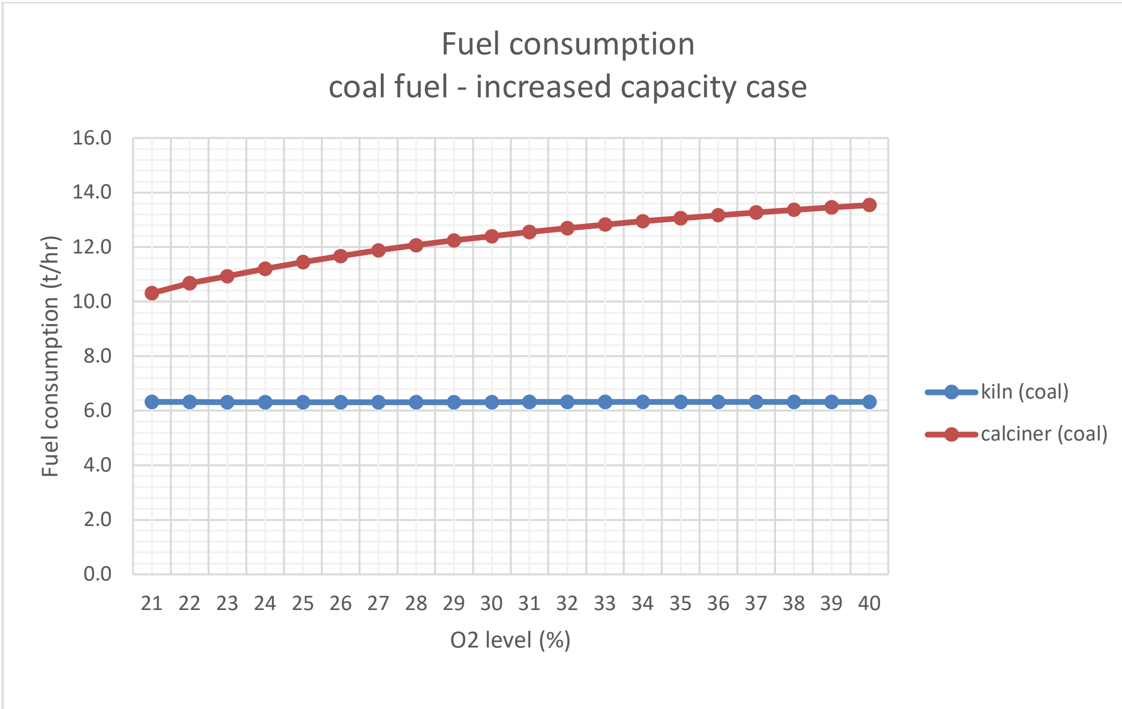

Figure 4.2 fuel consumption rate on oxygen enriched air combustion of coal fuel – increased capacity case

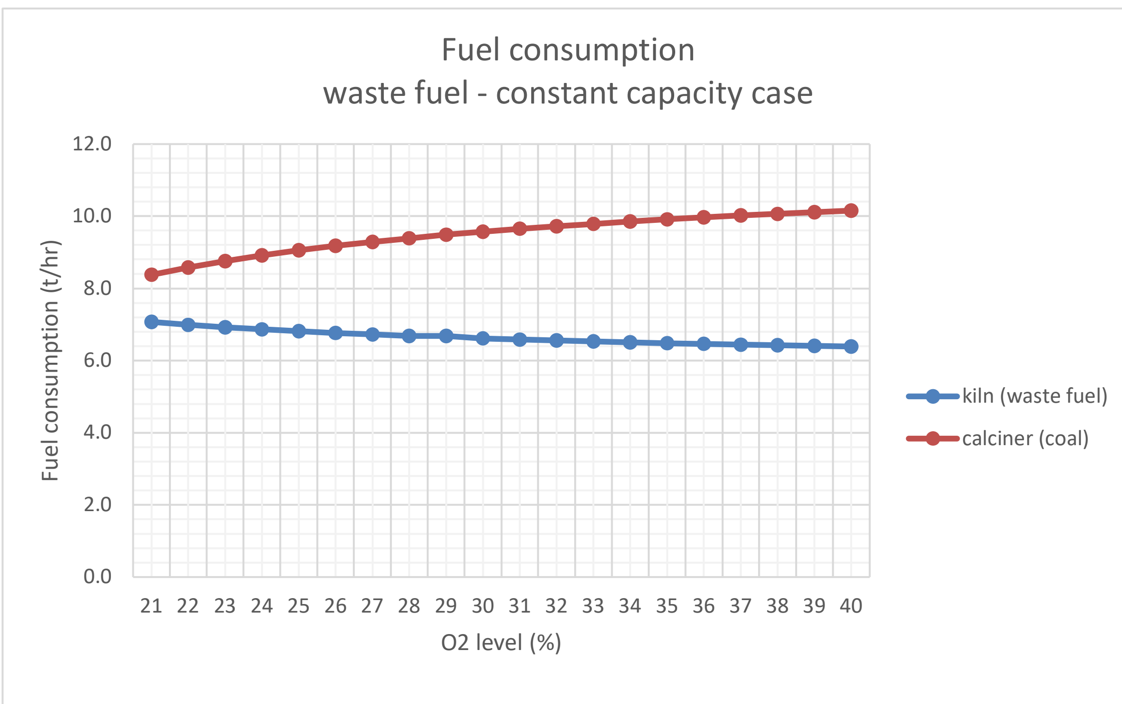

Figure 4.3 fuel consumption rate on oxygen enriched air combustion of waste fuel – constant capacity case

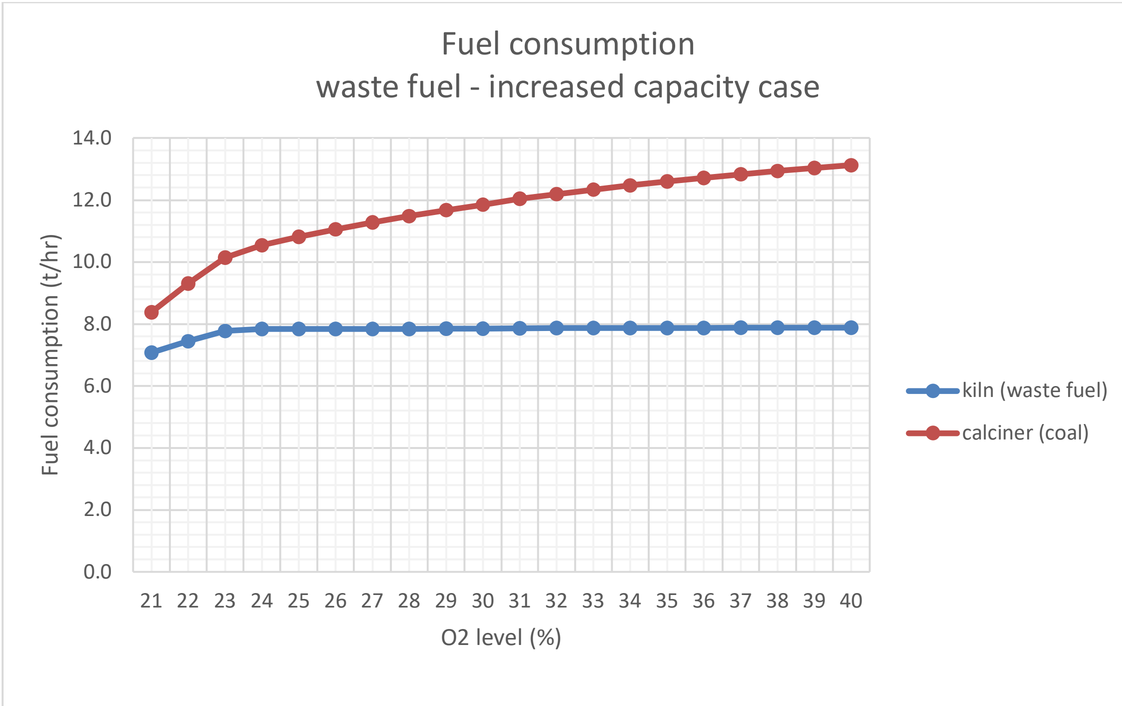

Figure 4.4 fuel consumption rate on oxygen enriched air combustion of waste fuel – constant capacity case

4.3 Impact on exit gas flow rate

Kiln exit gas flow rate is a product of combustion flue gas plus the carbon dioxide produced during calcination. For calciner, the exit gas is the total of calciner exit gas plus the kiln exit gas.

As shown on figure 4.5 and 4.6 for coal constant capacity case and by figure 4.9 and 4.10 for waste fuel constant capacity case, the exit gas flow rate on both kiln and calciner decreases following the decrease of the fuel rate as describe on previous part. On coal case, the exit gas of kiln drops from 304348 m3/h (79687 kg/h) to 146725 m3/h (41299 kg/h) and from 872095 m3/h (277525 kg/h) to 776665 m3/h (260989 kg/h) for calciner. Meanwhile on waste fuel case, the kiln exit gas drops from 304884 m3/h (78372 kg/h) to 148786 m3/h (40434 kg/h) and from 777307 m3/h (242910 kg/h) to 684416 m3/h (227096 kg/h) for calciner.

On capacity increased cases of both coal fuel and waste fuel, as shown by figure 4.7 and 4.8 for coal and by figure 4.11 and figure 4.12 for waste fuel cases, the calciner exit gas is purposely set to be constant as base case i.e 872767 m3/h as the restriction for capacity increment except for waste for fuel case oxygen level below 24% where the restriction is kiln exit gas flow rate at 305021 m3/h. Meanwhile the kiln exit gas drop from 304348 m3/h (79687 kg/h) to 161201 m3/h (45392 kg/h) for coal case and from 304884 m3/h (78372 kg/h) to 183962 m3/h (50044 kg/h) for waste case.

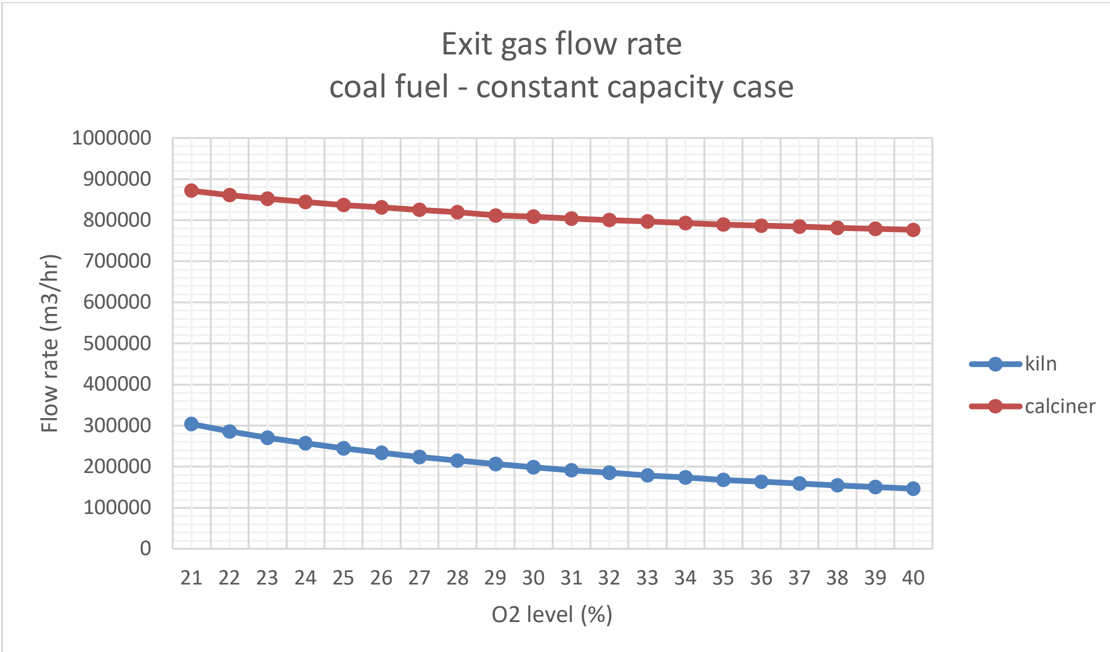

Figure 4.5 exit gas flow rate on oxygen enriched air combustion of coal fuel – constant capacity case

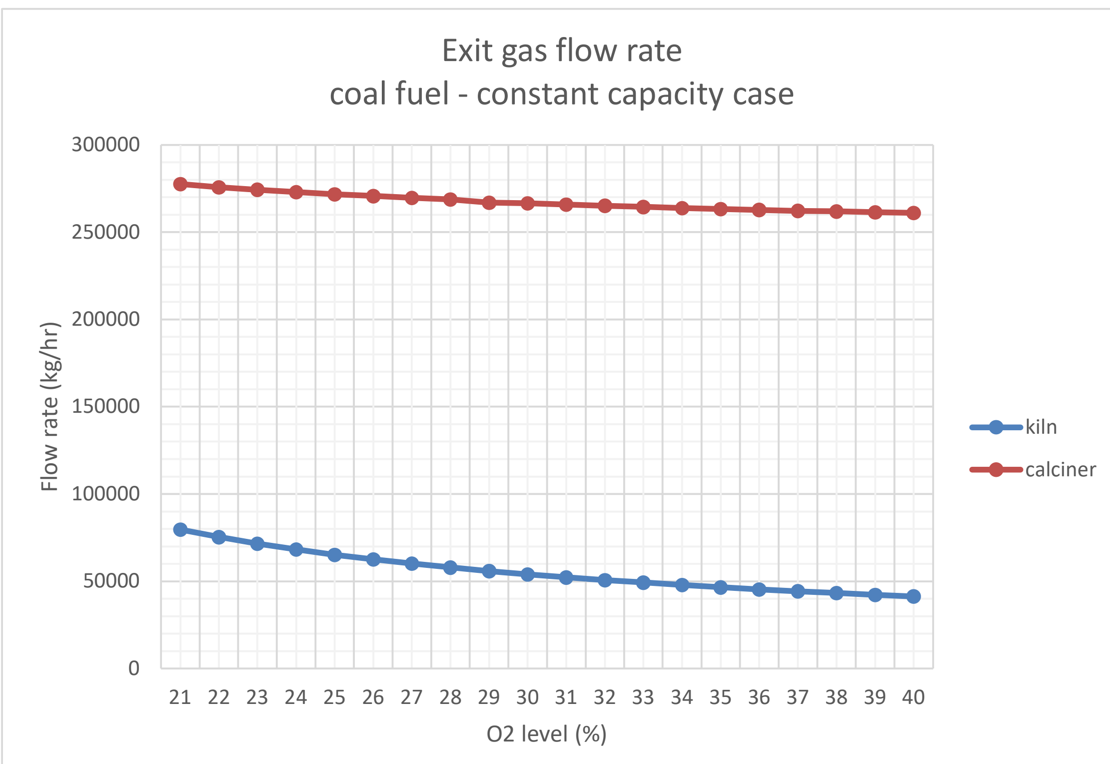

Figure 4.6 exit gas flow rate on oxygen enriched air combustion of coal fuel – constant capacity case

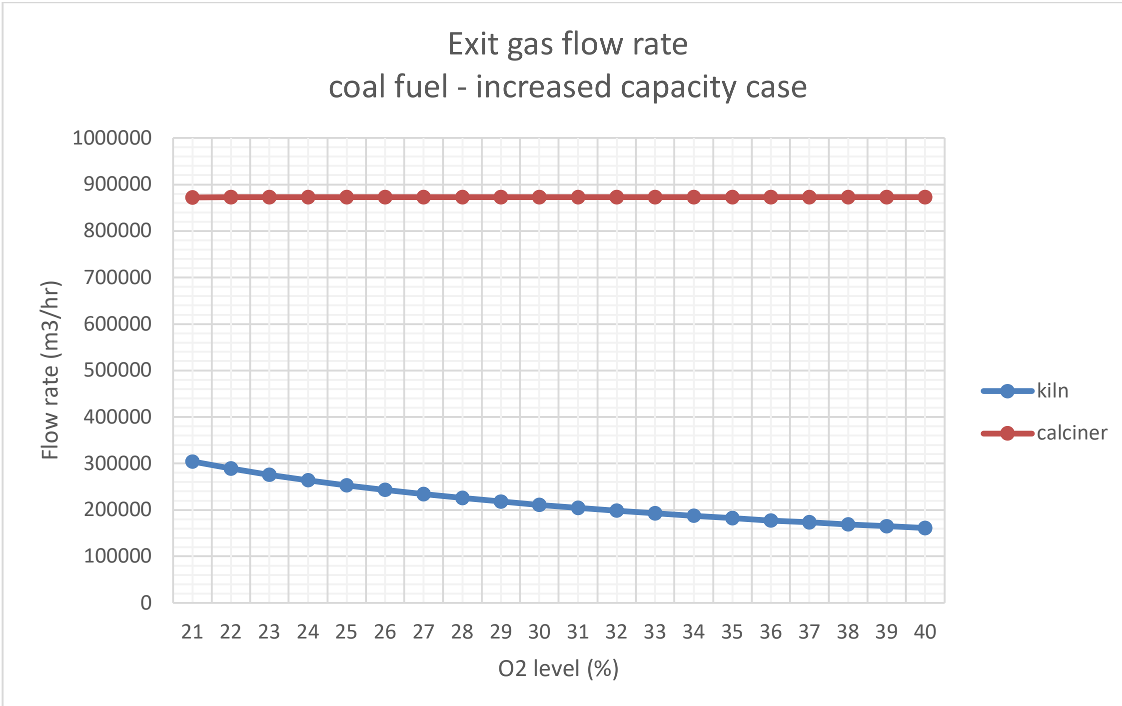

Figure 4.7 exit gas flow rate on oxygen enriched air combustion of coal fuel – increased capacity case

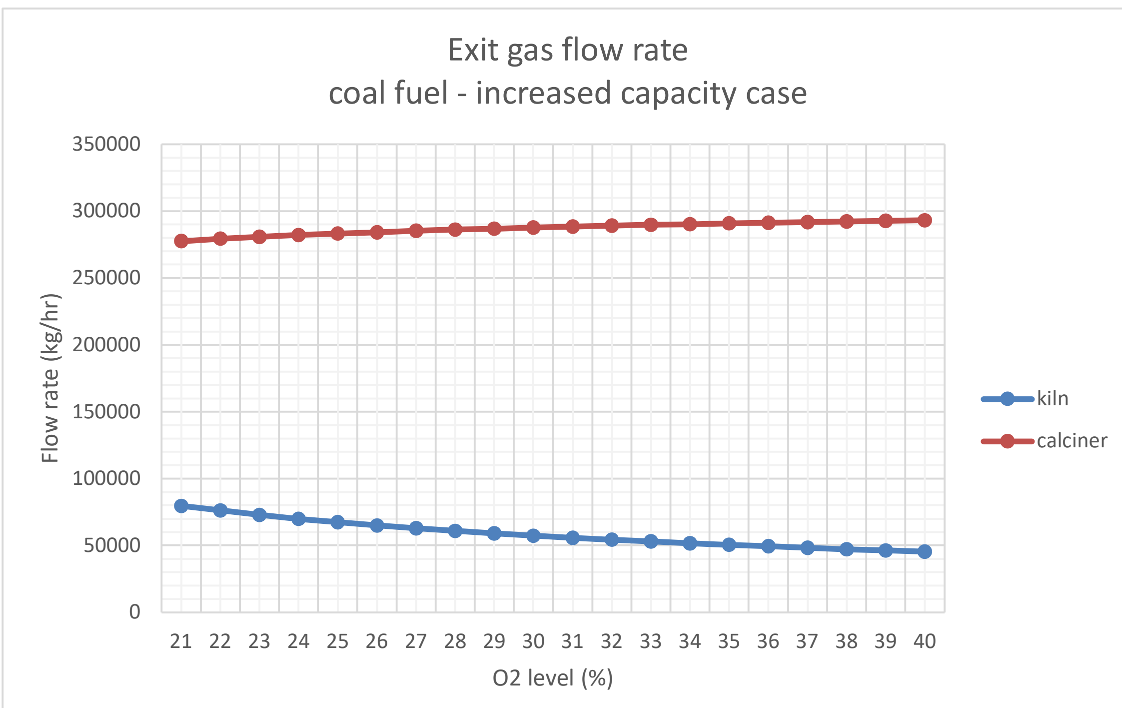

Figure 4.8 exit gas flow rate on oxygen enriched air combustion of coal fuel – increased capacity case

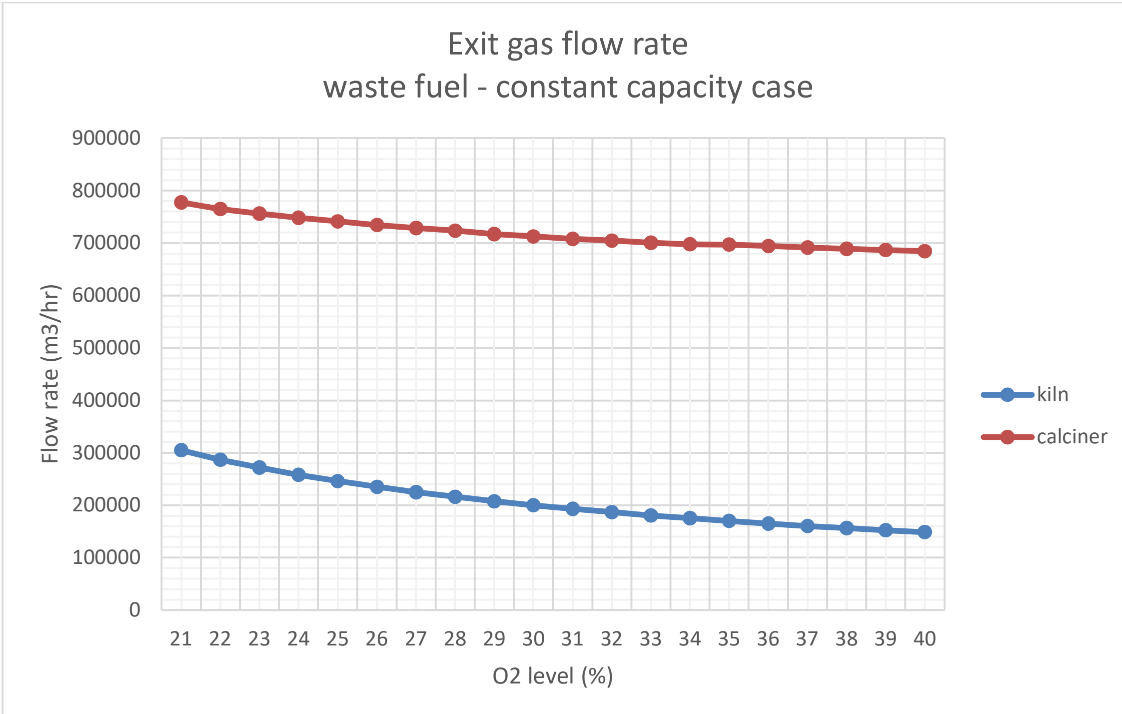

Figure 4.9 exit gas flow rate on oxygen enriched air combustion of waste fuel – constant capacity case

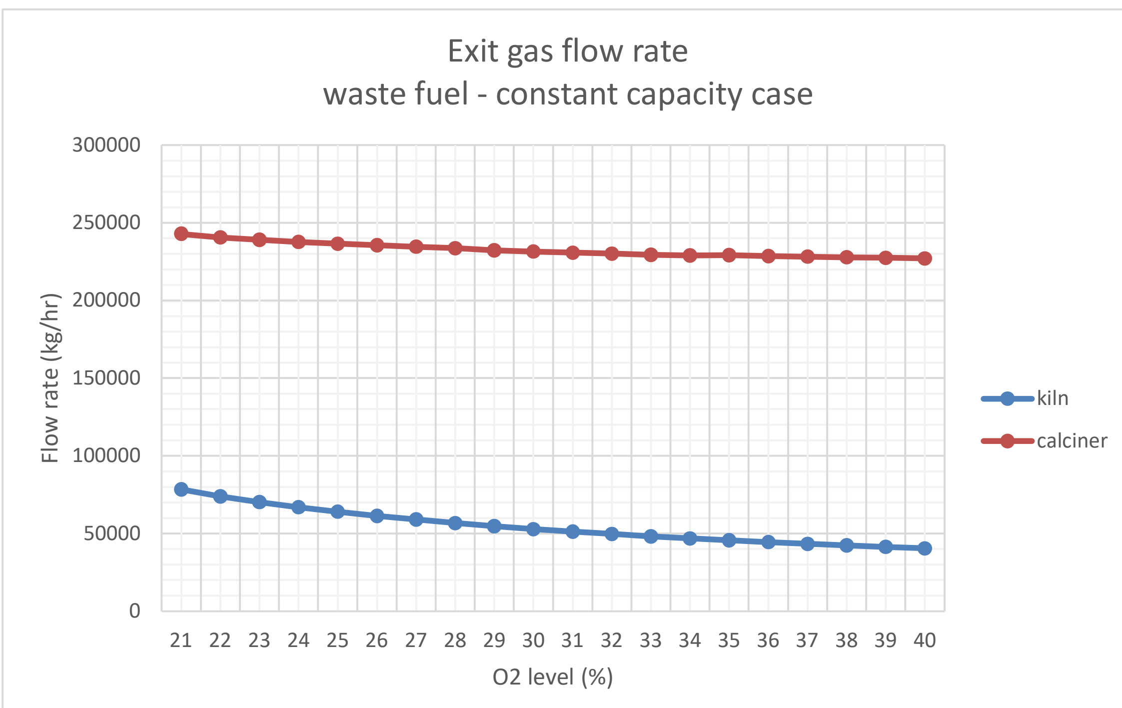

Figure 4.10 exit gas flow rate on oxygen enriched air combustion of waste fuel – constant capacity case

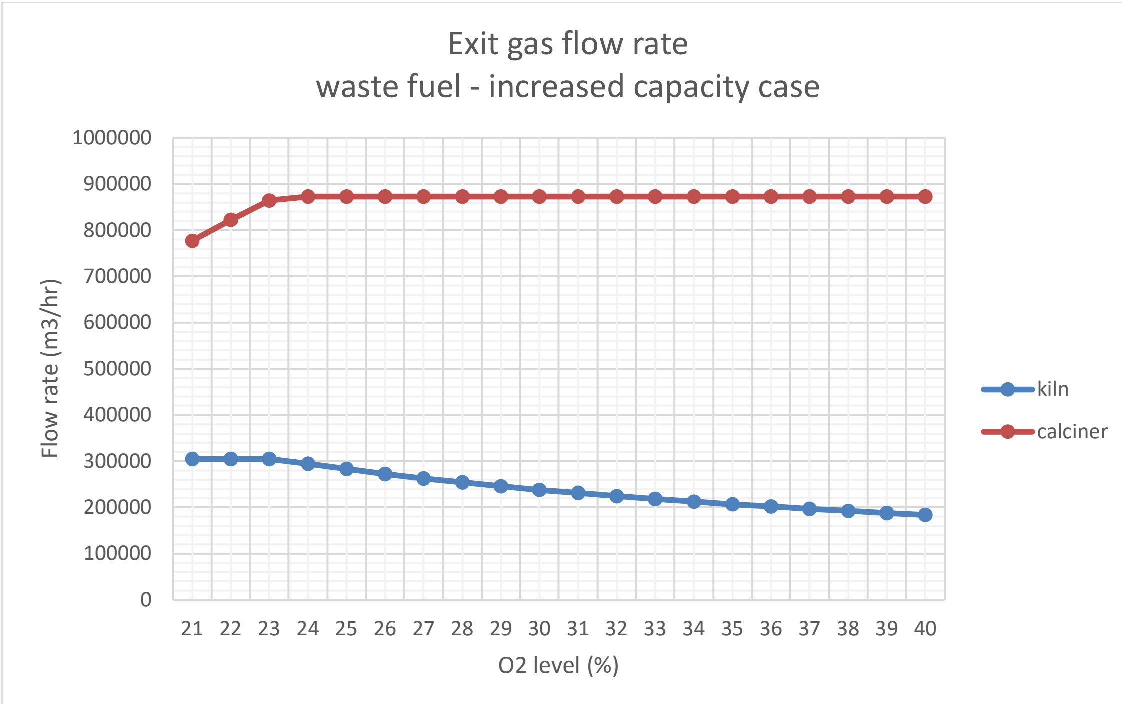

Figure 4.11 exit gas flow rate on oxygen enriched air combustion of waste fuel – increased capacity case

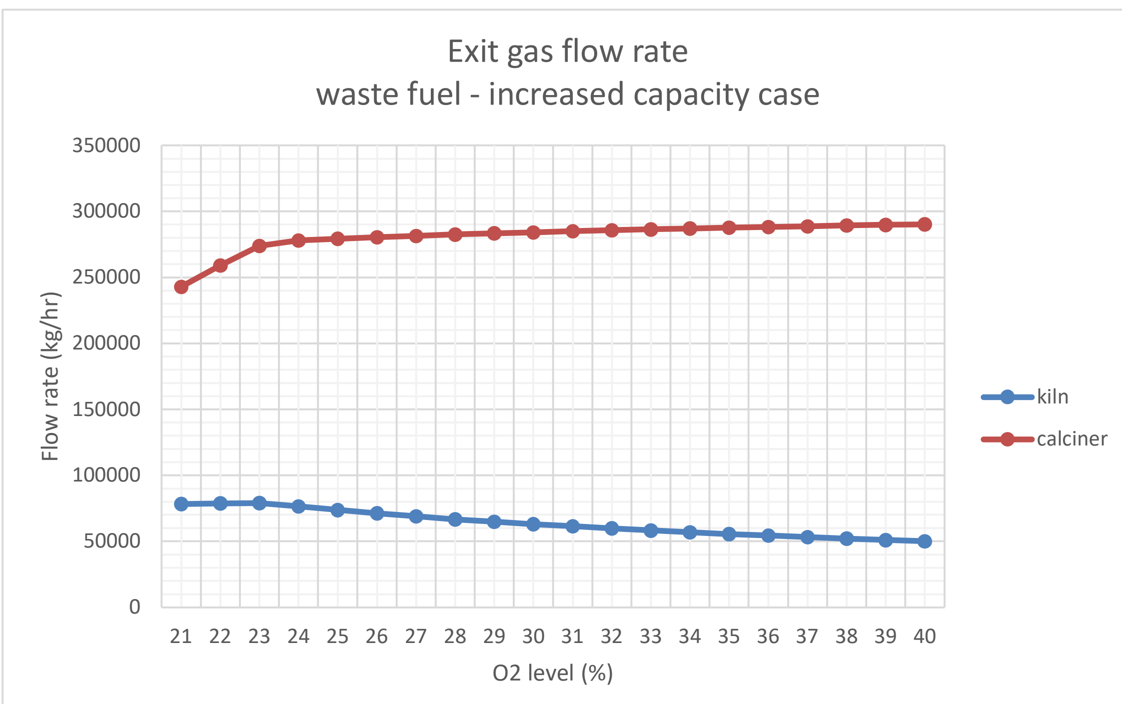

Figure 4.12 exit gas flow rate on oxygen enriched air combustion of waste fuel – increased capacity case

4.4 Impact on kiln exit gas temperature

Kiln exit gas temperature is maintained 1150ºC on all cases, meanwhile the calciner is set at 900 ºC.

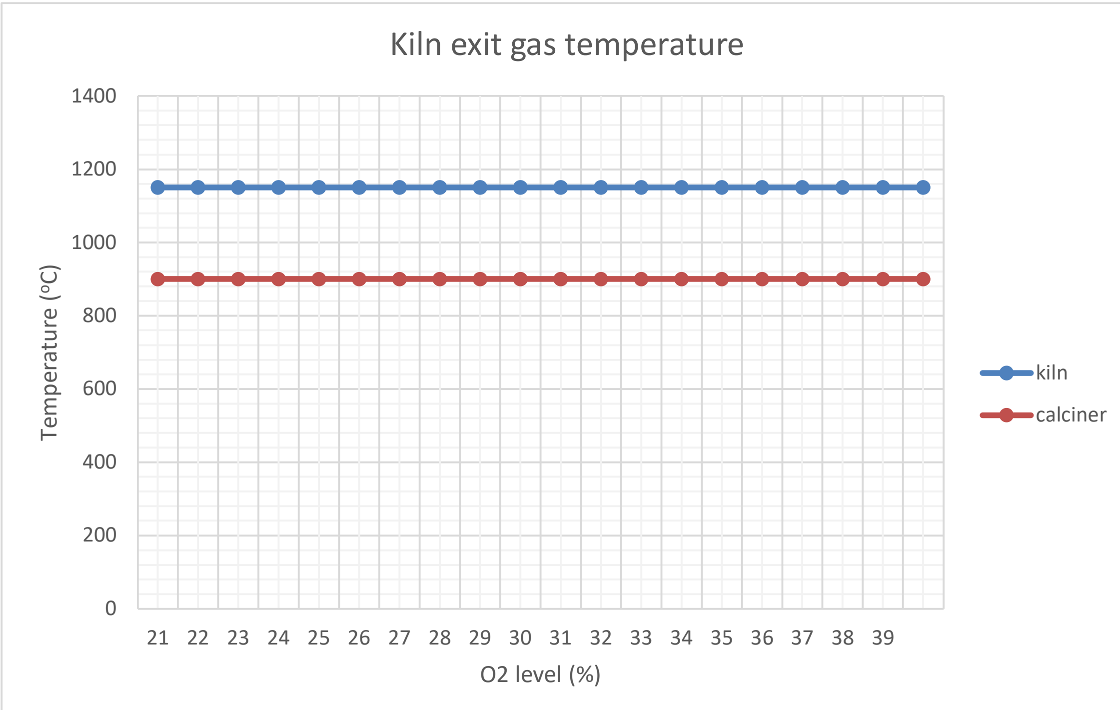

Figure 4.13 kiln exit gas temperature on oxygen enriched air combustion

4.5 Impact on exit gas composition

Oxygen gas concentration is adjusted at 3% on all cases at both rotary kiln and calciner. Sulfur dioxide (SO_2) concentration is almost constant on all case around 0.3% at kiln exit gas and 0.2% at calciner exit gas.

The nitrogen is the most exit gas component which changed in concentration beside of carbon dioxide due to oxygen enriched air combustion and calcination process effect. The concentration of the nitrogen of all cases is almost the same on both coal fuel and waste fuel case. The nitrogen concentration on kiln exit gas plummets from 68% to 44% and from 53% to 47% on the calciner exit gas.

On coal fuel cases, the carbon dioxide concentration on kiln exit gas increases from 26% to 47% and from 41% to 47% on the calciner exit gas. Meanwhile on waste fuel cases, the carbon dioxide concentration on kiln exit gas increases from 25% to 44% and from 40% to 47% on the calciner exit gas.

Water concentration of coal fuel case increases from 3% to 5% on kiln exit gas and from 2.3% to 2.5% on calciner exit gas. Meanwhile on waste fuel case, the water concentration increases from 5% to 9% on kiln exit gas and from 2.9% to 3.1% on calciner exit gas. For more detail of kiln and calciner exit gas composition, can be seen in the figure 4.14, to figure 4.21 below.

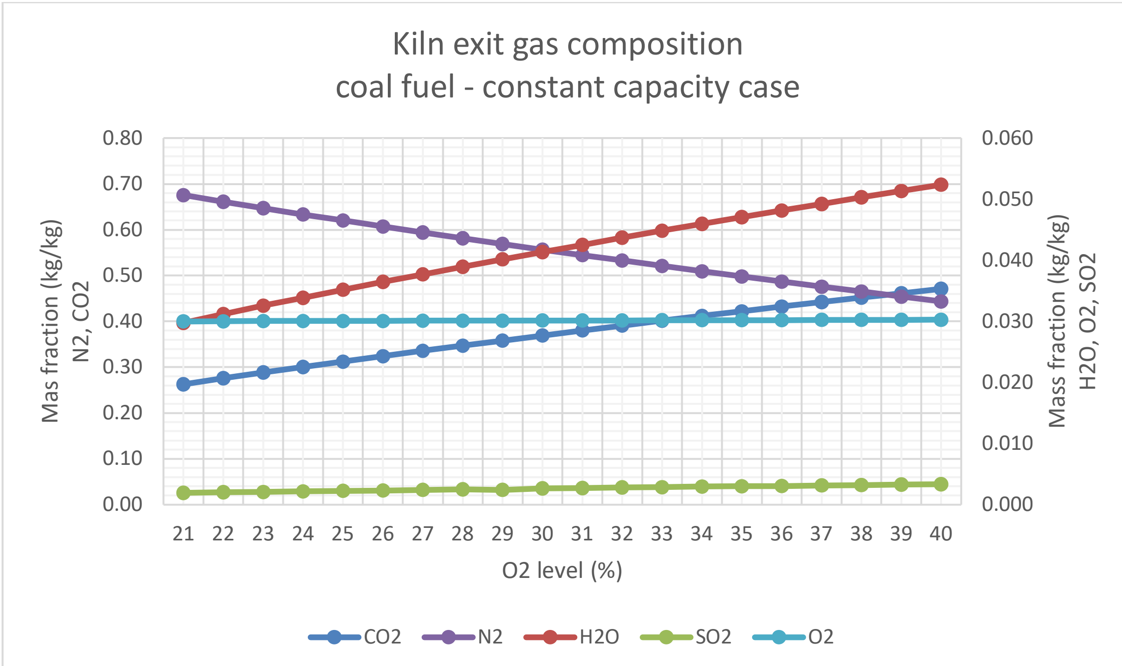

Figure 4.14 kiln exit gas composition on oxygen enriched air combustion of coal fuel – constant capacity case

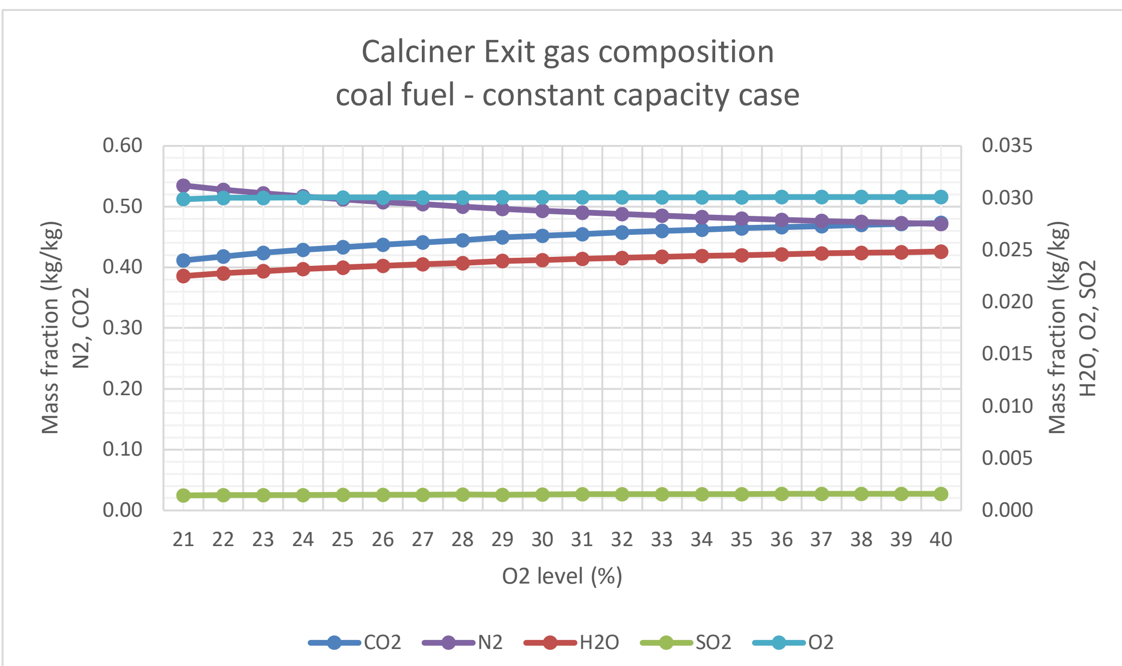

Figure 4.15 calciner exit gas composition on oxygen enriched air combustion of coal fuel – constant capacity case

Figure 4.16 kiln exit gas composition on oxygen enriched air combustion of coal fuel – increased capacity case

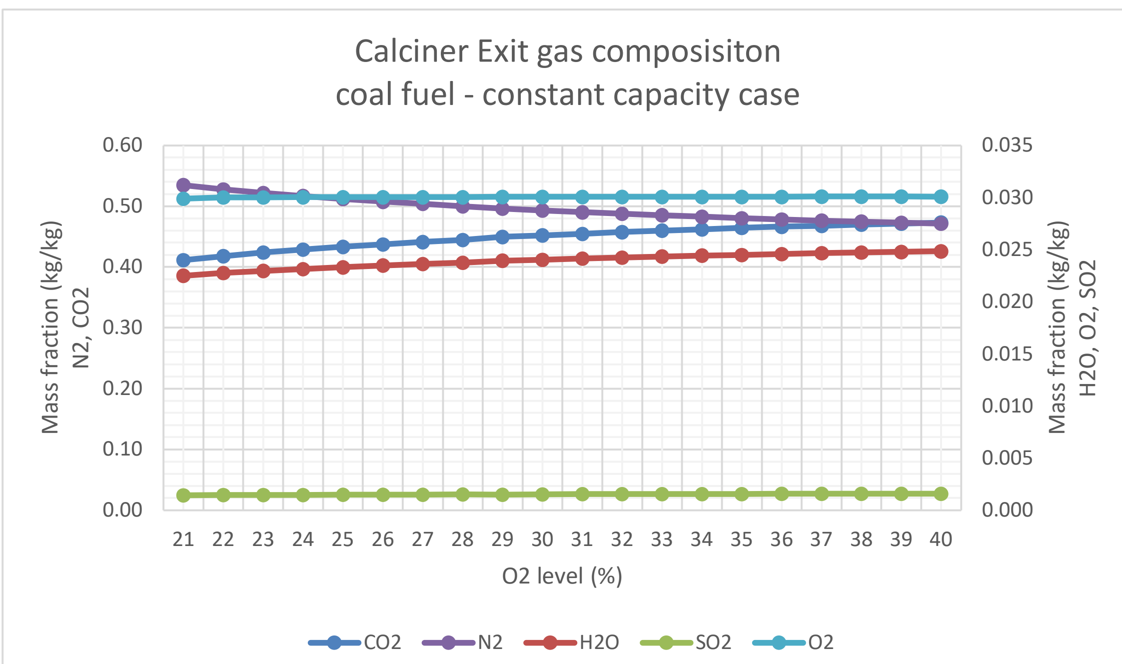

Figure 4.17 calciner exit gas composition on oxygen enriched air combustion of coal fuel – increased capacity case

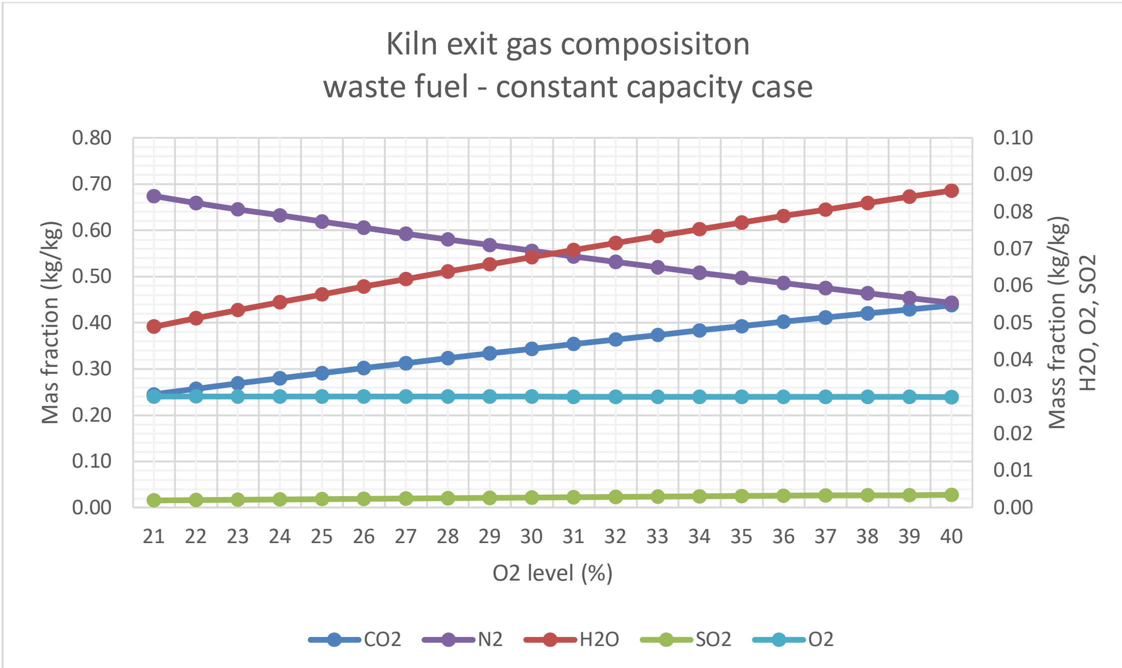

Figure 4.18 kiln exit gas composition of oxygen enriched air combustion of waste fuel – constant capacity case

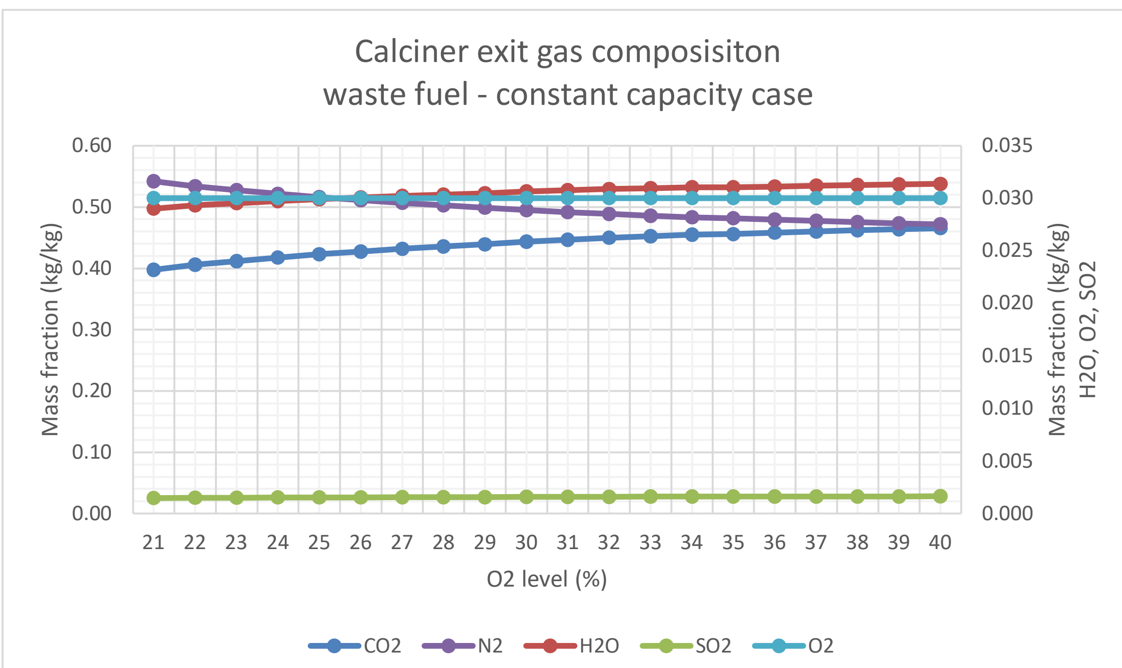

Figure 4.19 calciner exit gas composition on oxygen enriched air combustion of waste fuel – constant capacity case

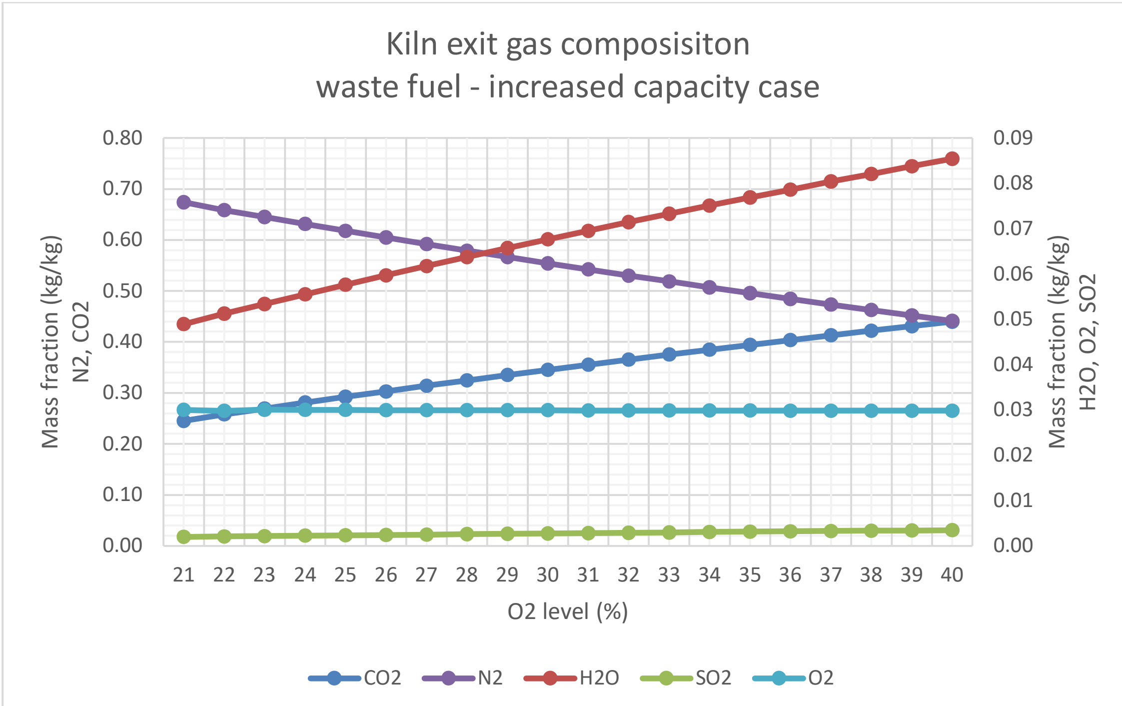

Figure 4.20 kiln exit gas composition on oxygen enriched air combustion of waste fuel – increased capacity case

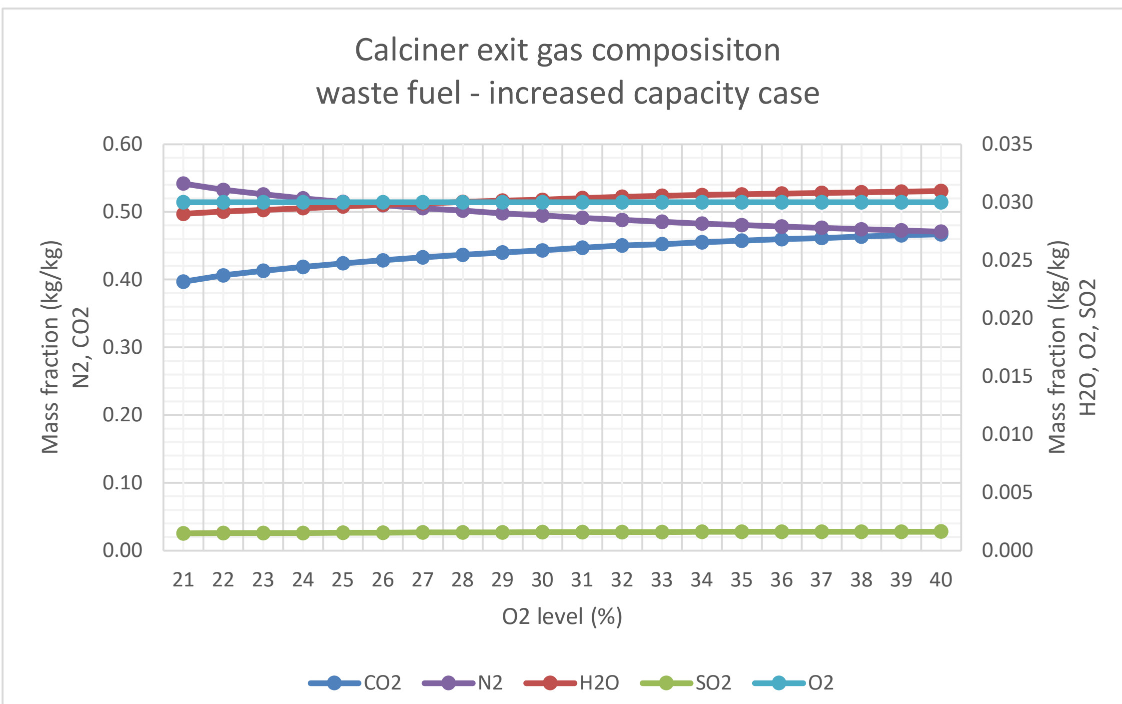

Figure 4.21 calciner exit gas composition on oxygen enriched air combustion of waste fuel – increased capacity case

4.6 Impact on production capacity

By having the decrease in exit gas flow rate due to the oxygen enriched air combustion, there will be a room for increasing the production capacity until the exit gas flowrate reach the same as the base case.

Capacity increase percentage Cap_{inc} (%) is defined as the ratio of the calculated capacity Cap_{cal} (t/h) over the base case capacity Cap_{base} (t/h) as below

$$Cap_{inc} = \frac{Cap_{cal}}{Cap_{base}} \tag{4.1}$$

On coal case, the production capacity increases gradually from 137 t/h to 153 t/h or increase up to 11.5% following oxygen level increment from 21% - 40% as shown by figure 4.22. Meanwhile on the waste fuel case, the production capacity increases from 118.3 t/h to 151.8 t/h or increase up to 10.8% relative to the base case following the same oxygen level increment as shown by figure 4.23.

As mentioned earlier in the previous chapter, for waste fuel case below 24% oxygen level, the restriction for production capacity increase is kiln exit gas flow rate instead of calciner exit gas flow rate, therefore there is significant changes on capacity below that range compare to increase on capacity at above 24% oxygen level which is more gradually.

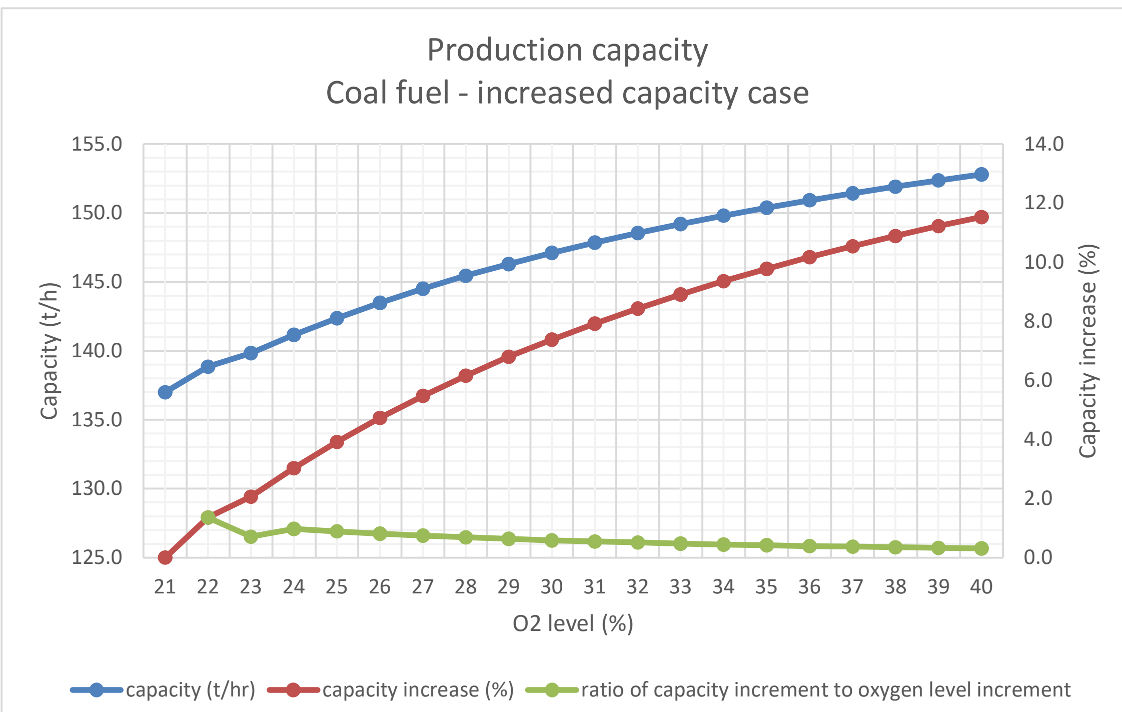

Figure 4.22 capacity increase on oxygen enriched air combustion of coal fuel

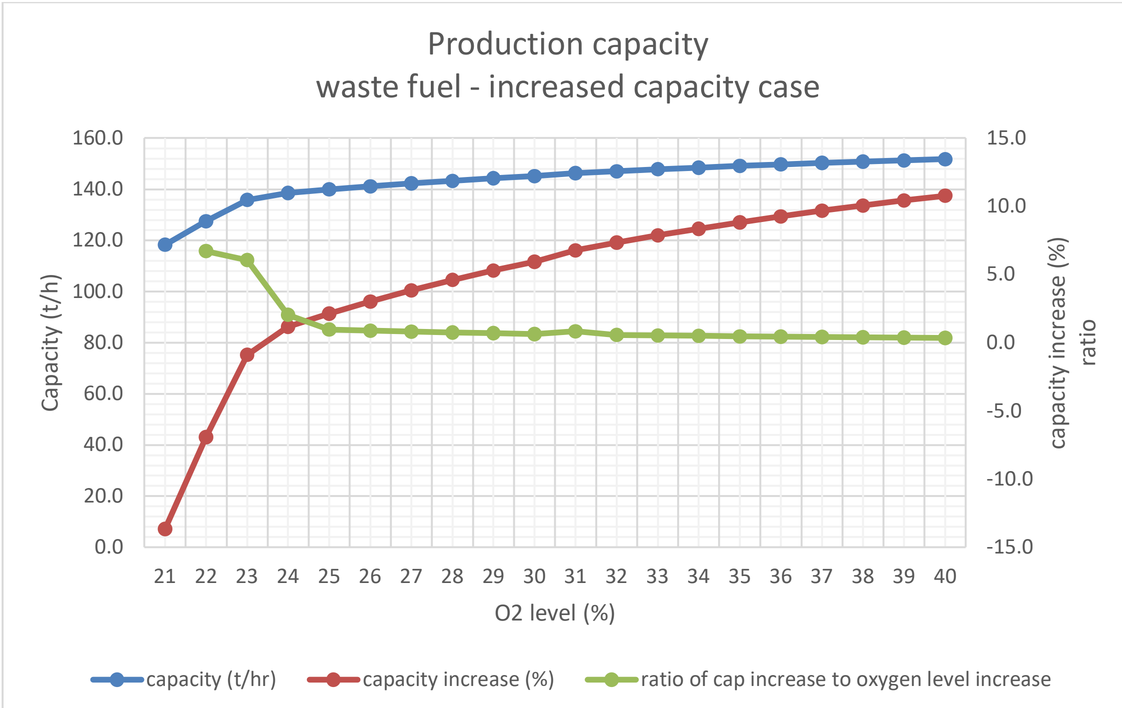

Figure 4.23 capacity increase on oxygen enriched air combustion of waste fuel

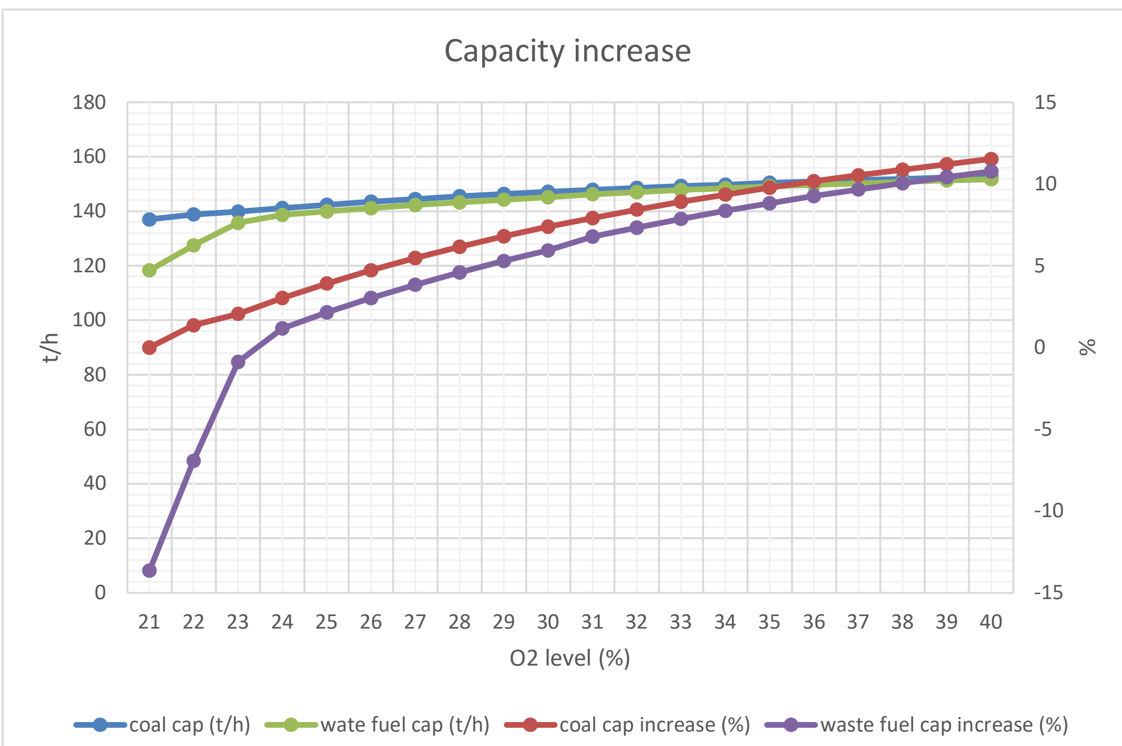

Figure 4.24 capacity increase on oxygen enriched air combustion

Figure 4.24 compares the coal and waste fuel capacity increases on oxygen enriched combustion. It can be seen that at oxygen level 24%, the capacity of the waste fuel case almost reaches the capacity of coal case, within the margin of 2 t/h.

The ratio of capacity increase to oxygen level increase decreases linearly with the oxygen level increament for both coal and waste fuel. The oxygen level of 22% is found as the optimum level in term of capacity increament againt the oxygen level increment.

4.7 Impact on pure oxygen flow rate

Pure oxygen flow rate increases gradually following the oxygen level increment form 21% - 40%. On coal fuel cases, the pure oxygen flow rate for constant capacity case and increased capacity case increases up to 6.1 t/h (44729 t/y) and 6.7 t/h (490480 t/y) respectively. Meanwhile on waste fuel cases, the pure oxygen flow rate increases up to 5.9 t/h (43426 t/y) and 7.3 t/h (53552 t/y) for constant capacity case and increased capacity case respectively.

The figure shown that the pure oxygen flow rate of capacity increase case is higher for waste fuel compare to coal, but for constant capacity case the pure oxygen flowrate is higher for coal compare to waste fuel.

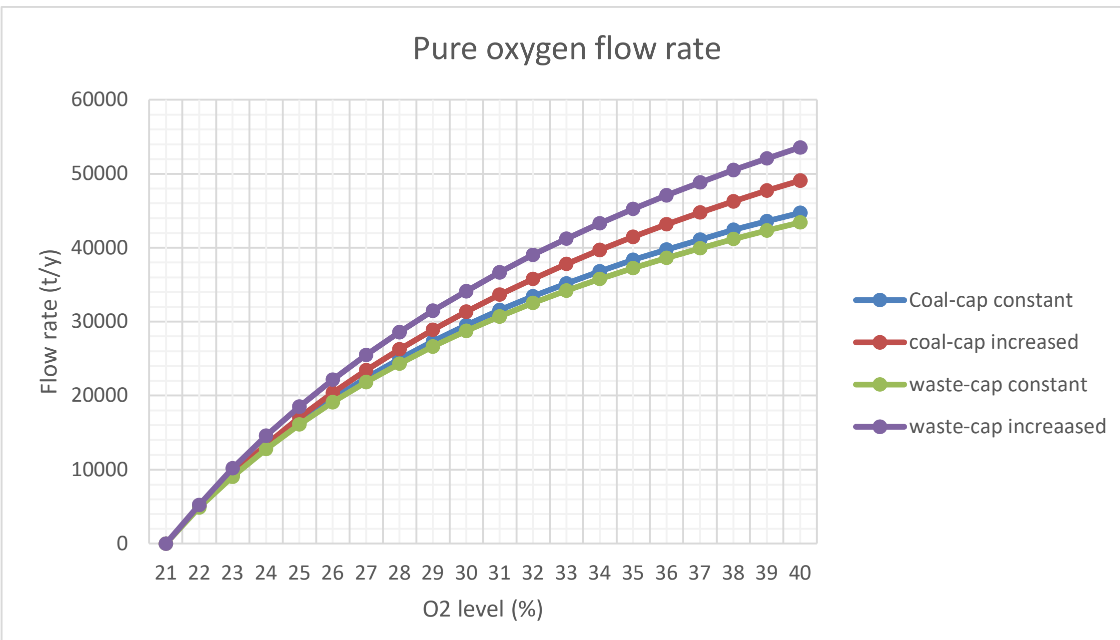

Figure 4.25 pure oxygen flow rate on oxygen enriched air combustion

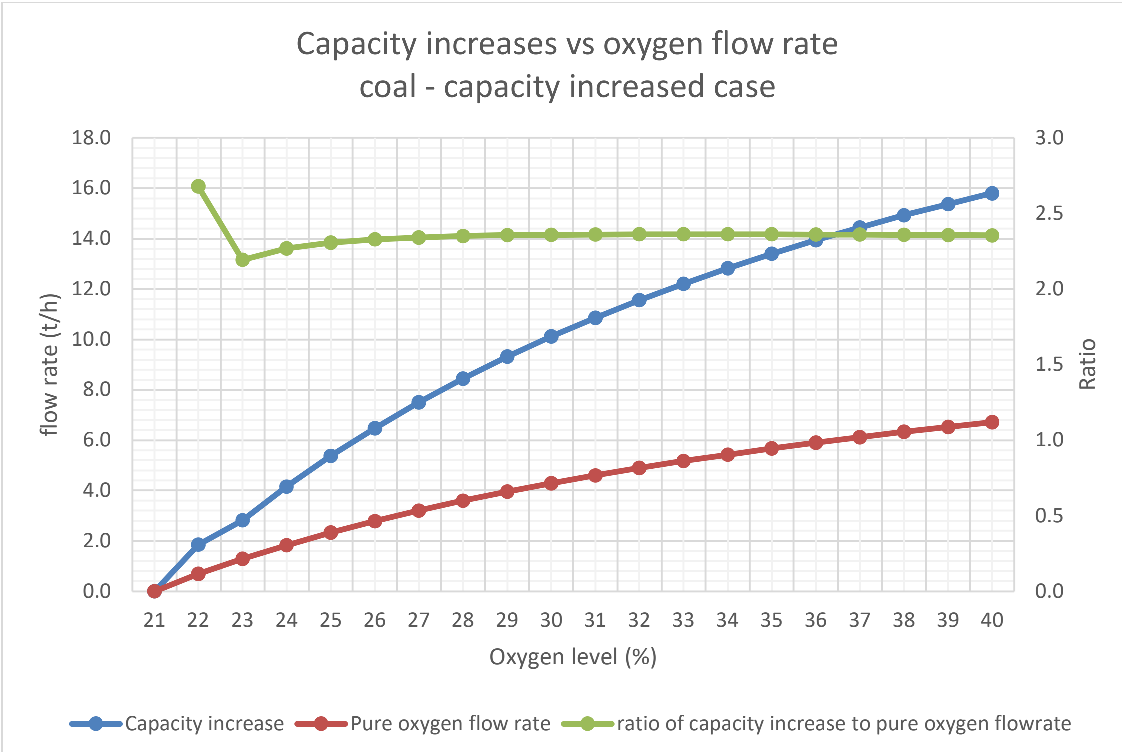

Figure 4.26 pure oxygen flow rate vs capacity increases – coal increased capacity case

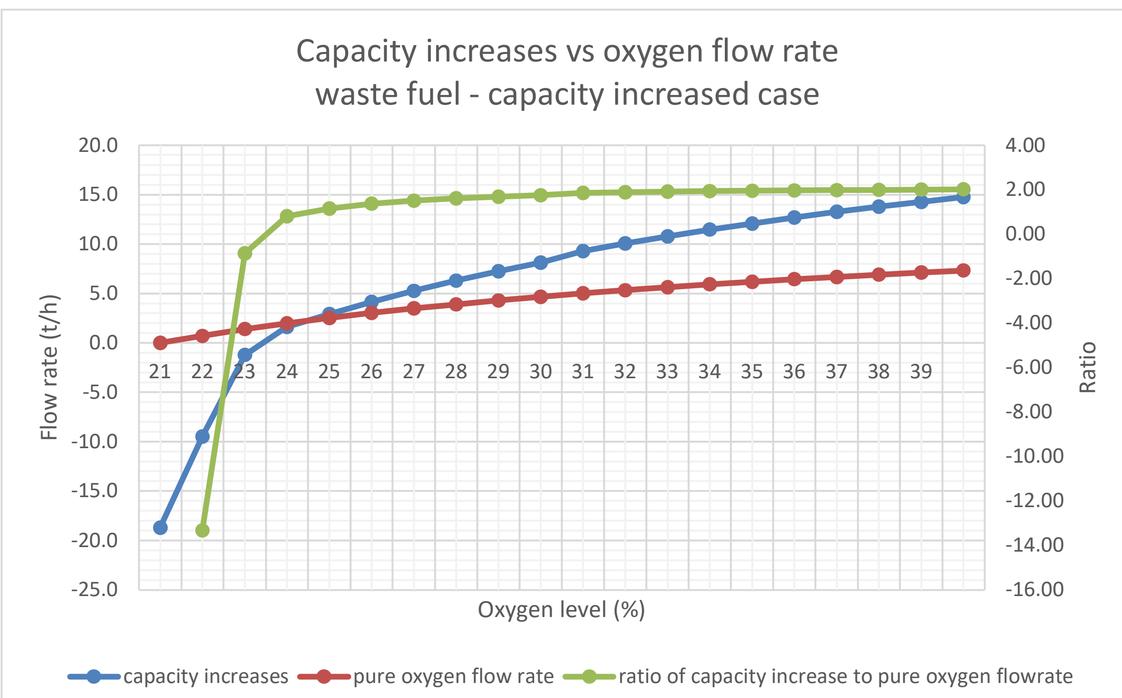

Figure 4.27 pure oxygen flow rate vs capacity increases – waste fuel increased capacity case

Figure 4.26 and 4.27 shows that, for coal case, the capacity increases 2.4 t/h for each t/h pure oxygen flow rate averagely. Meanwhile for waste fuel case, capacity increases 1.75 t/h for each t/h pure oxygen flow rate averagely. In term of ratio of capacity increases to pure oxygen flow rate, coal have a better output compared to waste fuel. But this result needs to be compared further by economic analysis that will be done in the next chapter.

Those ratios above are lower than what experienced by [2] where they got 4 t/h for each t/h pure oxygen, but it is needed more information to compare the result regarding the parameters and variables used by [2], which is insufficient provided by [2].

5 Constructional changes to kiln system

This chapter discussed about the constructional change deemed required regarding the implementation of oxygen enriched air combustion implementation.

5.1 Modification of refractory

Refractory is is a material that is resistant to decomposition by heat, pressure, or chemical attack, and retains strength and form at high temperature. There are varied type and shape of refractory available in the market for different application according to the severity of the condition.

Refractory is installed inside the rotary shell to protect the rotary kiln's metallic wall from overheating due to fuel combustion process inside the kiln. Refractory failure particularly is a common problem in rotary kiln, therefore regular replacement of refractory need to be done. Selection of appropriate refractory for rotary kiln shall consider the refractory resistance to high temperature, spalling, chemical attack, abrasion and its coatability [14].

Significant increase in flame temperature due to oxygen enriched air combustion required the higher temperature resistance refractory compared to the atmospheric air combustion. Based on the flame temperature calculation of oxygen enriched air combustion of coal at 30% oxygen level, the flame temperature can reach 3137ºC compared to 2649ºC at oxygen level of 21%. Assume this is as the worst-case condition, the super refractories class is required for example are zirconia (ZrO2), hafnium carbide (HfC). But in fact, during normal operational mode, the temperature inside the rotary kiln will not reach such that high temperature because of some of the heat is absorbed by the solid material inside the rotary kiln and by the kiln wall. Based on the information on [6], maximum operational temperature inside the rotary kiln is usually around 1500ºC. Therefore, it is no modification on the refractory seems required for now.

5.2 Modification of raw material delivery system

Since the implementation of this oxygen enriched air combustion is envisaged at the existing cement plant, major modification on the plant should be avoided. Capacity increase should be limited at 10% of the base capacity under assumption that the capacity allowance of whole plant is in order of 10%. Therefore, no modification of raw material delivery system seems required for capacity increase below 10%.

5.3 Modification of burner

Flame shape, heat transfer and temperature profile of the burner is out this thesis's scope. But referred to the studies done by others [10], [11], [12], there will be changes on the variables mentioned above on the oxygen enriched air combustion case. Existing burner is still can be used with some measures of inlet condition but some modification on the burner were also proposed to have the same condition as existing burner system.

5.4 Modification of air system

The principal modification required of oxygen enriched air combustion on kiln system is pure oxygen injection point. There are several possibilities of oxygen injection point around the kiln's system but comingle together with fuel supply is deemed as the most practical choice. Normally, fuel is delivered to the burner together with the primary air mainly in pneumatic supply system beside of the primary air is also functioned as cooler for the burner tip to prevent overheating at the burner tip.

Additional nozzle at rotary kiln body also seems as another option for pure oxygen injection point, particularly when the required pure oxygen flow rate exceeds the primary air flow rate.

Existing air compressor/blower flow rate also needs to be adjusted to get the appropriate air flow rate after balanced with pure oxygen injection.

5.5 Modification of fuel transfer system

Implementation of oxygen enriched air combustion create a room for capacity increase using the existing rotary kiln. Increasing in capacity is in line with increasing of fuel demand on the kiln and calciner. By assumption that the fuel transfer capacity allowance of the base case is within magnitude of 10%, the existing fuel system capacity needs to be upgraded to fulfill the increase of fuel rate above 10%.

5.6 Modification of clinker cooler

The increase on capacity due to the oxygen enriched air combustion will increase the cooler duty. Since the existing clinker cooler duty is designed based on the existing capacity with 10% allowance assumed, some amount of heat will not be recovered in the clinker cooler or must be released to atmosphere for capacity increase of 10%. To tackle this problem, the existing clinker cooler need to be upgraded to increase its duty or by modifying the existing clinker cooler or by installing new clinker cooler as additional to the existing clinker cooler.

5.7 Modification of heat recovery scheme

One drawback of oxygen enriched air combustion implementation on existing cement plant is there will be more heat loss through clinker cooler which is linearly increase with the increase of the oxygen level. Normally there will be heat recovery at the clinker cooler by kiln secondary air which keeps the system energy in balance. But in the oxygen enriched air combustion case, the kiln secondary air decrease linearly with the increase of the oxygen level, which causes higher heat loss to atmosphere via vent air.

It is needed modification on heat recovery scheme to keep the energy balance of the system. The most plausible way left to recover the heat from the clinker cooler is to utilize the heat for raw material preheating at the preheater section by installing new piping connection from preheater tied to kiln's secondary air pipe where some amount of heated air as a balancing of kiln's secondary air decrease is routed to preheater for raw material heating.

6 Local handling and intermediate storage of oxygen

This chapter will describe about oxygen storage system required and sizing of the relevant equipment.

6.1 Oxygen storage system

Oxygen enriched air combustion involved a large mass flow rate of pure oxygen. For long term analysis, it will be more economics to build the oxygen production unit. But for earlier stage of oxygen enriched air combustion implementation, it is envisaged that the pure oxygen will be purchased from the external resource.

When the external resource of oxygen supplier is located within existing cement plant vicinity area, the pure oxygen can be supplied via pipeline system. But if it is located at distance area from the existing cement plant, delivery of liquid oxygen by truck is the most plausible option.

It assumed that the pure liquid oxygen will be supplied by truck periodically. Therefore, Vertical cryogenic vessel completed with vaporizer as indicated in fig.6.1 is needed to be provided to store the liquid oxygen.

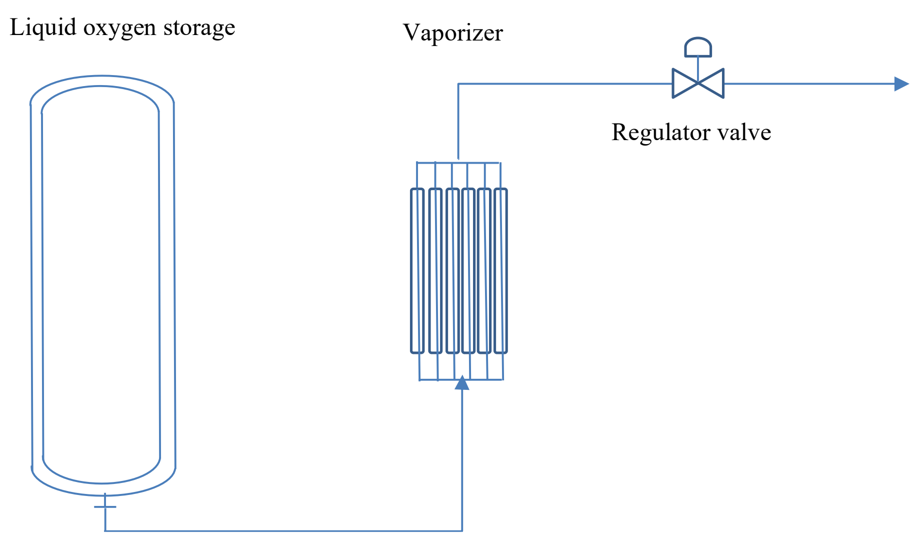

Figure 6.1 Liquid oxygen storage system schematic diagram

6.2 Oxygen storage system sizing

6.2.1 Vaporizer capacity

Capacity of vaporizer, $Cap_{vaporizer}$ (m3/h) is determined from the pure oxygen mass flow rate by

$$Cap_{vaporizer} = \frac{\dot{m}_{o2,pure}}{\rho_{O2,gas}} \tag{6.1}$$

Where,

$\rho_{O2,liq}$ = gas phase oxygen density (kg/Nm3)

$\dot{m}_{o2,pure}$ = pure oxygen flow rate in primary air (kg/h)

$\dot{m}_{air,prim,rk}$ = primary air flow rate in the rotary kiln (kg/h)

Tabel 6.1 Design basis value for vaporizer sizing

Parameter	Unit	Value
Pure oxygen flow rate	kg/h	7320
gas phase oxygen density	kg/Nm3	1.429
Allowance	%	10

Based on equation 6.1 and design basis value as table 6.1 above, it is determined that the capacity of vaporizer is at 5635 Nm3/h.

6.2.2 Oxygen storage sizing

It is envisaged that the pure oxygen will be supplied via truck, therefore oxygen storage capacity $m_{O2,stor}$ is specified as capacity of truck that is around 10 ton. No spare seems required for oxygen storage because during oxygen supply interruption, the operation of the rotary kiln can be switched back to atmospheric air combustion mode.

To determine the required volume of the vessel, V_{stor} (m3) is calculated by

$$V_{stor} = \frac{m_{O2,stor}}{\rho_{O2,liq}} \tag{6.2}$$

Where,

V_{stor} = Storage volume (m3)

$\rho_{O2,liq}$ = Liquid oxygen density (kg/m3)

Tabel 6.2 Design basis value for oxygen storage sizing

Parameter	Unit	Value
Pure oxygen mass	kg	10000
liquid phase oxygen density	kg/m3	1140
Allowance	%	10

Based on equation 6.2 and design basis value on table 6.2 above, it is determined that the capacity of vaporizer is at 9.6 m3.

Vessel is assumed as cylinder, by neglecting the volume of the head, the required volume of the storage is calculated as

$$V_{stor} = \frac{\pi D^2}{4} x L \tag{6.3}$$

Where,

$D =$ Diameter (m)

$L =$ Length (m)

Vessel dimension is determined after trial and error to have the L/D is 3. It is estimated that the dimension of the vessel is 1.6m (diameter) x 4.8 (length). However, the dimension of the vessel is not a mandatory one, readymade vessel vendor may offer a different dimension and it is acceptable.

7 Economic analysis

This chapter will describe about economic analysis of implementation of oxygen as partial replacement for air in cement kiln combustion consisted of CAPEX and relative nett present value (NPV).

Relative NPV calculation result is attached in appendix G.

7.1 Capital expenditure (CAPEX)

Implementation of oxygen enriched air combustion require CAPEX cost for oxygen liquid storage system provision as mentioned earlier. Since the oxygen liquid storage system has been a sort of a common package in industry therefore a complete ready made one is available in the market to be purchased. It seems more economics and simply to purchase a ready-made one compared to build a new one from a scratch.

Table 7.1 below is the purchase cost of oxygen storage system referred to the offers available in the internet [15], [16].

Tabel 7.1 Purchase cost of oxygen storage system equipment

equipment	Unit	Value
Liquid oxygen storage	KNOK	125
Vaporizer	KNOK	25
Regulator valve	KNOK	8.5
Transportation cost	KNOK	17
Installation	KNOK	15

7.2 Operational expenditure (OPEX)

For simplification, it is assumed that there is no significant changes in operational cost of the existing plant on oxygen enriched air combustion implementation in term of labor cost, utility cost, maintenance cost, tax cost except on the fuel cost, raw material cost and pure oxygen cost that are variable that will be taken into account in the next chapter.

7.3 Relative Net Present Value Analysis

This chapter is purposed to determine the economic price of the liquid oxygen per ton in KNOK of oxygen enriched air combustion for both coal and waste fuel at different O_2 level. The analysis is using relative net present value analysis where the base case is set as the reference point. Then, relative NPV is calculated as the amount of different of calculated NPV against the reference point.

Here below table 7.2 are calculation basis value of relative NPV analysis:

Tabel 7.2 Relative NPV analysis calculation basis

Param	Unit	Value
Gross margin/ton clinker	KNOK	0.5
Coal price	KNOK	1
Waste fuel price	KNOK	0.5
O2 cost	KNOK	Variable
Interest rate	-	7.5
Period	Year	24
DCF factor	-	11.98
USD/NOK	-	8.36
Operational hour	Hour/year	7315

Here, NPV_{rel} is NPV calculated relative to the base case by

$$NPV_{rel} = \sum_{N=zero}^{N=end} discounted\ cash\ flow\ (P_N) \tag{7.1}$$

$$NPV_{rel} = \sum_{N=zero}^{N=end} \{Nondiscounted\ cash\ flow\ (F_N)\ x\ discount\ factor\} \tag{7.2}$$

Discount factor is calculated by using

$$Discounted\ factor = \frac{1}{(1+i)^N} \tag{7.3}$$

Where,

i = discount rate per year

In this case, the discounted cash flow uses the precalculated discounted factor as stipulated in calculation basis.

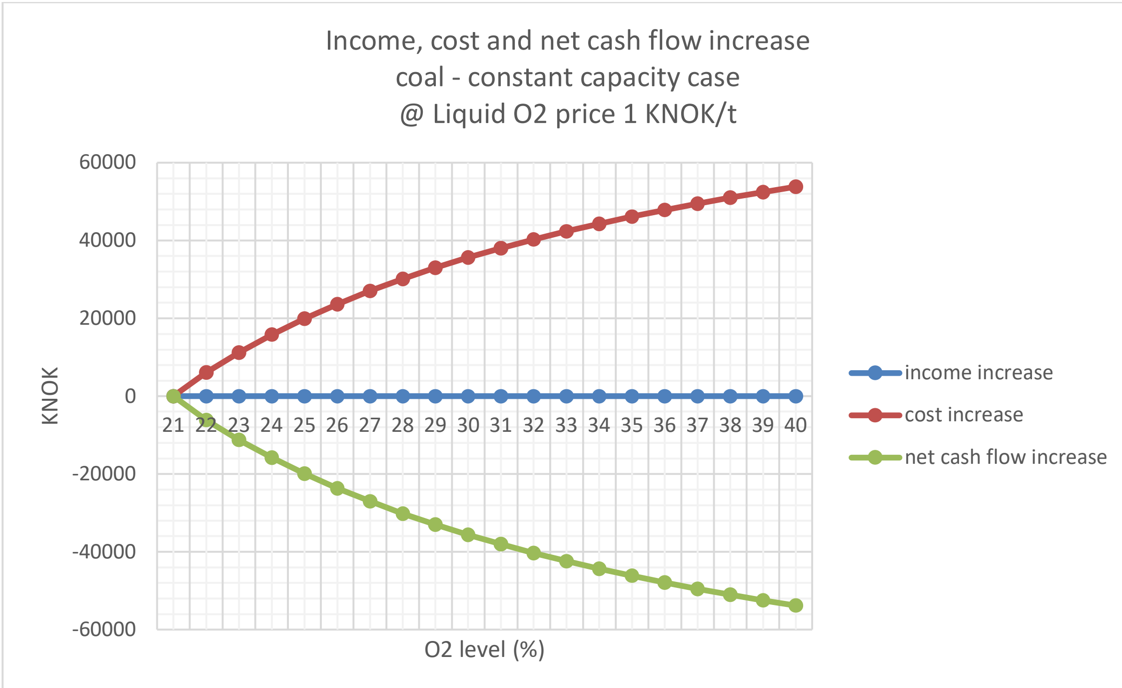

Figure 7.1 income, cost and net cash flow – coal constant capacity case

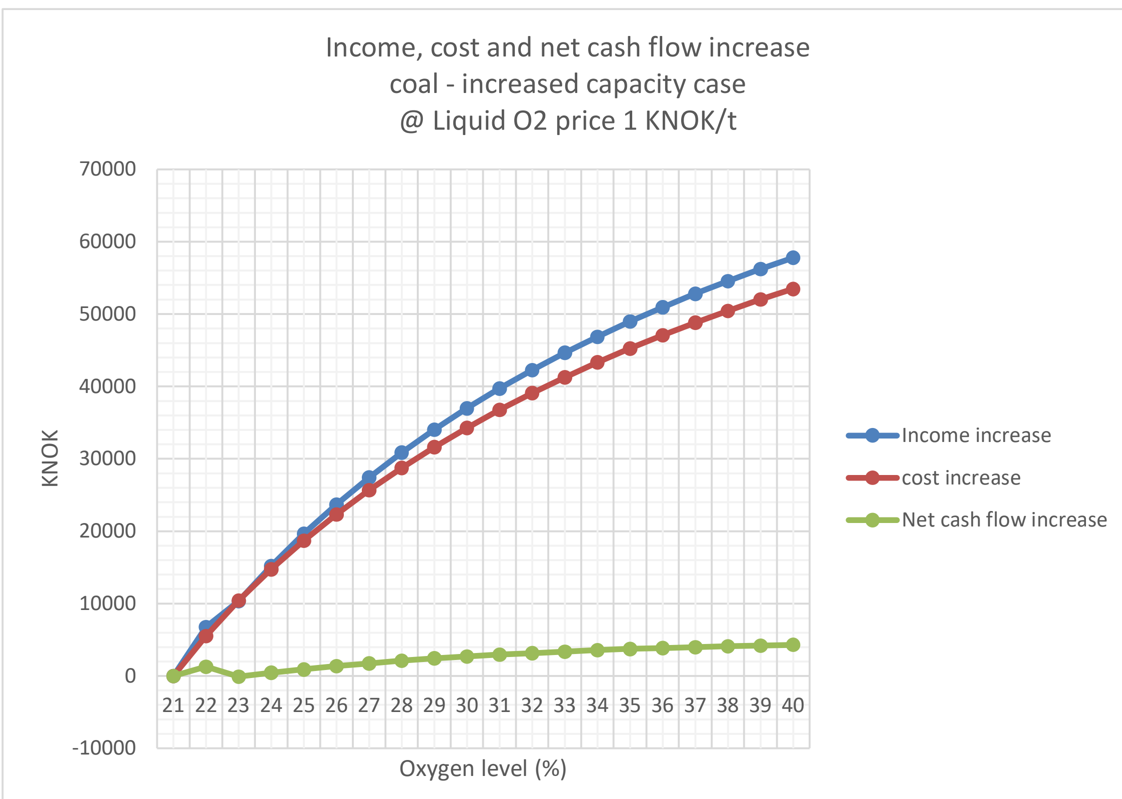

Figure 7.2 income, cost and net cash flow - coal increased capacity case

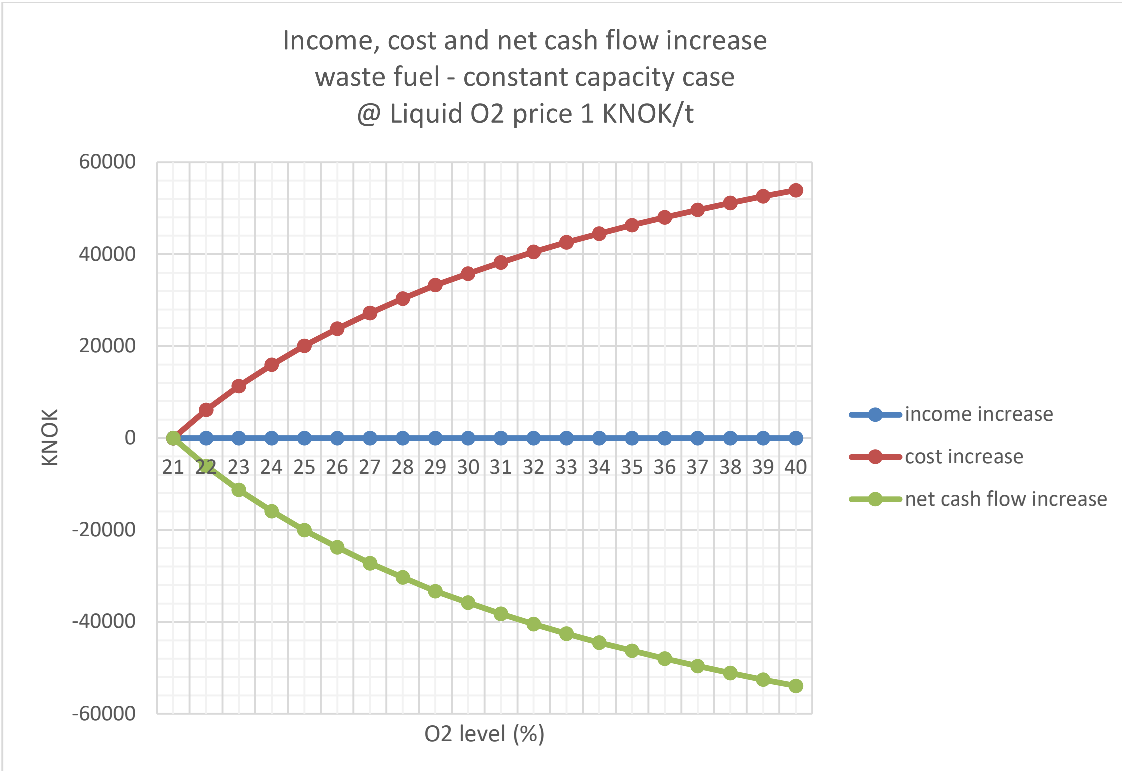

Figure 7.3 income, cost and net cash flow - waste fuel constant capacity case

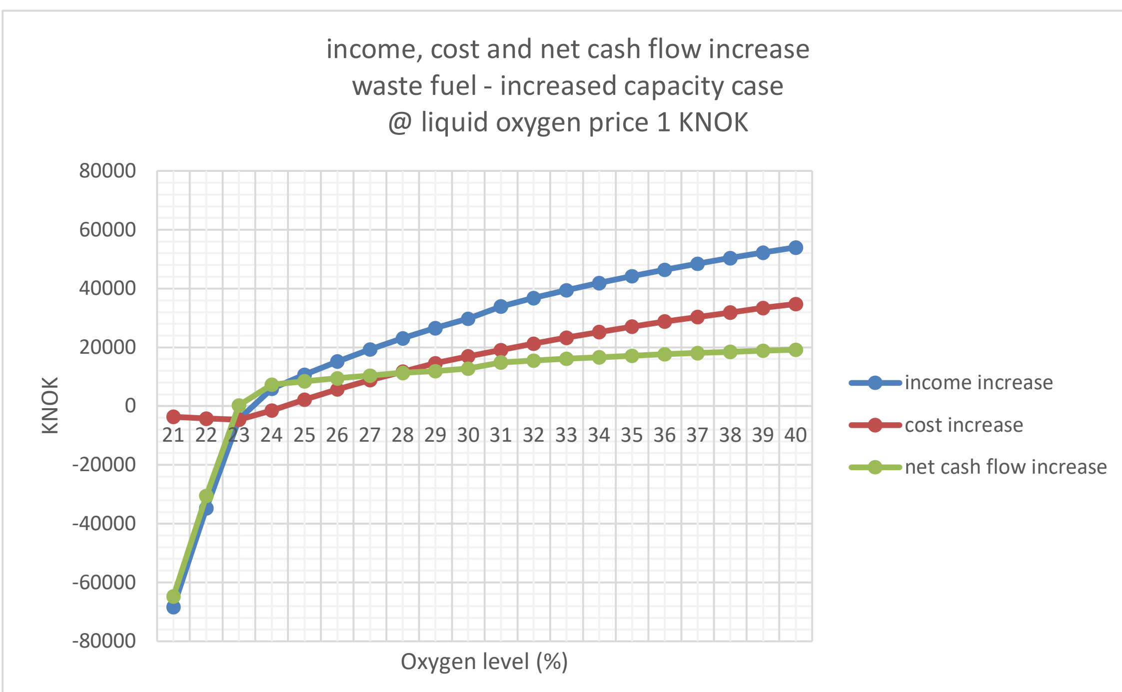

Figure 7.4 income, cost and net cash flow - waste fuel increased capacity case

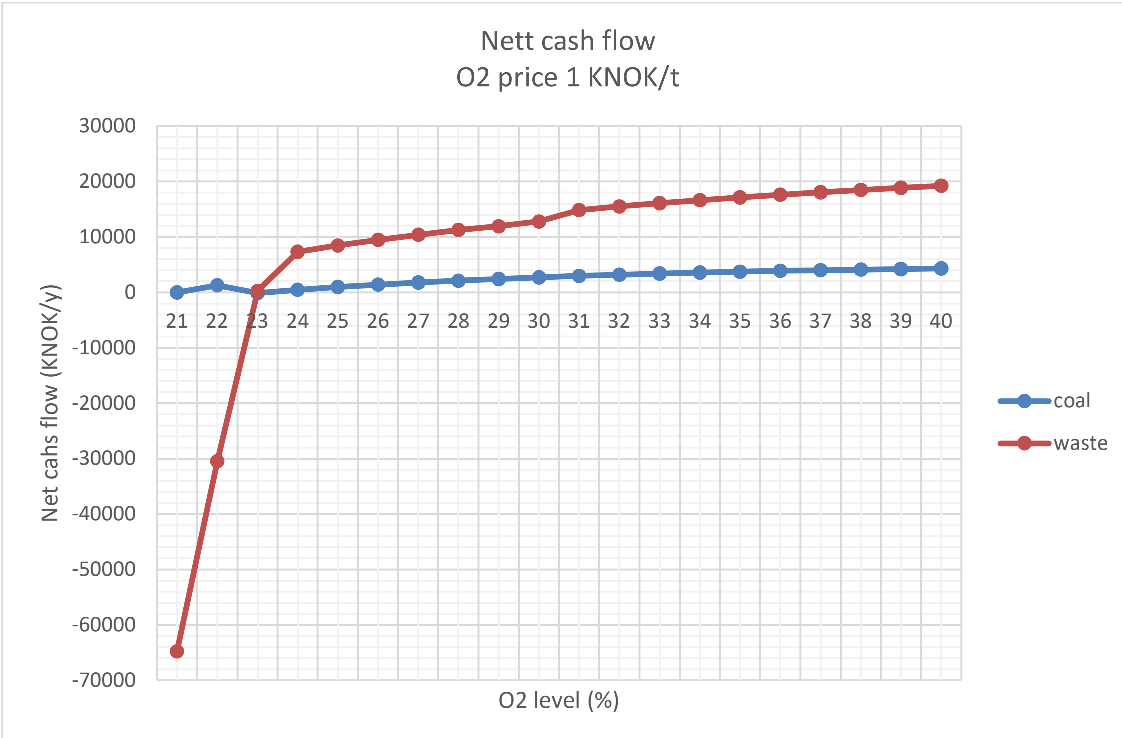

Figure 7.5 Nondiscounted nett cash flow at O2 price 1 KNOK/t

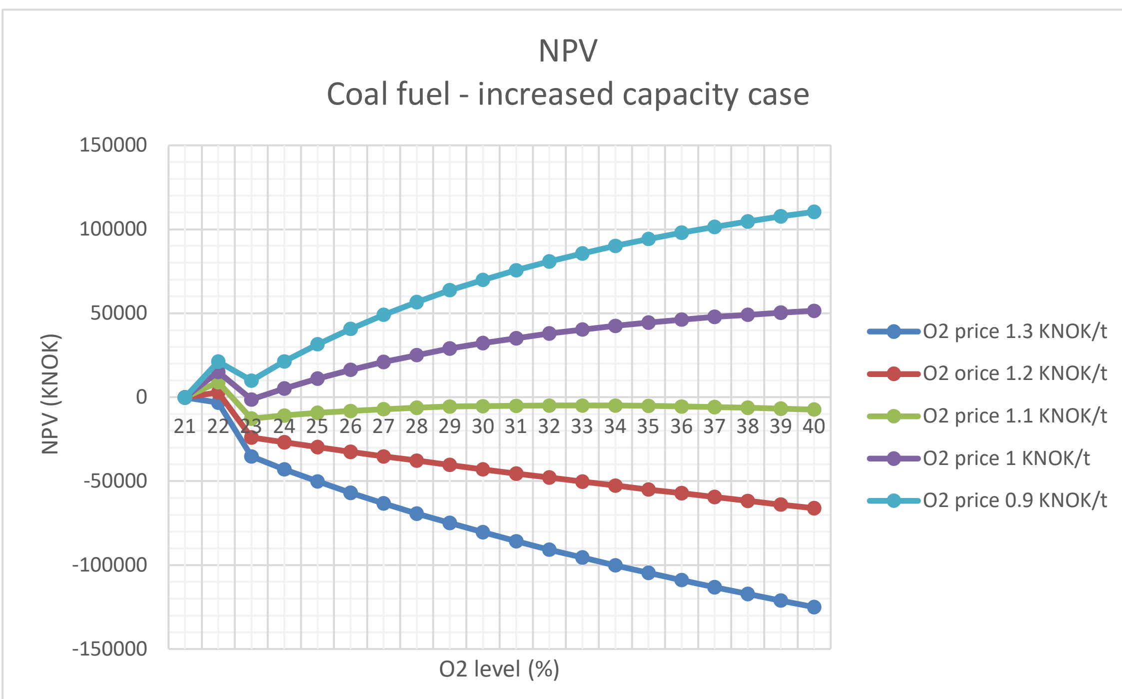

Figure 7.6 NPV analysis – coal fuel increased capacity case

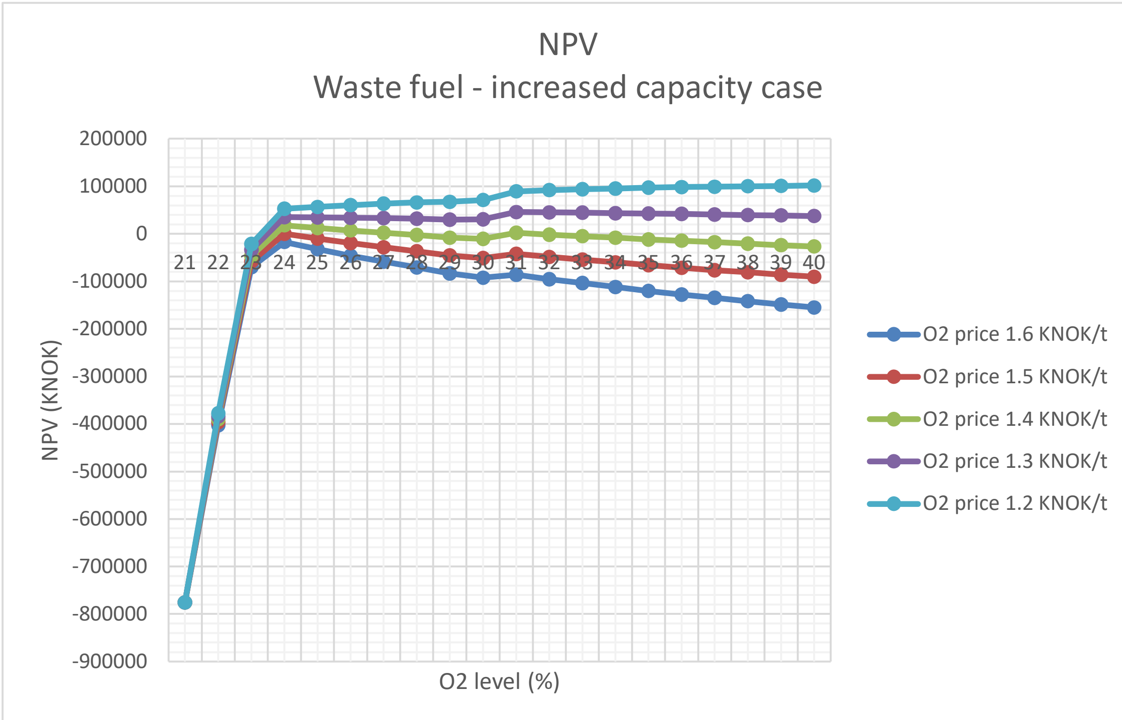

Figure 7.7 NPV analysis – waste fuel increased capacity case

Constant capacity case doesn't give a profit since the net cash flow is negative as shown by the figure 7.1 and 7.3. This is due to the nett positive fuel increase on both coal and waste fuel but no increase in capacity.

On increased capacity case as shown by figure 7.2 and 7.4, the nett positive fuel increase cost is compensated by the income from capacity increase so that result in net cash flow or margin. Figure 7.4 shows that the profit decrease on waste fuel combustion at initial due to the capacity drop. But by oxygen enriched air combustion, the profit soon increases and then cross the profit of the coal at oxygen level 23%. Above oxygen level 23%, it is found that oxygen enriched air combustion of waste fuel gives more profit than coal combustion. The benefit increases linearly by increasing of the oxygen level.

It is found that there is significant different in term of net cash flow between coal and waste fuel where the different increases linearly following the oxygen level increment and varies following the oxygen's price as illustrated by figure 7.5. This is caused by the lower price of waste fuel as stated in the calculation basis.

Figure 7.6 and 7.7 illustrated the different figures of net present value of coal and waste fuel at varied oxygen price. It is found that economic price of liquid oxygen is below 1.1 KNOK/t for coal and below 1.4 KNOK/t for waste fuel. It means that to make this oxygen enriched air combustion profitable, the price of oxygen should be below 1.1 KNOK for coal and below 1.4 KNOK for waste fuel case.

8 Conclusion

Theoretical adiabatic flame temperature is increasing by increasing of oxygen level. Although the actual temperature is below the adiabatic flame temperature, application of oxygen enriched air combustion at high oxygen level should consider a more heat resistant refractory.

At constant capacity case, fuel rate will decrease on the kiln but will increase on the calciner for both coal and waste fuel. But on increased capacity case, fuel rate on the kiln will almost be same as base case and keep increases on calciner for both coal and waste fuel.

Exist gas flow rate will decrease at both kiln and calciner on constant capacity case for both coal and waste fuel. On increased capacity case, kiln exit gas flow rate will decrease, but the calciner exit gas flow rate is kept constant as the base case, except for waste fuel increased capacity case below oxygen level 23% where the capacity increase is restricted by the kiln exit gas flow rate instead of calciner exit gas flow rate. Therefore, below 23% oxygen level, the production capacity of waste fuel increased capacity case drops to -13.6%.

In general, capacity increase of coal fuel is higher than the waste fuel case. But above 24% oxygen level, the capacity of waste fuel as well as its increase is almost the same as the coal, this will lead to economic benefit of waste fuel.

Pure oxygen flow rate will be higher on oxygen enriched air combustion of waste fuel due to the higher fuel rate of waste fuel compared to coal combustion.

According to the simulation result conducted, constant capacity case of oxygen enriched air combustion on existing cement plant doesn't give economics benefit. Although there is fuel rate decrease on the kiln, but it is also accompanied with fuel rate increase on the calciner, where nett positive fuel addition is required in total. This can be explained due to the decrease of kiln exhaust gas flowrate on oxygen enriched air combustion which brings less energy to calciner compared to atmospheric air combustion at same temperature i.e 1150. By modification of heat recovery scheme as proposed in previous chapter, the shortage of energy in calciner can be compensated that in turn will result in net negative fuel addition in total and can make the process more profitable.

By increasing the production capacity, oxygen enriched air combustion on existing cement plant give significant economic benefit, where waste fuel is more profitable compared to the coal fuel at oxygen level above 23%. The benefit is increasing linearly by increasing of oxygen level and by decreasing of pure oxygen price. It is found that the economics price of pure oxygen is below 1.1 KNOK/t for coal, meanwhile for waste fuel the economic price of liquid oxygen is below 1.4 KNOK/t.

9 Recommendation for future works

For future works on the same topic, study on kiln fuel mix composition to find the most optimum mixing composition is an interesting idea to be done. Study of implementation of oxygen enriched air combustion on calciner to study the impact of oxygen enriched air combustion on cement plant more comprehensively will be challenging. Also study of oxygen enriched air combustion with others different type of fuel seems an interesting idea to be explored.

For the model improvement, it is proposed to include into the model the modification on heat recovery scheme between clinker cooler and raw material preheating to make the oxygen enriched air combustion more feasible economically.

Despite of any advantages from using excel in the modelling, there were also several draw-back that faced during modelling with excel. Using Phyton or Matlab may be another option should be considered.

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Appendices

Appendix A – Task description

University of South-Eastern Norway

Faculty of Technology, Natural Sciences and Maritime Sciences, Campus Porsgrunn

FMH606 Master's Thesis

Title: Utilization of oxygen as partial replacement for air in cement kiln combustion processes

USN supervisors: Lars-André Tokheim

External partners: Norcem AS Brevik

Task background:

Hydrogen may play an important role as an energy carrier in future energy systems in Norway or other countries. For example, 'green' hydrogen may be formed by electrolysis, using electricity generated by renewable energy sources like wind or solar energy. The hydrogen itself may be used for different purposes (e.g. combustion processes with zero emissions, or chemical production processes), but in such cases, oxygen will also be produced, and this oxygen may be useful in other processes. If the oxygen can be used in other processes, it may make the green hydrogen production process more economically viable. One example of oxygen utilization is as a partial replacement for air in combustion of waste fuels in cement kilns (see figure 1), i.e. in the rotary kiln and/or in the calciner.

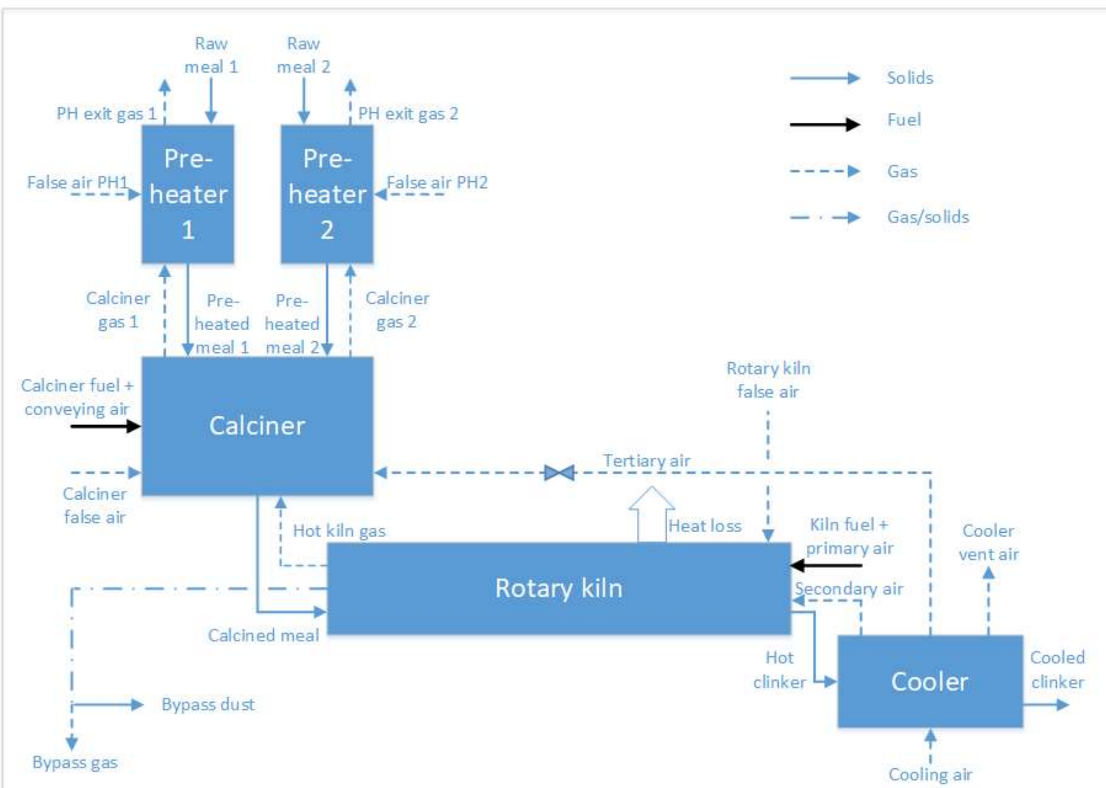


***Figure 1:** A regular cement kiln process with two preheater strings.*

Such use may be advantageous because it could make it possible to obtain higher combustion temperatures, and also because the total gas flow rate in the system may be reduced. The former (higher combustion temperatures) may make it possible to increase the waste fuel consumption and reduce the fossil fuel consumption, whereas the latter (reduced gas flow rates) may result in a higher plant production capacity. Hence, both environment and economy may benefit from this.

<u>Task description</u>:
The task may include the following:

- Describe the relevant chemical and physical processes
- Evaluate the impact of using oxygen as a partial replacement for air in waste fuel combustion; the impact on fuel feed rate and production capacity is of particularly high importance
- As part of the evaluation, make a model based on mass and energy balances for (part of) the system
- Select an appropriate simulation tool based on an assessment of different available options, and calculate relevant mass flow rates, temperatures, duties, etc. using the selected tool
- Make relevant process flow diagrams with process values for selected cases based on relevant design basis values
- Simulate different cases with the selected simulation tool, varying key parameters in the system
- Describe required constructional changes to the kiln system
- Assess local handling and intermediate storage of oxygen
- Determine the required size of relevant equipment units
- Make estimates of investment costs (CAPEX) and operational costs (OPEX) of the suggested process changes.
- Determine what oxygen purchasing prices that would be required for oxygen utilization to be economically viable in the cement plant

<u>Student category</u>: EET or PT students

<u>Is the task suitable for online students (not present at the campus)?</u> Yes, both online and campus students may select the task.

<u>Practical arrangements</u>:
There may be meetings with Norcem to discuss the task and the progress, most likely via Skype/Teams/Zoom (due to the corona situation).

<u>Supervision:</u>
As a general rule, the student is entitled to 15-20 hours of supervision. This includes necessary time for the supervisor to prepare for supervision meetings (reading material to be discussed, etc).

<u>Signatures</u>:

Supervisor (date and signature): 21-Jan-2021, Lars André Tokheim

Student (write clearly in all capitalized letters): SYAIFUL BAHRI,

Student (date and signature): 21-Jan-2021,

Appendix B - Adiabatic flame temperature calculation

ADIBATIC FLAME TEMPERATURE CALCULATION
COAL

$$C_\alpha H_\beta O_\gamma N_\delta S_\epsilon + \left(\alpha + \frac{\beta}{4} - \frac{\gamma}{2} + \epsilon\right)\ (O_2 + 3.77N_2)\ \rightarrow\ \alpha CO_2 + \frac{\beta}{2} H_2O + \epsilon SO_2 + \left(3.77\left(\alpha + \frac{\beta}{4} - \frac{\gamma}{2} + \epsilon\right) + \frac{\delta}{2}\right) N_2$$

$$T_p = T_R + \frac{LHV.N_{fuel}.M_{fuel}}{\sum_i N_{i,P}\hat{c}_{pi}}$$

N	699.5493	Kmol
MW of coal	2300.315	Kg/Kmol
LHV	28	MJ/kg
	0.012172	MJ/kmol
TR	1154	
Assume Flame Tp	2700	K
T ave	1499	K

Fuel			Flue Gas				
Substance	Coeff	MW	Substance	Coeff (Ni)	Cpi at Tav		Ni.Cpi
					(kcal/ kmol-K)	(kJ/ kmol-K)	(kJ / kmol-K)
C	161	12.011	CO2	160.6231	14.36	60.12	9657.226124
H	106	1.008	H2O	53.01753	11.46	47.96	2542.894183
O	10	15.999	SO2	3.052283	13.78	57.69	176.0947827
N	1	14.007	N2	556.7726	8.00	33.49	18646.43409
S	3	32.065					
SUM							31022.64918

Calculated Tp	3230.187	K
Average	1764.093	K

Fuel			Flue Gas				
Substance	Coeff	MW	Substance	Coeff (Ni)	Cpi at Tav		Ni.Cpi
					(kcal/ kmol-K)	(kJ/ kmol-K)	(kJ / kmol-K)
C	161	12.011	CO2	160.6231	15.11	63.27	10161.96178
H	106	1.008	H2O	53.01753	12.65	52.98	2809.017807
O	10	15.999	SO2	3.052283	14.47	60.57	184.8744757
N	1	14.007	N2	556.7726	8.26	34.60	19264.39206
S	3	32.065					
SUM							32420.24613

Calculated Tp	3140.685	K
Average	1719.342	K

Fuel			Flue Gas				
Substance	Coeff	MW	Substance	Coeff (Ni)	Cpi at Tav		Ni.Cpi
					(kcal/ kmol-K)	(kJ/ kmol-K)	(kJ / kmol-K)
C	161	12.011	CO2	160.6231	14.98	62.74	10077.27406

H	106	1.008	H2O	53.01753	12.44	52.08	2761.159994
O	10	15.999	SO2	3.052283	14.36	60.12	183.4969474
N	1	14.007	N2	556.7726	8.22	34.41	19160.07336
S	3	32.065					
SUM							32182.00437

Calculated Tp 3155.392 K
Average 1726.696 K

Fuel			Flue Gas				
Substance	Coeff	MW	Substance	Coeff (Ni)	Cpi at Tav		Ni.Cpi
					(kcal/ kmol-K)	(kJ/ kmol-K)	(kJ / kmol-K)
C	161	12.011	CO2	160.6231	15.01	62.83	10091.20224
H	106	1.008	H2O	53.01753	12.47	52.23	2768.942395
O	10	15.999	SO2	3.052283	14.38	60.19	183.7262257
N	1	14.007	N2	556.7726	8.23	34.44	19177.21546
S	3	32.065					
SUM							32221.08632

Calculated Tp 3152.965 K 2879.815
Average 1725.482 K

ADIBATIC FLAME TEMPERATURE CALCULATION

WASTE FUEL

$$C_\alpha H_\beta O_\gamma N_\delta S_\epsilon + \left(\alpha + \frac{\beta}{4} - \frac{\gamma}{2} + \epsilon\right)(O_2 + 3.77N_2) \rightarrow \alpha CO_2 + \frac{\beta}{2}H_2O + \epsilon SO_2 + \left(3.77\left(\alpha + \frac{\beta}{4} - \frac{\gamma}{2} + \epsilon\right) + \frac{\delta}{2}\right)N_2$$

$$T_p = T_R + \frac{LHV.N_{fuel}.M_{fuel}}{\sum_i N_{i,P}\hat{c}_{pi}}$$

N	662.7839	Kmol
MW of coal	2392.154	Kg/Kmol
LHV	23.1	MJ/kg
	0.009657	MJ/kmol
TR	1154	
Assume Flame Tp	2700	K
T ave	1499	K

Fuel			Flue Gas				
Substance	Coeff	MW	Substance	Coeff (Ni)	Cpi at Tav		Ni.Cpi
					(kcal/ kmol-K)	(kJ/ kmol-K)	(kJ / kmol-K)
C	146	12.011	CO2	145.8592	14.36	60.12	8769.568724
H	156	1.008	H2O	78.08036	11.46	47.96	3744.989615
O	22	15.999	SO2	1	13.78	57.69	57.69280262
N	7	14.007	N2	528.1291	8.00	33.49	17687.15531
S	1	32.065					
SUM							30259.40645

Calculated Tp	2980.168	K
Average	1639.084	K

Fuel			Flue Gas				
Substance	Coeff	MW	Substance	Coeff (Ni)	Cpi at Tav		Ni.Cpi
					(kcal/ kmol-K)	(kJ/ kmol-K)	(kJ / kmol-K)
C	146	12.011	CO2	145.8592	14.76	61.79	9012.661157
H	156	1.008	H2O	78.08036	12.07	50.52	3944.425737
O	22	15.999	SO2	1	14.16	59.27	59.27365854
N	7	14.007	N2	528.1291	8.14	34.08	17996.90488
S	1	32.065					
SUM							31013.26543

Calculated Tp	2935.778	K
Average	1616.889	K

Fuel			Flue Gas				
Substance	Coeff	MW	Substance	Coeff (Ni)	Cpi at Tav		Ni.Cpi
					(kcal/ kmol-K)	(kJ/ kmol-K)	(kJ / kmol-K)
C	146	12.011	CO2	145.8592	14.70	61.53	8974.294648

H	156	1.008	H2O	78.08036	11.97	50.10	3911.680884
O	22	15.999	SO2	1	14.10	59.03	59.03228026
N	7	14.007	N2	528.1291	8.12	33.98	17947.82815
S	1	32.065					
SUM							30892.83596

Calculated Tp 2942.724 K
Average 1620.362 K

Fuel			Flue Gas				
Substance	Coeff	MW	Substance	Coeff (Ni)	Cpi at Tav		Ni.Cpi
					(kcal/ kmol-K)	(kJ/ kmol-K)	(kJ / kmol-K)
C	146	12.011	CO2	145.8592	14.71	61.57	8980.301388
H	156	1.008	H2O	78.08036	11.98	50.16	3916.776156
O	22	15.999	SO2	1	14.11	59.07	59.07027591
N	7	14.007	N2	528.1291	8.12	34.00	17955.50743
S	1	32.065					
SUM							30911.65526

Calculated Tp 2941.635 K 2668.485
Average 1619.818 K

Appendix C – Mass and Energy balance model

ROTARY KILN CALCULATION
COAL CASE

Calculation Basis;

Table 3.2. Ultimate analysis (C, H, O, S, N), moisture content, ash content, lower heating value (LHV) and fossil fraction and split between rotary kiln and calciner for different fuels.

Parameter	Unit	Coal	RDF	SHW	AM	LHW	Fuel mix rotary kiln	Fuel mix calciner
Mass fraction of C	kg/kg	0.722	0.348	0.359	0.463	0.437	0.601	0.395
Mass fraction of H	kg/kg	0.040	0.050	0.053	0.065	0.080	0.054	0.049
Mass fraction of O	kg/kg	0.057	0.245	0.285	0.149	0.253	0.121	0.232
Mass fraction of S	kg/kg	0.012	0.003	0.012	0.004	0.016	0.011	0.006
Mass fraction of N	kg/kg	0.016	0.006	0.006	0.097	0.018	0.034	0.007
Mass fraction of moisture	kg/kg	0.018	0.250	0.118	0.029	0.198	0.060	0.192
Mass fraction of ash	kg/kg	0.135	0.098	0.167	0.192	0.000	0.118	0.119
Lower heating value	MJ/kg	28.0	14.2	15.9	19.4	14.6	23.1	16.2
Fossil fraction	-	100 %	30 %	70 %	0 %	100 %	78 %	48 %
Mass fraction used in the rotary kiln	-	56 %	0 %	0 %	22 %	22 %	100 %	0 %
Mass fraction used in the calciner	-	12 %	65 %	24 %	0 %	0 %	0 %	100 %
Energy fraction used in the rotary kiln	-	67 %	0 %	0 %	19 %	14 %	100 %	0 %
Energy fraction used in the calciner	-	20 %	57 %	23 %	0 %	0 %	0 %	100 %

LHV	28	Mj/kg
Raw material	207000	kg /h
Precalcined meal	143000	kg /h
CaCO3 weight fraction of raw material	77	%
clinker	137000	kg /h
Calcination in rotary kiln	6	%
Specific calcination enthalpy	3.4	MJ/kgclinker
Thermal energy fraction in rotary kiln	38	%

Calculation;

1. CO2 calculation

$CaCO_3(s) \rightarrow CaO(s) + CO_2(g)$

MW CaCO3	100	kg/kmol
Calcination in rotary kiln	9563.4	kg /h
	95.634	kmol/hr
CO2	95.634	kmol/hr
	4208.757	kg/hr
CO2/Clinker	0.031	kg CO2/kg clinker
	0.082	m3 CO2/kg clinker

2. Fuel calculation

Total heat required	465800.00	MJ/hr
Coal flow rate in rotary kiln	6321.6	kg/hr
fuel/precalcined meal	0.044	kg fuel /kg precalcined meal
precalcined meal/fuel	22.62	kg precalcined meal/kg fuel

3. Dust calculation

Dust out from rotary kiln	2644.7	kg/hr
Ash from fuel rotary kiln	853.4	kg/hr
Dust from calciner	1791.2	kg/hr

4. Air calculation

Oxygen req	0.07	Kmol/kg fuel
Oxygen required from air	2.20	kg/kg fuel
O2 level in air	20.95	%
N2 level in air	79.05	%
N2/O2 ratio	3.77	
Nitrogen in air	0.26	Kmol/kg fuel
	7.25	kg/kg fuel

Coal

	kg/kg	MW	kmol/kg
C	0.722	12.011	0.060111564
H	0.04	1.008	0.03968254
O	0.057	15.999	0.003562723
S	0.012	32.065	0.00037424
N	0.016	14.007	0.001142286
Moisture	0.018	18.015	0.000999167
Ash	0.135	60	0.00225

$$C_\alpha H_\beta O_\gamma N_\delta S_\epsilon + \left(\alpha + \frac{\beta}{4} - \frac{\gamma}{2} + \epsilon\right)(O_2 + 3.77N_2) \rightarrow \alpha CO_2 + \frac{\beta}{2}H_2O + \epsilon SO_2 + \left(3.77\left(\alpha + \frac{\beta}{4} - \frac{\gamma}{2} + \epsilon\right) + \frac{\delta}{2}\right)N_2$$

O2 stoichiometric	0.068625078	kmol/kg fuel
	2.19586524	kg/kg fuel

Total air	9.45	kg/kg fuel
Excess air	17.50	%
	70175.97	kg/hr
Primary air	5614.08	kg/hr
Secondary air	60451.89	kg/hr
False air	4110	kg/hr

5. Flue gas calculation

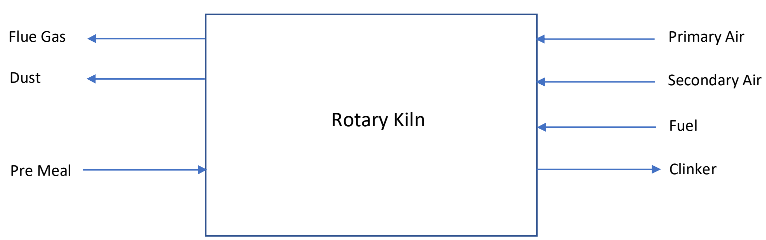
Flue gas 1150.00

	MW	kg/kg fuel	Excess air (kg/kg fuel)	N2 in fuel (kg/kg fuel)	Moisture in fuel (kg/kg fuel)	Total Kg/kg fuel	Cp @ 1000K kJ/kg.K	Cpi x wi	mas flow kg/hr			w (kg/kg)	density	Volume flow
CO2	44.009	2.65				2.65	1.23	0.272	16723.40011	4208.8	20932.2	0.262134	0.37688	55541.37
H2O	18.015	0.36			0.018	0.38	2.28	0.072			2373.373787	0.029722	0.15427	15384.24
SO2	64.063	0.02				0.02	0.85	0.002			151.5592068	0.001898	0.54861	276.2607
N2	28.014	7.25	1.27	0.02		8.54	1.16	0.828			53966.5667	0.675825	0.23990	224954
O2	31.998		0.38			0.38	1.09	0.035			2429.230818	0.030421	0.27402	8865.231

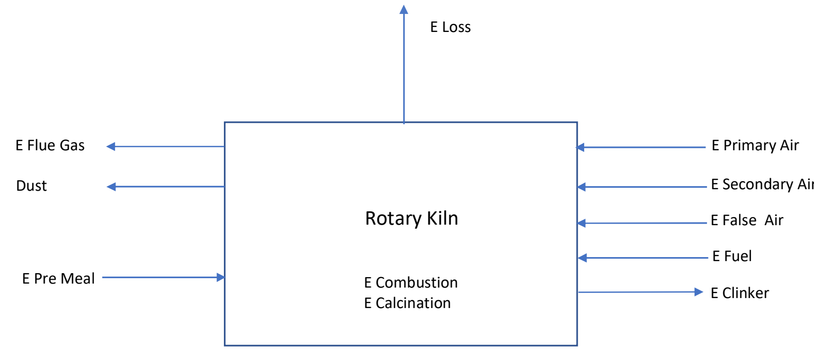
SUM	11.966	kg/kg fuel
Cp mixture	1.208	kJ/kg.K
Density of mixture	0.262	kg/m3
Flue gas from combustion	75644.1	kg/hr
CO2 from calcination	4208.8	kg/hr
Flue gas Total	79852.9	kg/hr
	305021.11	m3/hr
	84.73	m3/s
O2 in flue gas	2429.2	kg/hr
	3.04	%

ROTARY KILN MASS AND ENERGY BALANCE
COAL CASE

Mass Balance



Energy Balance



m in		m out	
kg/hr		kg/hr	
Precalcined meal	143000	Flue gas	79852.89
Fuel	6321.57	Clinker	137000
Primary air	5614.08	Dust	2644.7
Secondary air	60451.89		
False air	4110		
219498		219498	

0.00

	m	Cp	T	LHV / H	Ein	Egen	Eout
	kg/hr	kJ/kg K	K	KJ/kg	kW	kW	kW
Precalcined meal	143000	0.795048	1173		27633.42		
Fuel	6321.57	1.26	303		11.06		
Primary air	5614.08	1.01	303		7.84		
Secondary air	60451.89	1.16	1154		16673.98		
False air	4110.00	1.01	303		5.74		
Combustion	6321.57			28000		49167.78	
Post calcination	4208.76			-3600		-4208.76	
Flue gas	79852.89	1.21	1423				30138.21
Dust	2644.7	1	1423				826.45
Loss							6000.00
Cooled clinker	137000	1	1673				52326.39
					44332.04	44959.02	
			Total	MW	89.29		89.29

Flue gas T 1150 C 0.00

CALCINER CALCULATION
COAL CASE

Calculation Basis;

Table 3.2. Ultimate analysis (C, H, O, S, N), moisture content, ash content, lower heating value (LHV) and fossil fraction and split between rotary kiln and calciner for different fuels.

Parameter	Unit	Coal	RDF	SHW	AM	LHW	Fuel mix rotary kiln	Fuel mix calciner
Mass fraction of C	kg/kg	0.722	0.348	0.359	0.463	0.437	0.601	0.395
Mass fraction of H	kg/kg	0.040	0.050	0.053	0.065	0.080	0.054	0.049
Mass fraction of O	kg/kg	0.057	0.245	0.285	0.149	0.253	0.121	0.232
Mass fraction of S	kg/kg	0.012	0.003	0.012	0.004	0.016	0.011	0.006
Mass fraction of N	kg/kg	0.016	0.006	0.006	0.097	0.018	0.034	0.007
Mass fraction of moisture	kg/kg	0.018	0.250	0.118	0.029	0.198	0.060	0.192
Mass fraction of ash	kg/kg	0.135	0.098	0.167	0.192	0.000	0.118	0.119
Lower heating value	MJ/kg	28.0	14.2	15.9	19.4	14.6	23.1	16.2
Fossil fraction	-	100 %	30 %	70 %	0 %	100 %	78 %	48 %
Mass fraction used in the rotary kiln	-	56 %	0 %	0 %	22 %	22 %	100 %	0 %
Mass fraction used in the calciner	-	12 %	65 %	24 %	0 %	0 %	0 %	100 %
Energy fraction used in the rotary kiln	-	67 %	0 %	0 %	19 %	14 %	100 %	0 %
Energy fraction used in the calciner	-	20 %	57 %	23 %	0 %	0 %	0 %	100 %

LHV	28	Mj/kg
Raw material	207000	kg /h
Precalcined meal	143000	kg /h
CaCO3 weight fraction of raw material	77	%
clinker	137000	kg /h
Calcination in calciner	94	%
Specific calcination enthalpy	3.4	MJ/kgclinker
Thermal energy fraction in calciner	62	%

Calculation;

1. CO2 calculation

$CaCO_3(s) \rightarrow CaO(s) + CO_2(g)$

MW CaCO3	100	kg/kmol
Calcination in calciner	149826.6	kg /h
	1498.266	kmol/hr
CO2	1498.266	kmol/hr
	65937.188	kg/hr
CO2/Clinker	0.481	kg CO2/kg clinker
	1.053	m3 CO2/kg clinker

2. Fuel calculation

Total heat required	465800.00	MJ/hr
Coal flow rate in calciner	10314.1	kg/hr
fuel/precalcined meal	0.072	kg fuel /kg precalcined meal
precalcined meal/fuel	13.86	kg precalcined meal/kg fuel

3. Dust calculation

Dust from kiln	544.8	kg/hr
Ash from fuel calciner	1392.4	kg/hr

4. Air calculation

Oxygen req	0.07	Kmol/kg fuel
Oxygen required from air	2.20	kg/kg fuel
O2 level in air	20.95	%
N2 level in air	79.05	%
N2/O2 ratio	3.77	

Coal

	kg/kg	MW	kmol/kg
C	0.722	12.011	0.06011156
H	0.04	1.008	0.03968254
O	0.057	15.999	0.00356272
S	0.012	32.065	0.00037424
N	0.016	14.007	0.00114229
Moisture	0.018	18.015	0.00099917
Ash	0.135	60	0.00225

$$C_\alpha H_\beta O_\gamma N_\delta S_\epsilon + \left(\alpha + \frac{\beta}{4} - \frac{\gamma}{2} + \epsilon\right)(O_2 + 3.77N_2) \rightarrow \alpha CO_2 + \frac{\beta}{2}H_2O + \epsilon SO_2 + \left(3.77\left(\alpha + \frac{\beta}{4} - \frac{\gamma}{2} + \epsilon\right) + \frac{\delta}{2}\right)N_2$$

O2 stoichiometric	0.068625078	kmol/kg fuel
	2.19586524	kg/kg fuel

Nitrogen in air	0.26	Kmol/kg fuel
	7.25	kg/kg fuel
Total air	9.45	kg/kg fuel
Excess air	26.20	%
	122979.19	kg/hr
Primary air	9838.34	kg/hr
Secondary air	109030.85	kg/hr
False air	4110	kg/hr

5. Flue gas calculation

Flue gas

	MW	kg/kg fuel	Excess air (kg/kg fuel)	N2 in fuel (kg/kg fuel)	Moisture in fuel (kg/kg fuel)	Total Kg/kg fuel	Cp @ 1000K kJ/kg.K	Cpi x wi	mas flow kg/hr			w (kg/kg)	density	Volume flow
CO2	44.009	2.65				2.65	1.23	0.254	27285.54755	65937.2	93222.7	0.471207	0.45719	203904.7
H2O	18.015	0.36			0.018	0.38	2.28	0.067			3872.346706	0.019573	0.18715	20691.26
SO2	64.063	0.02				0.02	0.85	0.002			247.280811	0.00125	0.66552	371.561
N2	28.014	7.25	1.90	0.02		9.17	1.16	0.832			94560.95216	0.477971	0.29102	324925.4
O2	31.998		0.58			0.58	1.09	0.049			5934.794049	0.029998	0.33241	17853.76

SUM	12.788	kg/kg fuel
Cp mixture	1.204	kJ/kg.K
Density of mixture	0.348	kg/m3
Flue gas from combustion	131900.9	kg/hr
CO2 from calcination	65937.2	kg/hr
Flue gas Total (calciner)	197838.1	kg/hr
	567746.71	m3/hr
	157.71	m3/s
O2 in flue gas	5934.8	kg/hr
	3.00	%
Spesific energy from kiln gas to calciner	0.21	MW/ton precalcined meal

Total Flue gas out from calciner

Total flue gas rate = Calciner flue gas + Rotary kiln flue gas

242.44 m3/s

ROTARY KILN CALCULATION
WASTE FUEL CASE

Calculation Basis;

Table 3.2. Ultimate analysis (C, H, O, S, N), moisture content, ash content, lower heating value (LHV) and fossil fraction and split between rotary kiln and calciner for different fuels.

Parameter	Unit	Coal	RDF	SHW	AM	LHW	Fuel mix rotary kiln	Fuel mix calciner
Mass fraction of C	kg/kg	0.722	0.348	0.359	0.463	0.437	0.601	0.395
Mass fraction of H	kg/kg	0.040	0.050	0.053	0.065	0.080	0.054	0.049
Mass fraction of O	kg/kg	0.057	0.245	0.285	0.149	0.253	0.121	0.232
Mass fraction of S	kg/kg	0.012	0.003	0.012	0.004	0.016	0.011	0.006
Mass fraction of N	kg/kg	0.016	0.006	0.006	0.097	0.018	0.034	0.007
Mass fraction of moisture	kg/kg	0.018	0.250	0.118	0.029	0.198	0.060	0.192
Mass fraction of ash	kg/kg	0.135	0.098	0.167	0.192	0.000	0.118	0.119
Lower heating value	MJ/kg	28.0	14.2	15.9	19.4	14.6	23.1	16.2
Fossil fraction	-	100 %	30 %	70 %	0 %	100 %	78 %	48 %
Mass fraction used in the rotary kiln	-	56 %	0 %	0 %	22 %	22 %	100 %	0 %
Mass fraction used in the calciner	-	12 %	65 %	24 %	0 %	0 %	0 %	100 %
Energy fraction used in the rotary kiln	-	67 %	0 %	0 %	19 %	14 %	100 %	0 %
Energy fraction used in the calciner	-	20 %	57 %	23 %	0 %	0 %	0 %	100 %

LHV	23.1	Mj/kg
Raw material	178746.7679	kg /h
Precalcined meal	123482.07	kg /h
CaCO3 weight fraction of raw material	77	%
clinker	118301.00	kg /h
Calcination in rotary kiln	6	%
Specific calcination enthalpy	3.634082123	MJ/kgclinker
Thermal energy fraction in rotary kiln	38	%

Calculation;

1. CO2 calculation

$CaCO_3(s) \rightarrow CaO(s) + CO_2(g)$

MW CaCO3	100	kg/kmol
Calcination in rotary kiln	8258.1	kg /h
	82.58100678	kmol/hr
CO2	82.581	kmol/hr
	3634.308	kg/hr
CO2/Clinker	0.031	kg CO2/kg clinker
	0.082	m3 CO2/kg clinker

2. Fuel calculation

Total heat required	429915.55	MJ/hr
Coal flow rate in rotary kiln	7072.2	kg/hr
fuel/precalcined meal	0.057	kg fuel /kg precalcined meal
precalcined meal/fuel	17.46	kg precalcined meal/kg fuel

3. Dust calculation

Dust out from rotary kiln	2388.4	kg/hr
Ash from fuel rotary kiln	841.6	kg/hr
Dust from calciner	1546.8	kg/hr

4. Air calculation

Oxygen req	0.06	Kmol/kg fuel
Oxygen required from air	1.92	kg/kg fuel
O2 level in air	20.95	%
N2 level in air	79.05	%
N2/O2 ratio	3.77	

Waste

	kg/kg	MW	kmol/kg
C	0.601	12.011	0.050037466
H	0.054	1.008	0.053571429
O	0.121	15.999	0.007562973
S	0.011	32.065	0.000343053
N	0.034	14.007	0.002427358
Moisture	0.06	18.015	0.003330558
Ash	0.119	60	0.001983333

$$C_\alpha H_\beta O_\gamma N_\delta S_\epsilon + \left(\alpha + \frac{\beta}{4} - \frac{\gamma}{2} + \epsilon\right)(O_2 + 3.77N_2) \rightarrow \alpha CO_2 + \frac{\beta}{2}H_2O + \epsilon SO_2 + \left(3.77\left(\alpha + \frac{\beta}{4} - \frac{\gamma}{2} + \epsilon\right) + \frac{\delta}{2}\right)N_2$$

O2 stoichiometric	0.05999189	kmol/kg fuel
	1.919620484	kg/kg fuel

Nitrogen in air 0.23 Kmol/kg fuel
6.34 kg/kg fuel

Total air 8.26 kg/kg fuel
Excess air 17.35 %
68541.87 kg/hr

Primary air 5483.35 kg/hr
Secondary air 59509.49 kg/hr
False air 3549.03003 kg/hr

5. Flue gas calculation

Flue gas

	MW	kg/kg fuel	Excess air (kg/kg fuel)	N2 in fuel (kg/kg fuel)	Moisture in fuel (kg/kg fuel)	Total Kg/kg fuel	Cp @ 1000K kJ/kg.K	Cpi x wi	mas flow kg/hr			w (kg/kg)	density	Volume flow
CO2	44.009	2.20				2.20	1.23	0.256	15573.69191	3634.3	19208.0	0.244979	0.37688	50966.49
H2O	18.015	0.48			0.06	0.54	2.28	0.117			3836.986339	0.048937	0.15427	24871.39
SO2	64.063	0.02				0.02	0.85	0.002			155.4259343	0.001982	0.54861	283.309
N2	28.014	6.34	1.10	0.03		7.47	1.16	0.820			52851.58247	0.674069	0.23990	220306.3
O2	31.998		0.33			0.33	1.09	0.034			2354.79974	0.030033	0.27402	8593.602

SUM 10.573 kg/kg fuel
Cp mixture 1.229 kJ/kg.K
Density of mixture 0.257 kg/m3

Flue gas from combustion 74772.5 kg/hr
CO2 from calcination 3634.3 kg/hr
Flue gas Total 78406.8 kg/hr
305021.11 m3/hr
84.73 m3/s

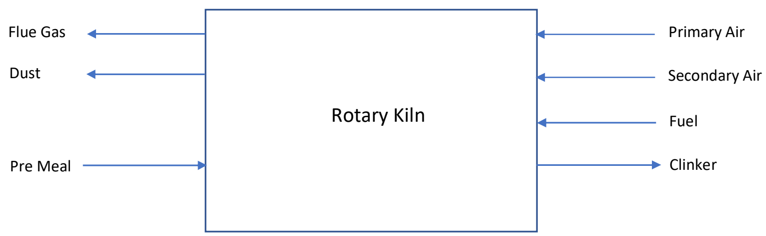
O2 in flue gas 2354.8 kg/hr
3.00 %

Total flue gas rate = Calciner flue gas + Rotary kiln flue gas
216.01 m3/s

Pure oxygen required 0.00 kg/hr
Primary air - pure oxygen 5483.35 kg/hr
Oxygen concentration in primary ai 20.95 %

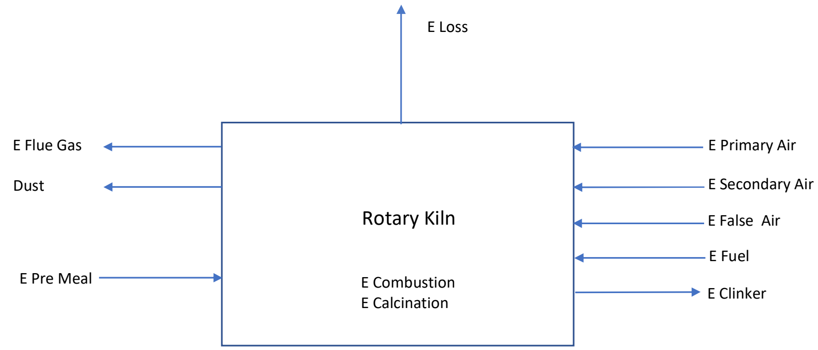
ROTARY KILN MASS AND ENERGY BALANCE
WASTE FUEL CASE

Mass Balance



m in		m out	
kg/hr		kg/hr	
Precalcined meal	123482.07	Flue gas	78406.79
Fuel	7072.20	Clinker	118301.001
Primary air	5483.35	Dust	2388.4
Secondary air	59509.49		
False air	3549.03		
199096		199096	

Energy Balance



	m	Cp	T	LHV / H	Ein	Egen	Eout
	kg/hr	kJ/kg K	K	KJ/kg	kW	kW	kW
Precalcined meal	123482.07	0.795048	1173		23861.76		
Fuel	7072.20	1.26	303		12.38		
Primary air	5483.35	1.01	303		7.65		
Secondary air	59509.49	1.16	1154		16414.04		
False air	3549.03	1.01	303		4.95		
Combustion	7072.20			23100		45379.98	
Post calcination	3634.31			-3600		-3634.31	
Flue gas	78406.79	1.23	1423				30118.17
Dust	2388.4	1	1423				746.36
Loss							6000.00
Cooled clinker	118301	1	1673				45184.41
					40300.79	41745.67	
				Total (MW)	82.05		82.05

CALCINER CALCULATION
WASTE FUEL CASE

Calculation Basis;

Table 3.2. Ultimate analysis (C, H, O, S, N), moisture content, ash content, lower heating value (LHV) and fossil fraction and split between rotary kiln and calciner for different fuels.

Parameter	Unit	Coal	RDF	SHW	AM	LHW	Fuel mix rotary kiln	Fuel mix calciner
Mass fraction of C	kg/kg	0.722	0.348	0.359	0.463	0.437	0.601	0.395
Mass fraction of H	kg/kg	0.040	0.050	0.053	0.065	0.080	0.054	0.049
Mass fraction of O	kg/kg	0.057	0.245	0.285	0.149	0.253	0.121	0.232
Mass fraction of S	kg/kg	0.012	0.003	0.012	0.004	0.016	0.011	0.006
Mass fraction of N	kg/kg	0.016	0.006	0.006	0.097	0.018	0.034	0.007
Mass fraction of moisture	kg/kg	0.018	0.250	0.118	0.029	0.198	0.060	0.192
Mass fraction of ash	kg/kg	0.135	0.098	0.167	0.192	0.000	0.118	0.119
Lower heating value	MJ/kg	28.0	14.2	15.9	19.4	14.6	23.1	16.2
Fossil fraction	-	100 %	30 %	70 %	0 %	100 %	78 %	48 %
Mass fraction used in the rotary kiln	-	56 %	0 %	0 %	22 %	22 %	100 %	0 %
Mass fraction used in the calciner	-	12 %	65 %	24 %	0 %	0 %	0 %	100 %
Energy fraction used in the rotary kiln	-	67 %	0 %	0 %	19 %	14 %	100 %	0 %
Energy fraction used in the calciner	-	20 %	57 %	23 %	0 %	0 %	0 %	100 %

LHV	28	Mj/kg
Raw material	178746.7679	kg /h
Precalcined meal	123482.0667	kg /h
CaCO3 weight fraction of raw material	77	%
clinker	118301.001	kg /h
Calcination in calciner	94	%
Specific calcination enthalpy	3.4	MJ/kgclinker
Thermal energy fraction in calciner	62	%

Calculation;

1. CO2 calculation

$CaCO_3(s) \rightarrow CaO(s) + CO_2(g)$

MW CaCO3	100	kg/kmol
Calcination in calciner	129376.9	kg /h
	1293.769106	kmol/hr
CO2	1293.769	kmol/hr
	56937.485	kg/hr
CO2/Clinker	0.481	kg CO2/kg clinker
	1.053	m3 CO2/kg clinker

2. Fuel calculation

Total heat required	402223.40	MJ/hr
Coal flow rate in calciner	8906.4	kg/hr
fuel/precalcined meal	0.072	kg fuel /kg precalcined meal
precalcined meal/fuel	13.86	kg precalcined meal/kg fuel

3. Dust calculation

Dust from kiln	470.4	kg/hr
Ash from fuel calciner	1202.4	kg/hr

4. Air calculation

Oxygen req	0.07	Kmol/kg fuel
Oxygen required from air	2.20	kg/kg fuel

Coal

	kg/kg	MW	kmol/kg
C	0.722	12.011	0.060111564
H	0.04	1.008	0.03968254
O	0.057	15.999	0.003562723
S	0.012	32.065	0.00037424
N	0.016	14.007	0.001142286
Moisture	0.018	18.015	0.000999167
Ash	0.135	60	0.00225

$$C_\alpha H_\beta O_\gamma N_\delta S_\epsilon + \left(\alpha + \frac{\beta}{4} - \frac{\gamma}{2} + \epsilon\right)(O_2 + 3.77N_2) \rightarrow \alpha CO_2 + \frac{\beta}{2}H_2O + \epsilon SO_2 + \left(3.77\left(\alpha + \frac{\beta}{4} - \frac{\gamma}{2} + \epsilon\right) + \frac{\delta}{2}\right)N_2$$

O2 stoichiometric	0.068625078	kmol/kg fuel
	2.19586524	kg/kg fuel

O2 level in air 20.95 %
N2 level in air 79.05 %
N2/O2 ratio 3.77

Nitrogen in air 0.26 Kmol/kg fuel
7.25 kg/kg fuel

Total air 9.45 kg/kg fuel
Excess air 26.83 %
106723.92 kg/hr

Primary air 8537.91 kg/hr
Secondary air 94636.98 kg/hr
False air 3549.03003 kg/hr

5. Flue gas calculation

Flue gas

	MW	kg/kg fuel	Excess air (kg/kg fuel)	N2 in fuel (kg/kg fuel)	Moisture in fuel (kg/kg fuel)	Total Kg/kg fuel	Cp @ 1000K kJ/kg.K	Cpi x wi	mas flow kg/hr			w (kg/kg)	density	Volume flow
CO2	44.009	2.65				2.65	1.23	0.253	23561.36925	56937.5	80498.9	0.46975	0.45719	176074
H2O	18.015	0.36			0.018	0.38	2.28	0.067			3343.813806	0.019513	0.18715	17867.13
SO2	64.063	0.02				0.02	0.85	0.002			213.5296896	0.001246	0.66552	320.847
N2	28.014	7.25	1.95	0.02		9.21	1.16	0.832			82061.26752	0.478867	0.29102	281974.6
O2	31.998		0.59			0.59	1.09	0.050			5247.954409	0.030624	0.33241	15787.53

SUM 12.848 kg/kg fuel
Cp mixture 1.203 kJ/kg.K
Density of mixture 0.348 kg/m3

Flue gas from combustion 114427.9 kg/hr

Spesific energy of kiln gas to calciner (O2 enriched case) 0.2439 MW/ton precalcined meal
Spesific energy of kiln gas to calciner (ref case) 0.2108 MW/ton precalcined meal
Specific energy difference (reference case - o2 enriched case) -0.0332 MW/ton precalcined meal
-0.0332 MJ/s/ton precalcined meal
-119.3419 MJ/hr/ton precalcined meal
Total energy shortage in calciner -14736.5791 MJ/hr

Additional fuel needed at calciner -526 kg/hr
-0.53 ton/hr

Total fuel required at calciner 8380.07 kg/hr

Additional flue gas from calciner -6761.9 kg/hr
CO2 from calcination 56937.5 kg/hr
Flue gas total (calciner) 164603.5 kg/hr
472609.31 m3/hr
131.28 m3/s

O2 in flue gas 4937.8 kg/hr

3.00 %

Total Flue gas out from calciner

Total flue gas rate (Calciner flue gas + Rotary kiln flue gas)

216.01 m3/s

	Clinker Prod Cap T/hr	Calciner Flue gas flow rate m3/s	Cap increase %
Ref case (coal)	137000	242.44	-
O2 enriched cas	118301	216.01	-13.65

Appendix D - Procedure for modelling and simulation

Procedure for modelling and simulation using excel spreadsheet

I. Kiln mass balance calculation steps
1. Define the design basis value
2. Estimated the raw material flow rate
3. Calculate the precalcined meal and clinker production capacity from the estimated raw material
4. Calculate the CO2 flow rate produced from calcination process at kiln
5. Calculate the fuel rate required based on the estimated production capacity
6. Calculate the ash flow rate based on the calculated fuel rate
7. Calculate the dust flow rate based on the balance of clinker, ash and precalcined meal flow rate
8. Calculate the air flow rate required from the fuel flow rate according to stoichiometric ratio
9. Define/change the O2 level
10. Estimate the excess air required (should be done by goal seek to get the 3 % of oxygen in the exit gas composition)
11. Calculate the primary, secondary, and false air flow rate
12. Calculate the exit gas composition based on the stoichiometric ratio
13. Calculate the oxygen concentration in the kiln exit gas
14. Calculate the kiln exit gas flow rate
15. Set the mass balance around the kiln

II. Kiln energy balance calculation steps

16. Define the heat capacity value for precalcined meal, fuel, air, flue gas, dust, and clinker
17. Define the temperature of precalcined meal, fuel, air, flue gas, dust, and clinker
18. Calculate the energy flow of precalcined meal
19. Calculate the energy flow of fuel
20. Calculate the energy flow of primary air
21. Calculate the energy flow of Secondary air
22. Calculate the energy flow of false air
23. Calculate the generated energy from fuel combustion
24. Calculate the required energy for calcination
25. Calculate the energy flow of flue gas
26. Calculate the energy flow of dust
27. Calculate the energy flow of clinker
28. Estimate the energy loss
29. Set the energy balance around the rotary kiln

III. Calciner mass balance calculation steps

30. Define the design basis value

31. Referred to capacity calculated from kiln calculation, calculate the CO_2 flow rate produced from calcination process
32. Calculate the CO_2 flow rate produced from calcination process at calciner
33. Calculate the fuel rate required based on the estimated precalcined meal
34. Calculate the ash flow rate based on the calculated fuel rate
35. Calculate the dust flow rate based on the balance of raw meal, ash and precalcined meal flow rate.
36. Calculate the air flow rate required from the fuel flow rate according to the stoichiometric ratio
37. Define the O_2 level
38. Estimate the excess air required
39. Calculate the primary, secondary, and false air flow rate
40. Calculate the exit gas composition based on the stoichiometric ratio
41. Calculate the oxygen concentration in the calciner exit gas
42. Calculate the kiln calciner gas flow rate

IV. Simulation Procedure steps

Once the model has been built and confirmed works for base case, the model can be used for simulation of other cases as following steps;

43. Change the oxygen level
44. Goal seek the excess air required to get the 3 % of oxygen in the exit gas composition
45. Goal seek the raw material flow rate to have the calciner exit gas as required
46. Goal seek the specific energy value to balance the energy at required temperature.

Appendix E – Coal case simulation result

SIMULATION RESULT
COAL - CONSTANT CAPACITY CASE

		Rotary Kiln										Calciner											Pure O2
		Fuel		Exit Gas								Fuel		Exit Gas									
No.	O2 (%)	(kg/hr)	(ton/hr)	(m^3/h)	(kg/h)	(°C)	CO2	H2O	SO2	N2	O2	(kg/hr)	(ton/hr)	(m^3/h)	(kg/h)	Tot (m^3/h)	Tot (kg/h)	CO2	H2O	SO2	N2	O2	(kg/hr)
1	21	6322	6.3	304348	79687	1150	0.26	0.030	0.002	0.68	0.030	10314	10.3	567747	197838	872095	277525	0.41	0.023	0.001	0.53	0.030	0
2	22	6254	6.3	286082	75236	1150	0.28	0.031	0.002	0.66	0.030	10530	10.5	575323	200495	861405	275731	0.42	0.023	0.001	0.53	0.030	685.76
3	23	6200	6.2	270780	71509	1150	0.29	0.033	0.002	0.65	0.030	10704	10.7	581713	202722	852493	274231	0.42	0.023	0.001	0.52	0.030	1266.68
4	24	6151	6.2	257139	68187	1150	0.30	0.034	0.002	0.63	0.030	10860	10.9	587408	204707	844547	272894	0.43	0.023	0.001	0.52	0.030	1786.18
5	25	6108	6.1	244904	65207	1150	0.31	0.035	0.002	0.62	0.030	10999	11.0	592515	206486	837419	271694	0.43	0.023	0.001	0.51	0.030	2253.75
6	26	6069	6.1	233869	62520	1150	0.32	0.036	0.002	0.61	0.030	11124	11.1	597119	208091	830988	270611	0.44	0.023	0.001	0.51	0.030	2677.02
7	27	6033	6.0	223864	60083	1150	0.34	0.038	0.002	0.59	0.030	11238	11.2	601293	209545	825157	269628	0.44	0.024	0.002	0.50	0.030	3062.20
8	28	6001	6.0	214753	57864	1150	0.35	0.039	0.002	0.58	0.030	11342	11.3	605092	210869	819845	268734	0.44	0.024	0.002	0.50	0.030	3414.40
9	29	5971	6.0	206419	55835	1150	0.36	0.040	0.002	0.57	0.030	11437	11.4	605404	211128	811823	266963	0.45	0.024	0.001	0.50	0.030	3737.84
10	30	5944	5.9	198769	53972	1150	0.37	0.041	0.003	0.56	0.030	11524	11.5	609667	212562	808436	266534	0.45	0.024	0.002	0.49	0.030	4036.05
11	31	5919	5.9	191721	52256	1150	0.38	0.043	0.003	0.54	0.030	11604	11.6	612424	213531	804145	265787	0.45	0.024	0.002	0.49	0.030	4312.02
12	32	5896	5.9	185207	50669	1150	0.39	0.044	0.003	0.53	0.030	11678	11.7	614969	214426	800176	265095	0.46	0.024	0.002	0.49	0.030	4568.26
13	33	5875	5.9	179168	49199	1150	0.40	0.045	0.003	0.52	0.030	11746	11.7	617326	215254	796494	264453	0.46	0.024	0.002	0.49	0.030	4806.95
14	34	5855	5.9	173555	47832	1150	0.41	0.046	0.003	0.51	0.030	11810	11.8	619515	216024	793069	263856	0.46	0.024	0.002	0.48	0.030	5029.93
15	35	5837	5.8	168323	46558	1150	0.42	0.047	0.003	0.50	0.030	11869	11.9	621552	216740	789875	263298	0.46	0.025	0.002	0.48	0.030	5238.80
16	36	5820	5.8	163435	45368	1150	0.43	0.048	0.003	0.49	0.030	11925	11.9	623454	217409	786889	262777	0.47	0.025	0.002	0.48	0.030	5434.96
17	37	5804	5.8	158858	44253	1150	0.44	0.049	0.003	0.48	0.030	11976	12.0	625233	218035	784091	262288	0.47	0.025	0.002	0.48	0.030	5619.62
18	38	5789	5.8	154564	43207	1150	0.45	0.050	0.003	0.46	0.030	12025	12.0	626900	218621	781464	261828	0.47	0.025	0.002	0.47	0.030	5793.85
19	39	5775	5.8	150527	42224	1150	0.46	0.051	0.003	0.45	0.030	12071	12.1	628466	219172	778993	261396	0.47	0.025	0.002	0.47	0.030	5958.58
20	40	5762	5.8	146725	41299	1150	0.47	0.052	0.003	0.44	0.030	12114	12.1	629940	219690	776665	260989	0.47	0.025	0.002	0.47	0.030	6114.64

SIMULATION RESULT

COAL - INCREAASED CAPACITY CASE

		Rotary Kiln										Calciner															
		Fuel		Exit Gas								Fuel		Exit Gas										Pure O2	Cap		Cap increase
No.	O2 (%)	(kg/hr)	(ton/hr)	(m^3/h)	(kg/h)	(°C)	CO2	H2O	SO2	N2	O2	(kg/hr)	(ton/hr)	(m^3/h)	(kg/h)	Tot (m^3/h)	Tot (m^3/s)	Tot (kg/h)	CO2	H2O	SO2	N2	O2	(kg/hr)	kg/hr	t/hr	%
1	21	6322	6.3	304348	79687	1150	0.26	0.030	0.002	0.68	0.03	10314	10.3	567747	197838	872095	242	277525	0.41	0.023	0.001	0.53	0.03	0.00	137000	137.0	0.0
2	22	6326	6.3	289374	76105	1150	0.28	0.031	0.002	0.66	0.03	10680	10.7	583393	203307	872768	242	279412	0.42	0.023	0.001	0.53	0.03	693.60	138857	138.9	1.4
3	23	6307	6.3	275522	72766	1150	0.29	0.033	0.002	0.65	0.03	10936	10.9	597241	207981	872763	242	280747	0.42	0.023	0.001	0.52	0.03	1288.70	139825	139.8	2.1
4	24	6309	6.3	263773	69952	1150	0.30	0.034	0.002	0.63	0.03	11204	11.2	608994	212074	872768	242	282027	0.43	0.023	0.001	0.52	0.03	1831.90	141157	141.2	3.0
5	25	6310	6.3	253065	67388	1150	0.31	0.035	0.002	0.62	0.03	11448	11.4	619699	215802	872764	242	283190	0.43	0.023	0.001	0.51	0.03	2328.21	142370	142.4	3.9
6	26	6311	6.3	243275	65044	1150	0.32	0.036	0.002	0.61	0.03	11672	11.7	629488	219211	872764	242	284255	0.44	0.023	0.001	0.51	0.03	2783.74	143480	143.5	4.7
7	27	6311	6.3	234289	62892	1150	0.34	0.038	0.002	0.59	0.03	11877	11.9	638479	222342	872768	242	285234	0.44	0.024	0.002	0.50	0.03	3203.48	144500	144.5	5.5
8	28	6312	6.3	226010	60909	1150	0.35	0.039	0.002	0.58	0.03	12066	12.1	646758	225225	872768	242	286134	0.44	0.024	0.002	0.50	0.03	3591.66	145439	145.4	6.2
9	29	6313	6.3	218364	59079	1150	0.36	0.040	0.003	0.57	0.03	12241	12.2	654428	227896	872792	242	286975	0.45	0.024	0.002	0.50	0.03	3951.97	146311	146.3	6.8
10	30	6314	6.3	211265	57379	1150	0.37	0.041	0.003	0.56	0.03	12403	12.4	661503	230360	872768	242	287738	0.45	0.024	0.002	0.49	0.03	4287.14	147112	147.1	7.4
11	31	6315	6.3	204680	55802	1150	0.38	0.042	0.003	0.54	0.03	12553	12.6	668088	232653	872768	242	288455	0.45	0.024	0.002	0.49	0.03	4600.35	147859	147.9	7.9
12	32	6316	6.3	198534	54331	1150	0.39	0.044	0.003	0.53	0.03	12693	12.7	674234	234793	872768	242	289124	0.45	0.024	0.002	0.49	0.03	4893.35	148557	148.6	8.4
13	33	6317	6.3	192790	52955	1150	0.40	0.045	0.003	0.52	0.03	12824	12.8	679978	236793	872768	242	289749	0.46	0.024	0.002	0.49	0.03	5168.22	149210	149.2	8.9
14	34	6318	6.3	187418	51669	1150	0.41	0.046	0.003	0.51	0.03	12947	12.9	685350	238664	872768	242	290333	0.46	0.024	0.002	0.48	0.03	5426.96	149820	149.8	9.4
15	35	6318	6.3	182372	50461	1150	0.42	0.047	0.003	0.50	0.03	13062	13.1	690396	240421	872768	242	290882	0.46	0.024	0.002	0.48	0.03	5670.73	150394	150.4	9.8
16	36	6319	6.3	177632	49326	1150	0.43	0.048	0.003	0.49	0.03	13170	13.2	695136	242072	872768	242	291398	0.46	0.024	0.002	0.48	0.03	5901.17	150933	150.9	10.2
17	37	6320	6.3	173166	48257	1150	0.44	0.049	0.003	0.48	0.03	13272	13.3	699602	243627	872768	242	291884	0.46	0.024	0.002	0.48	0.03	6119.24	151441	151.4	10.5
18	38	6321	6.3	168952	47248	1150	0.45	0.050	0.003	0.46	0.03	13368	13.4	703816	245095	872768	242	292343	0.47	0.024	0.002	0.48	0.03	6326.05	151920	151.9	10.9
19	39	6321	6.3	164970	46295	1150	0.46	0.051	0.003	0.45	0.03	13459	13.5	707798	246481	872768	242	292776	0.47	0.025	0.002	0.48	0.03	6522.53	152373	152.4	11.2
20	40	6322	6.3	161201	45392	1150	0.47	0.052	0.003	0.44	0.03	13545	13.5	711567	247794	872768	242	293186	0.47	0.025	0.002	0.47	0.03	6709.51	152802	152.8	11.5

Appendix F – Waste fuel case simulation result

SIMULATION RESULT
WASTE FUEL - CONSTANT CAPACITY CASE

		Rotary Kiln										Calciner													
		Fuel		Exit Gas								Fuel		Exit Gas											Pure O2
No.	O2 (%)	(kg/hr)	(ton/hr)	(m^3/h)	(kg/h)	(°C)	CO2	H2O	SO2	N2	O2	(kg/hr)	(ton/hr)	(m^3/h)	(kg/h)	Tot (m^3/h)	Tot (kg/h)	CO2	H2O	SO2	N2			O2	(kg/hr)
1	21	7070	7.1	304884	78372	1150	0.25	0.05	0.002	0.674	0.030	8377	8.4	472422	164538	777307	242910	0.40	0.029	0.001	0.54	0.03	1.01	0.03	0
2	22	6991	7.0	286913	74005	1150	0.26	0.05	0.002	0.659	0.030	8582	8.6	477993	166576	764906	240581	0.41	0.029	0.002	0.53	0.03	1.00	0.03	671
3	23	6924	6.9	271631	70290	1150	0.27	0.05	0.002	0.646	0.030	8756	8.8	484386	168804	756017	239094	0.41	0.030	0.002	0.53	0.03	1.00	0.03	1238
4	24	6864	6.9	258039	66986	1150	0.28	0.06	0.002	0.632	0.030	8911	8.9	490070	170785	748108	237771	0.42	0.030	0.002	0.52	0.03	1.00	0.03	1745
5	25	6811	6.8	245856	64025	1150	0.29	0.06	0.002	0.619	0.030	9050	9.1	495163	172560	741019	236585	0.42	0.030	0.002	0.52	0.03	1.00	0.03	2200
6	26	6763	6.8	234888	61359	1150	0.30	0.06	0.002	0.606	0.030	9175	9.2	499746	174157	734634	235516	0.43	0.030	0.002	0.51	0.03	1.00	0.03	2612
7	27	6720	6.7	224957	58946	1150	0.31	0.06	0.003	0.593	0.030	9288	9.3	503894	175603	728851	234548	0.43	0.030	0.002	0.51	0.03	1.00	0.03	2986
8	28	6680	6.7	215923	56750	1150	0.32	0.06	0.003	0.580	0.030	9391	9.4	507667	176917	723589	233667	0.44	0.030	0.002	0.50	0.03	1.00	0.03	3328
9	29	6680	6.7	207669	54744	1150	0.33	0.07	0.003	0.568	0.030	9485	9.5	509590	177660	717259	232404	0.44	0.030	0.002	0.50	0.03	1.00	0.03	3641
10	30	6611	6.6	200099	52904	1150	0.34	0.07	0.003	0.556	0.030	9571	9.6	512479	178680	712578	231584	0.44	0.031	0.002	0.49	0.03	0.98	0.03	3930
11	31	6581	6.6	193131	51211	1150	0.35	0.07	0.003	0.544	0.030	9650	9.7	515137	179618	708269	230829	0.45	0.031	0.002	0.49	0.03	0.98	0.03	4197
12	32	6553	6.6	186697	49647	1150	0.36	0.07	0.003	0.532	0.030	9723	9.7	517591	180484	704288	230131	0.45	0.031	0.002	0.49	0.03	0.98	0.03	4445
13	33	6527	6.5	180736	48198	1150	0.37	0.07	0.003	0.520	0.030	9791	9.8	519863	181285	700599	229483	0.45	0.031	0.002	0.49	0.03	0.98	0.03	4675
14	34	6503	6.5	175199	46852	1150	0.38	0.08	0.003	0.508	0.030	9854	9.9	522569	182209	697768	229062	0.45	0.031	0.002	0.48	0.03	0.98	0.03	4891
15	35	6481	6.5	170041	45599	1150	0.39	0.08	0.003	0.497	0.030	9913	9.9	526794	183583	696836	229182	0.46	0.031	0.002	0.48	0.03	1.00	0.03	5092
16	36	6460	6.5	165226	44429	1150	0.40	0.08	0.003	0.486	0.030	9967	10.0	528798	184281	694024	228710	0.46	0.031	0.002	0.48	0.03	1.00	0.03	5281
17	37	6441	6.4	160720	43334	1150	0.41	0.08	0.003	0.475	0.030	10018	10.0	530672	184934	691392	228268	0.46	0.031	0.002	0.48	0.03	1.00	0.03	5460
18	38	6423	6.4	156494	42307	1150	0.42	0.08	0.003	0.464	0.030	10066	10.1	532428	185546	688923	227853	0.46	0.031	0.002	0.48	0.03	1.00	0.03	5628
19	39	6406	6.4	152524	41342	1150	0.43	0.08	0.003	0.454	0.030	10111	10.1	534078	186121	686602	227463	0.46	0.031	0.002	0.47	0.03	1.00	0.03	5786
20	40	6390	6.4	148786	40434	1150	0.44	0.09	0.003	0.443	0.030	10154	10.2	535630	186662	684416	227096	0.47	0.031	0.002	0.47	0.03	1.00	0.03	5937

SIMULATION RESULT

WASTE FUEL - INCREASED CAPACITY CASE

		Rotary Kiln										Calciner															
		Fuel		Exit Gas								Fuel		Exit Gas										Pure O2	Cap		Cap increase
No.	O2 (%)	(kg/hr)	(ton/hr)	(m^3/h)	(kg/h)	(°C)	CO2	H2O	SO2	N2	O2	(kg/hr)	(ton/hr)	(m^3/h)	(kg/h)	Tot (m^3/h)	Tot (m^3/s)	Tot (kg/h)	CO2	H2O	SO2	N2	O2	(kg/hr)	kg/hr	t/hr	%
1	21	7070	7.1	304884	78372	1150	0.25	0.05	0.002	0.67	0.03	8377	8.4	472422	164538	777307	216	242910	0.40	0.029	0.001	0.54	0.030	0.00	118301	118.3	-13.6
2	22	7439	7.4	305021	78696	1150	0.26	0.05	0.002	0.66	0.03	9307	9.3	517355	180293	822376	228	258989	0.41	0.029	0.002	0.53	0.030	713.02	127501	127.5	-6.9
3	23	7769	7.8	305021	78960	1150	0.27	0.05	0.002	0.64	0.03	10143	10.1	559453	194964	864475	240	273925	0.41	0.029	0.002	0.53	0.030	1389.53	135797	135.8	-0.9
4	24	7836	7.8	294841	76575	1150	0.28	0.06	0.002	0.63	0.03	10541	10.5	577926	201402	872768	242	277977	0.42	0.029	0.002	0.52	0.030	1991.91	138598	138.6	1.2
5	25	7838	7.8	283225	73793	1150	0.29	0.06	0.002	0.62	0.03	10807	10.8	589543	205450	872768	242	279244	0.42	0.030	0.002	0.51	0.030	2531.81	139931	139.9	2.1
6	26	7840	7.8	272631	71257	1150	0.30	0.06	0.002	0.60	0.03	11051	11.1	600137	209142	872768	242	280400	0.43	0.030	0.002	0.51	0.030	3027.75	141146	141.1	3.0
7	27	7842	7.8	262887	68924	1150	0.31	0.06	0.003	0.59	0.03	11274	11.3	609881	212538	872768	242	281462	0.43	0.030	0.002	0.51	0.030	3484.67	142265	142.3	3.8
8	28	7844	7.8	253906	66774	1150	0.32	0.06	0.003	0.58	0.03	11480	11.5	618861	215668	872768	242	282442	0.44	0.030	0.002	0.50	0.030	3907.30	143296	143.3	4.6
9	29	7846	7.8	245608	64787	1150	0.34	0.07	0.003	0.57	0.03	11671	11.7	627160	218560	872768	242	283347	0.44	0.030	0.002	0.50	0.030	4299.58	144249	144.2	5.3
10	30	7848	7.8	237913	62945	1150	0.35	0.07	0.003	0.55	0.03	11847	11.8	634854	221241	872768	242	284187	0.44	0.030	0.002	0.49	0.030	4664.80	145133	145.1	5.9
11	31	7865	7.9	231205	61351	1150	0.36	0.07	0.003	0.54	0.03	12039	12.0	641563	223670	872768	242	285022	0.45	0.030	0.002	0.49	0.030	5015.38	146283	146.3	6.8
12	32	7867	7.9	224530	59754	1150	0.37	0.07	0.003	0.53	0.03	12193	12.2	648238	225997	872768	242	285751	0.45	0.030	0.002	0.49	0.030	5335.39	147055	147.1	7.3
13	33	7868	7.9	218294	58261	1150	0.38	0.07	0.003	0.52	0.03	12336	12.3	654474	228171	872768	242	286432	0.45	0.031	0.002	0.49	0.030	5635.78	147777	147.8	7.9
14	34	7870	7.9	212454	56863	1150	0.38	0.08	0.003	0.51	0.03	12471	12.5	660313	230207	872768	242	287070	0.46	0.031	0.002	0.48	0.030	5918.42	148453	148.5	8.4
15	35	7872	7.9	206974	55551	1150	0.39	0.08	0.003	0.50	0.03	12597	12.6	665794	232118	872768	242	287669	0.46	0.031	0.002	0.48	0.030	6184.93	149087	149.1	8.8
16	36	7874	7.9	201821	54318	1150	0.40	0.08	0.003	0.48	0.03	12715	12.7	670947	233914	872768	242	288232	0.46	0.031	0.002	0.48	0.030	6436.75	149685	149.7	9.3
17	37	7875	7.9	196968	53156	1150	0.41	0.08	0.003	0.47	0.03	12827	12.8	675800	235607	872768	242	288763	0.46	0.031	0.002	0.48	0.030	6675.19	150247	150.2	9.7
18	38	7877	7.9	192388	52060	1150	0.42	0.08	0.003	0.46	0.03	12932	12.9	680380	237203	872768	242	289263	0.46	0.031	0.002	0.47	0.030	6901.37	150778	150.8	10.1
19	39	7878	7.9	188059	51024	1150	0.43	0.08	0.003	0.45	0.03	13032	13.0	684709	238712	872768	242	289737	0.47	0.031	0.002	0.47	0.030	7116.29	151281	151.3	10.4
20	40	7880	7.9	183962	50044	1150	0.44	0.09	0.003	0.44	0.03	13126	13.1	688806	240141	872768	242	290184	0.47	0.031	0.002	0.47	0.030	7320.86	151756	151.8	10.8

Appendix G – Relative NPV calculation result

NPV CALCULATION
COAL - CONSTANT CAPACITY CASE

Product price		0.5	KNOK/t
Coal price		1	KNOK/t
Waste fuel price		0.5	KNOK/t
O2 cost		1	KNOK/t
Interest rate		7.5	%
Periode		24	year
DCF factor		11.98	
USD/NOK		8.36	
Operational hour		7315	h/y
Capex	USD	NOK	
Storage	15000	125.4	KNOK
Vaporizer	3000	25.08	KNOK
Regulator valve	1000	8.36	KNOK
Transportation	2000	16.72	KNOK
Installation	10.00	15.884	KNOK
		191.44	KNOK

O2 level	capacity increase		normal case coal rate	O2 enriched case coal rate			coal rate increase	Pure oxygen rate		Income increase		cost increase				Net cash flow		NPV
				(kiln)	(calciner)	(total)						fuel cost	O2 cost	Total				
%	%	t/h	t/h	t/h	t/h	t/h	t/h	kg/h	t/h	KNOK/h	KNOK/y	KNOK/h	KNOK/h	KNOK/h	KNOK/y	KNOK/h	KNOK/y	NOK
21	0	0	16.64	6.32	10.31	16.64	0.00	0.00	0.00	0.00	0	0.00	0.00	0.0	0	0.0	0	-191
22	0	0	16.64	6.25	10.53	16.78	0.15	685.76	0.69	0.00	0	0.15	0.69	0.8	6102	-0.8	-6102	-73296
23	0	0	16.64	6.20	10.70	16.90	0.27	1266.68	1.27	0.00	0	0.27	1.27	1.5	11228	-1.5	-11228	-134705
24	0	0	16.64	6.15	10.86	17.01	0.38	1786.18	1.79	0.00	0	0.38	1.79	2.2	15810	-2.2	-15810	-189590
25	0	0	16.64	6.11	11.00	17.11	0.47	2253.75	2.25	0.00	0	0.47	2.25	2.7	19931	-2.7	-19931	-238960
26	0	0	16.64	6.07	11.12	17.19	0.56	2677.02	2.68	0.00	0	0.56	2.68	3.2	23659	-3.2	-23659	-283626
27	0	0	16.64	6.03	11.24	17.27	0.64	3062.20	3.06	0.00	0	0.64	3.06	3.7	27050	-3.7	-27050	-324247
28	0	0	16.64	6.00	11.34	17.34	0.71	3414.40	3.41	0.00	0	0.71	3.41	4.1	30148	-4.1	-30148	-361364
29	0	0	16.64	5.97	11.44	17.41	0.77	3737.84	3.74	0.00	0	0.77	3.74	4.5	32991	-4.5	-32991	-395426
30	0	0	16.64	5.94	11.52	17.47	0.83	4036.05	4.04	0.00	0	0.83	4.04	4.9	35611	-4.9	-35611	-426809
31	0	0	16.64	5.92	11.60	17.52	0.89	4312.02	4.31	0.00	0	0.89	4.31	5.2	38033	-5.2	-38033	-455830
32	0	0	16.64	5.90	11.68	17.57	0.94	4568.26	4.57	0.00	0	0.94	4.57	5.5	40281	-5.5	-40281	-482756
33	0	0	16.64	5.88	11.75	17.62	0.99	4806.95	4.81	0.00	0	0.99	4.81	5.8	42373	-5.8	-42373	-507817
34	0	0	16.64	5.86	11.81	17.67	1.03	5029.93	5.03	0.00	0	1.03	5.03	6.1	44325	-6.1	-44325	-531210
35	0	0	16.64	5.84	11.87	17.71	1.07	5238.80	5.24	0.00	0	1.07	5.24	6.3	46153	-6.3	-46153	-553104
36	0	0	16.64	5.82	11.92	17.74	1.11	5434.96	5.43	0.00	0	1.11	5.43	6.5	47868	-6.5	-47868	-573649
37	0	0	16.64	5.80	11.98	17.78	1.14	5619.62	5.62	0.00	0	1.14	5.62	6.8	49481	-6.8	-49481	-592972
38	0	0	16.64	5.79	12.03	17.81	1.18	5793.85	5.79	0.00	0	1.18	5.79	7.0	51001	-7.0	-51001	-611187
39	0	0	16.64	5.77	12.07	17.85	1.21	5958.58	5.96	0.00	0	1.21	5.96	7.2	52438	-7.2	-52438	-628394
40	0	0	16.64	5.76	12.11	17.88	1.24	6114.64	6.11	0.00	0	1.24	6.11	7.4	53797	-7.4	-53797	-644679

NPV CALCULATION
COAL - INCREASED CAPACITY CASE

Product price		0.5	NOK/t
Coal price		1	NOK/t
Waste fuel price		0.5	NOK/t
O2 cost		1	NOK/t
Interest rate		7.5	%
Periode		24	year
DCF factor		11.98	
USD/NOK		8.36	
Operational hour		7315	h/y
Capex	USD	NOK	
Storage	15000	125.4	KNOK
Vaporizer	3000	25.08	KNOK
Regulator valve	1000	8.36	KNOK
Transportation	2000	16.72	KNOK
Installation	10.00	15.884	KNOK
		191.44	KNOK

O2 level	capacity increase			normal O2 case coal rate	O2 enriched case coal rate			coal rate increase	Pure oxygen rate		Income increase		cost increase				Net cash flow		NPV	Cap increase/ oxygen flow rate	Ratio Cap increase/ O2 increase
					(kiln)	(calciner)	(total)						fuel cost	O2 cost	Total	Total					
%	%		t/h	t/h	t/h	t/h	t/h	t/h	kg/h	t/h	KNOK/h	KNOK/y	KNOK/h	KNOK/h	KNOK/h	KNOK/y	KNOK/h	KNOK/y	NOK		
21	0.000	137.00	0.0	16.64	6.32	10.31	16.64	0.00	0.00	0.0	0.00	0	0.00	0.00	0.0	0	0.00	0	-191		0.0
22	1.36	138.86	1.9	16.94	6.33	10.68	17.01	0.06	693.60	0.7	0.93	6794	0.06	0.69	0.8	5520	0.17	1274	15066	2.7	1.4
23	2.06	139.83	2.8	17.11	6.31	10.94	17.24	0.14	1288.70	1.3	1.41	10333	0.14	1.29	1.4	10432	-0.01	-99	-1379	2.2	0.7
24	3.03	141.16	4.2	17.33	6.31	11.20	17.51	0.19	1831.90	1.8	2.08	15205	0.19	1.83	2.0	14755	0.06	450	5198	2.3	1.0
25	3.92	142.37	5.4	17.53	6.31	11.45	17.76	0.23	2328.21	2.3	2.68	19641	0.23	2.33	2.6	18705	0.13	936	11022	2.3	0.9
26	4.73	143.48	6.5	17.71	6.31	11.67	17.98	0.27	2783.74	2.8	3.24	23700	0.27	2.78	3.1	22328	0.19	1371	16236	2.3	0.8
27	5.47	144.50	7.5	17.88	6.31	11.88	18.19	0.31	3203.48	3.2	3.75	27430	0.31	3.20	3.5	25666	0.24	1764	20936	2.3	0.7
28	6.16	145.44	8.4	18.04	6.31	12.07	18.38	0.34	3591.66	3.6	4.22	30864	0.34	3.59	3.9	28752	0.29	2112	25113	2.3	0.7
29	6.80	146.31	9.3	18.18	6.31	12.24	18.55	0.37	3951.97	4.0	4.66	34054	0.37	3.95	4.3	31614	0.33	2440	29042	2.4	0.6
30	7.38	147.11	10.1	18.32	6.31	12.40	18.72	0.40	4287.14	4.3	5.06	36984	0.40	4.29	4.7	34278	0.37	2706	32222	2.4	0.6
31	7.93	147.86	10.9	18.44	6.32	12.55	18.87	0.43	4600.35	4.6	5.43	39718	0.43	4.60	5.0	36767	0.40	2952	35168	2.4	0.5
32	8.44	148.56	11.6	18.56	6.32	12.69	19.01	0.45	4893.35	4.9	5.78	42271	0.45	4.89	5.3	39093	0.43	3178	37880	2.4	0.5
33	8.91	149.21	12.2	18.67	6.32	12.82	19.14	0.47	5168.22	5.2	6.10	44657	0.47	5.17	5.6	41274	0.46	3383	40337	2.4	0.5
34	9.36	149.82	12.8	18.77	6.32	12.95	19.26	0.50	5426.96	5.4	6.41	46890	0.50	5.43	5.9	43326	0.49	3564	42500	2.4	0.4
35	9.78	150.39	13.4	18.86	6.32	13.06	19.38	0.52	5670.73	5.7	6.70	48987	0.52	5.67	6.2	45259	0.51	3728	44472	2.4	0.4
36	10.17	150.93	13.9	18.95	6.32	13.17	19.49	0.54	5901.17	5.9	6.97	50959	0.54	5.90	6.4	47086	0.53	3873	46205	2.4	0.4
37	10.54	151.44	14.4	19.04	6.32	13.27	19.59	0.55	6119.24	6.1	7.22	52816	0.55	6.12	6.7	48813	0.55	4003	47765	2.4	0.4
38	10.89	151.92	14.9	19.12	6.32	13.37	19.69	0.57	6326.05	6.3	7.46	54570	0.57	6.33	6.9	50451	0.56	4119	49151	2.4	0.3
39	11.22	152.37	15.4	19.19	6.32	13.46	19.78	0.59	6522.53	6.5	7.69	56228	0.59	6.52	7.1	52007	0.58	4221	50377	2.4	0.3
40	11.53	152.80	15.8	19.26	6.32	13.54	19.87	0.60	6709.51	6.7	7.90	57797	0.60	6.71	7.3	53486	0.59	4311	51457	2.4	0.3

NPV at O2 price				
1.3 NOK	1.2 NOK	1.1 NOK	1 NOK	0.9 NOK
-176	-175.719	-175.719	-175.719	-175.719
-3153	2925.463	9003.723	15081.98	21160.24
-35243	-23949.8	-12656.4	-1363.05	9930.309
-42947	-26893.5	-10840	5213.607	21267.19
-50171	-29768.5	-9365.55	11037.38	31440.31
-56933	-32538.2	-8143.21	16251.76	40646.72
-63268	-35194.6	-7121.29	20952.02	49025.33
-69297	-37821.6	-6346.5	25128.55	56603.6
-74840	-40207.7	-5575.1	29057.48	63690.05
-80472	-42902.2	-5332.34	32237.49	69807.31
-85760	-45445.6	-5131.03	35183.51	75498.05
-90750	-47868.2	-4985.99	37896.22	80778.44
-95521	-50229.7	-4938.68	40352.36	85643.41
-100160	-52601.5	-5043.03	42515.43	90073.9
-104596	-54901.6	-5206.89	44487.82	94182.53
-108921	-57207.2	-5493.05	46221.11	97935.27
-113095	-59469.8	-5844.64	47780.54	101405.7
-117146	-61708.7	-6271.14	49166.4	104603.9
-121085	-63925.8	-6766.44	50392.93	107552.3
-124921	-66122.9	-7325.01	51472.91	110270.8

NPV CALCULATION
WASTE FUEL - CONSTANT CAPACITY CASE

Product price		0.5	NOK/t
Coal price		1	NOK/t
Waste fuel price		0.5	NOK/t
O2 cost		1	NOK/t
Interest rate		7.5	%
Periode		24	year
DCF factor		11.98	
USD/NOK		8.36	
Operational hour		7315	h/y
Capex	USD	NOK	
Storage	15000	125.4	KNOK
Vaporizer	3000	25.08	KNOK
Regulator valve	1000	8.36	KNOK
Transportation	2000	16.72	KNOK
Installation	10.00	15.884	KNOK
		191.44	KNOK

O2 level	capacity increase		base case		O2 enriched case coal rate		waste rate increase	coal rate increase	Pure oxygen rate		income increase		cost increase						Net cash flow		NPV
			waste	coal	waste	coal							waste	coal	total fuel	O2 cost	total cost				
%	%	t/h	t/h	t/h	t/h	t/h	t/h	t/h	kg/h	t/h	KNOK/h	KNOK/y	KNOK/h	KNOK/h	KNOK/h	KNOK/h	KNOK/h	KNOK/y	KNOK/h	KNOK/y	NOK
21	0.00	0.00	7.07	8.38	7.07	8.38	0.00	0.00	0.00	0.00	0.00	0	0.00	0.00	0.00	0.00	0.0	0	0.00	0	-191
22	0.00	0.00	7.07	8.38	6.99	8.58	-0.08	0.21	671.08	0.67	0.00	0	-0.04	0.21	0.17	0.67	0.8	6120	-0.84	-6120	-73505
23	0.00	0.00	7.07	8.38	6.92	8.76	-0.15	0.38	1238.45	1.24	0.00	0	-0.07	0.38	0.31	1.24	1.5	11300	-1.54	-11300	-135565
24	0.00	0.00	7.07	8.38	6.86	8.91	-0.21	0.53	1745.02	1.75	0.00	0	-0.10	0.53	0.43	1.75	2.2	15921	-2.18	-15921	-190930
25	0.00	0.00	7.07	8.38	6.81	9.05	-0.26	0.67	2200.20	2.20	0.00	0	-0.13	0.67	0.54	2.20	2.7	20072	-2.74	-20072	-240653
26	0.00	0.00	7.07	8.38	6.76	9.18	-0.31	0.80	2611.76	2.61	0.00	0	-0.15	0.80	0.64	2.61	3.3	23821	-3.26	-23821	-285571
27	0.00	0.00	7.07	8.38	6.72	9.29	-0.35	0.91	2985.84	2.99	0.00	0	-0.18	0.91	0.74	2.99	3.7	27227	-3.72	-27227	-326366
28	0.00	0.00	7.07	8.38	6.68	9.39	-0.39	1.01	3327.52	3.33	0.00	0	-0.19	1.01	0.82	3.33	4.1	30334	-4.15	-30334	-363597
29	0.00	0.00	7.07	8.38	6.68	9.48	-0.39	1.11	3641.00	3.64	0.00	0	-0.19	1.11	0.91	3.64	4.6	33315	-4.55	-33315	-399302
30	0.00	0.00	7.07	8.38	6.61	9.57	-0.46	1.19	3929.79	3.93	0.00	0	-0.23	1.19	0.96	3.93	4.9	35805	-4.89	-35805	-429138
31	0.00	0.00	7.07	8.38	6.58	9.65	-0.49	1.27	4196.83	4.20	0.00	0	-0.24	1.27	1.03	4.20	5.2	38228	-5.23	-38228	-458157
32	0.00	0.00	7.07	8.38	6.55	9.72	-0.52	1.35	4444.61	4.44	0.00	0	-0.26	1.35	1.09	4.44	5.5	40473	-5.53	-40473	-485059
33	0.00	0.00	7.07	8.38	6.53	9.79	-0.54	1.41	4675.26	4.68	0.00	0	-0.27	1.41	1.14	4.68	5.8	42561	-5.82	-42561	-510076
34	0.00	0.00	7.07	8.38	6.50	9.85	-0.57	1.48	4890.61	4.89	0.00	0	-0.28	1.48	1.19	4.89	6.1	44509	-6.08	-44509	-533409
35	0.00	0.00	7.07	8.38	6.48	9.91	-0.59	1.54	5092.23	5.09	0.00	0	-0.29	1.54	1.24	5.09	6.3	46331	-6.33	-46331	-555233
36	0.00	0.00	7.07	8.38	6.46	9.97	-0.61	1.59	5281.49	5.28	0.00	0	-0.30	1.59	1.29	5.28	6.6	48039	-6.57	-48039	-575696
37	0.00	0.00	7.07	8.38	6.44	10.02	-0.63	1.64	5459.58	5.46	0.00	0	-0.31	1.64	1.33	5.46	6.8	49644	-6.79	-49644	-594931
38	0.00	0.00	7.07	8.38	6.42	10.07	-0.65	1.69	5627.53	5.63	0.00	0	-0.32	1.69	1.37	5.63	7.0	51157	-6.99	-51157	-613051
39	0.00	0.00	7.07	8.38	6.41	10.11	-0.66	1.73	5786.27	5.79	0.00	0	-0.33	1.73	1.40	5.79	7.2	52585	-7.19	-52585	-630158
40	0.00	0.00	7.07	8.38	6.39	10.15	-0.68	1.78	5936.61	5.94	0.00	0	-0.34	1.78	1.44	5.94	7.4	53936	-7.37	-53936	-646340

NPV CALCULATION
WASTE FUEL - INCREASED CAPACITY CASE

Product price		0.5	NOK/t
Coal price		1	NOK/t
Waste fuel price		0.5	NOK/t
O2 cost		1	NOK/t
Interest rate		7.5	%
Periode		24	year
DCF factor		11.98	
USD/NOK		8.36	
Operational hour		7315	h/y
Capex	USD	NOK	
Storage	15000	125.4	KNOK
Vaporizer	3000	25.08	KNOK
Regulator valve	1000	8.36	KNOK
Transportation	2000	16.72	KNOK
Installation	10.00	15.884	KNOK
		191.44	KNOK
		8.19	10.3

O2 level	capacity increase			Normal O2 case		O2 enriched case coal rate		waste rate increase	coal rate increase	Pure oxygen rate		income increase		cost increase						Net cash flow		NPV	Cap increase/oxygen flow rate	Ratio Cap increase/ O2
				waste	coal	waste	coal							waste	coal	total fuel	O2 cost	total cost						
%	%		t/h	t/h	t/h	t/h	t/h	t/h	t/h	kg/h	t/h	KNOK/h	KNOK/y	KNOK/h	KNOK/h	KNOK/h	KNOK/h	KNOK/h	KNOK/y	KNOK/h	KNOK/y	NOK		
21	-13.65	118.30	-18.7	7.01	8.91	7.07	8.38	0.06	-0.53	0.00	0.0	-9.35	-68392	0.03	-0.53	-0.50	0.00	-0.5	-3639	-8.85	-64753	-775927		0.00
22	-6.93	127.50	-9.5	7.55	9.60	6.99	8.58	-0.56	-1.02	713.02	0.7	-4.75	-34744	-0.28	-1.02	-1.30	0.71	-0.6	-4272	-4.17	-30473	-365252	-13.32	6.72
23	-0.88	135.80	-1.2	8.04	10.22	6.92	8.76	-1.12	-1.47	1389.53	1.4	-0.60	-4401	-0.56	-1.47	-2.03	1.39	-0.6	-4659	0.04	257	2890	-0.87	6.06
24	1.17	138.60	1.6	8.21	10.43	6.86	8.91	-1.34	-1.52	1991.91	2.0	0.80	5846	-0.67	-1.52	-2.20	1.99	-0.2	-1486	1.00	7332	87641	0.80	2.04
25	2.14	139.93	2.9	8.29	10.53	6.81	9.05	-1.48	-1.48	2531.81	2.5	1.47	10718	-0.74	-1.48	-2.22	2.53	0.3	2262	1.16	8456	101116	1.16	0.97
26	3.03	141.15	4.1	8.36	10.63	6.76	9.18	-1.60	-1.45	3027.75	3.0	2.07	15165	-0.80	-1.45	-2.25	3.03	0.8	5696	1.29	9469	113252	1.37	0.89
27	3.84	142.26	5.3	8.43	10.71	6.72	9.29	-1.71	-1.42	3484.67	3.5	2.63	19256	-0.85	-1.42	-2.27	3.48	1.2	8849	1.42	10407	124485	1.51	0.82
28	4.60	143.30	6.3	8.49	10.79	6.68	9.39	-1.81	-1.40	3907.30	3.9	3.15	23027	-0.90	-1.40	-2.30	3.91	1.6	11757	1.54	11270	134819	1.61	0.75
29	5.29	144.25	7.2	8.54	10.86	6.68	9.48	-1.86	-1.37	4299.58	4.3	3.62	26514	-0.93	-1.37	-2.31	4.30	2.0	14583	1.63	11931	142740	1.69	0.70
30	5.94	145.13	8.1	8.59	10.93	6.61	9.57	-1.98	-1.36	4664.80	4.7	4.07	29747	-0.99	-1.36	-2.35	4.66	2.3	16954	1.75	12793	153070	1.74	0.65
31	6.78	146.28	9.3	8.66	11.01	6.58	9.65	-2.08	-1.36	5015.38	5.0	4.64	33951	-1.04	-1.36	-2.40	5.02	2.6	19106	2.03	14845	177656	1.85	0.84
32	7.34	147.05	10.1	8.71	11.07	6.55	9.72	-2.16	-1.35	5335.39	5.3	5.03	36775	-1.08	-1.35	-2.43	5.34	2.9	21287	2.12	15488	185360	1.88	0.56
33	7.87	147.78	10.8	8.75	11.13	6.53	9.79	-2.22	-1.33	5635.78	5.6	5.39	39415	-1.11	-1.33	-2.45	5.64	3.2	23331	2.20	16084	192494	1.91	0.53
34	8.36	148.45	11.5	8.79	11.18	6.50	9.85	-2.29	-1.32	5918.42	5.9	5.73	41888	-1.14	-1.32	-2.47	5.92	3.5	25253	2.27	16636	199104	1.94	0.49
35	8.82	149.09	12.1	8.83	11.22	6.48	9.91	-2.35	-1.31	6184.93	6.2	6.04	44210	-1.17	-1.31	-2.49	6.18	3.7	27062	2.34	17148	205244	1.95	0.46
36	9.26	149.68	12.7	8.86	11.27	6.46	9.97	-2.40	-1.30	6436.75	6.4	6.34	46394	-1.20	-1.30	-2.50	6.44	3.9	28769	2.41	17625	210951	1.97	0.44
37	9.67	150.25	13.2	8.90	11.31	6.44	10.02	-2.46	-1.29	6675.19	6.7	6.62	48452	-1.23	-1.29	-2.52	6.68	4.2	30385	2.47	18067	216254	1.98	0.41
38	10.06	150.78	13.8	8.93	11.35	6.42	10.07	-2.51	-1.29	6901.37	6.9	6.89	50395	-1.25	-1.29	-2.54	6.90	4.4	31916	2.53	18479	221187	2.00	0.39
39	10.42	151.28	14.3	8.96	11.39	6.41	10.11	-2.55	-1.28	7116.29	7.1	7.14	52231	-1.28	-1.28	-2.55	7.12	4.6	33369	2.58	18862	225780	2.01	0.37
40	10.77	151.76	14.8	8.99	11.43	6.39	10.15	-2.60	-1.27	7320.86	7.3	7.38	53971	-1.30	-1.27	-2.57	7.32	4.8	34752	2.63	19219	230058	2.02	0.35

NPV at O2 price						
1.6 NOK	1.5 NOK	1.4 NOK	1.3 NOK	1.2 NOK	1.1 NOK	1 NOK
-775927	-775927	-775927	-775927	-775927	-775927	-775927
-402743	-396495	-390246	-383998	-377749	-371501	-365252
-70172	-57995.1	-45818	-33641.2	-21464.3	-9287.31	2889.643
-17094	361.8089	17818	35273.48	52729.32	70185.16	87641
-32007	-9820.24	12367	34554.18	56741.39	78928.61	101115.8
-45948	-19415.1	7118	33651.57	60184.89	86718.21	113251.5
-58739	-28202	2336	32873.02	63410.51	93948	124485.5
-70627	-36386.3	-2145	32095.95	66337.06	100578.2	134819.3
-83333	-45654.6	-7976	29703.05	67381.87	105060.7	142739.5
-92207	-51327.1	-10448	30431.71	71311.11	112190.5	153069.9
-86053	-42101.8	1850	45801.37	89752.97	133704.6	177656.2
-95176	-48420.1	-1664	45091.89	91847.91	138603.9	185359.9
-103836	-54447.8	-5059	44328.99	93717.4	143105.8	192494.2
-112088	-60222.6	-8357	43508.07	95373.41	147238.8	199104.1
-119961	-65760.2	-11559	42641.47	96842.33	151043.2	205244
-127495	-71086.9	-14679	41728.43	98136.09	154543.7	210951.4
-134729	-76232	-17735	40762.35	99259.55	157756.8	216254
-141688	-81208.9	-20730	39749.58	100228.9	160708.1	221187.4
-148396	-86033.4	-23671	38691.98	101054.7	163417.4	225780.1
-154875	-90719.3	-26564	37591.44	101746.8	165902.2	230057.6