Energy Efficiency and Carbon Management in Mineral Processing Plants

A Master of Science Thesis

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April 2014

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# TABLE OF CONTENTS

TABLE OF CONTENTS ........................................................................................................................... iv
LIST OF TABLES .......................................................................................................................................... ix
NOMENCLATURE ........................................................................................................................................ x
ACRONYMS AND ABBREVIATIONS ........................................................................................................... xi
AKNOWLEDGEMENTS ................................................................................................................................. xiii
ABSTRACT ................................................................................................................................................... xiv
CHAPTER 1 .................................................................................................................................................. 1
  1.1 Background ........................................................................................................................................... 1
    1.1.1 Nchanga smelter ....................................................................................................................... 1
      1.1.1.1 Flash smelting furnace ................................................................................................. 3
      1.1.1.2 Other furnaces and heat points ..................................................................................... 3
    1.1.2 Steam Distribution of steam from the WHRB .............................................................................. 4
      1.1.2.1 Steam dryer .................................................................................................................... 4
      1.1.2.2 Steam heat exchanger nitrogen heating ....................................................................... 4
      1.1.2.3 Air cooled condenser ..................................................................................................... 4
    1.1.3 Nkana Refinery ............................................................................................................................. 4
    1.1.4 Nchanga tailings leach plant ....................................................................................................... 5
  1.2 Problem Statement ................................................................................................................................. 6
  1.2.1 Aim of this work ............................................................................................................................ 6
  1.3 Methodology ......................................................................................................................................... 6
  1.4 Significance of the study ..................................................................................................................... 7
  1.5 Motivation to undertake the thesis subject ........................................................................................ 7
CHAPTER 2 .................................................................................................................................................. 8
LITERATURE REVIEW ................................................................................................................................. 8
  2.1 Energy management and carbon strategies in copper ........................................................................... 8
  2.2 Carbon management ........................................................................................................................... 9
  2.3 Carbon management in the global village .......................................................................................... 10
CHAPTER 2 ..............................................................................................................................................11
2.1 Waste heat scenarios in copper smelters .................................................................................................11
2.1.1 Smelter flue gas heat recovery ..............................................................................................................11
2.1.2 Electric furnaces ....................................................................................................................................12
2.1.3 Anode furnaces .......................................................................................................................................12
2.1.4 Waste heat recovery boilers .................................................................................................................13
2.2 Other waste heat recovery equipment ........................................................................................................13
2.2.1 Air-To-Gas Exchangers (Recuperators) ..................................................................................................14
2.2.2 Organic Rankine Cycle (ORC) Heat Recovery and Power Generation ..................................................14
2.2.3 Thermal Oil Heat Exchangers ...............................................................................................................14
2.2.4 Thermal Storage and Power Generation ..............................................................................................16
2.3 Benefits of process flue gas heat recovery ..................................................................................................17
2.4 Challenges and solutions in recovering waste heat ....................................................................................18
2.4.1 Example of waste heat recovery in a plant similar to the smelter ..........................................................18
2.4.1.1 The Kennecott Utah Copper Smelter ..............................................................................................18
2.4.2 other examples of waste heat recovery ..................................................................................................19
2.4.2.1 Sterlite Industries India Limited (Siil), Tuticorin ...........................................................................19
2.4.2.2 Anhui Huaibei Mining (Group) Cement Co. Limited, China ..........................................................19
2.5 Integrated efforts aimed at energy efficiency improvement and carbon management in copper processing ..........................................................................................................20
CHAPTER 3 ...............................................................................................................................................21
3.1 The Nchanga smelter complex ....................................................................................................................21
3.2 Nchanga smelter waste heat scenarios ........................................................................................................21
3.2.1 Flash smelting furnace and the waste heat recovery boiler .....................................................................21
3.2.2 Slag cleaning furnace ............................................................................................................................22
3.2.3 Cobalt recovery furnace .......................................................................................................................24
3.2.4 Anode furnaces ....................................................................................................................................24
3.3 Available energy extractable from smelter waste heat .............................................................................24
3.3.1 Installation of heat recovery steam generator for SCF and CRF .......................................................25
3.3.1.1 Slag cleaning furnace waste heat and HRSG installation .................................................................27
3.3.1.2 Summary of steam generated from envisaged HRSG .................................................................33
3.3.1.3 Power generation from HRSG steam at SCF .................................................................................34
v
LIST OF FIGURES

Fig. 1.1: Outlook of KCM Nchanga site operations..................................................2
Fig. 2.1: Radiative and convective sections across the WHRB...............................13
Fig. 2.2: Organic Rankine cycle – direct heat exchange.......................................15
Fig. 2.3: Organic Rankine cycle – indirect heat exchange.....................................15
Fig. 2.4: Thermal oil heat recovery system...........................................................16
Fig. 2.5: Thermal storage heat recovery system......................................................17
Fig. 3.1: Nchanga smelter complex and its installations.......................................23
Fig. 3.2: Incinerator, HRSG and steam turbine for SCF waste heat.......................26
Fig. 3.3: Flue gas and combustion air flow into the SCF incinerator.......................27
Fig. 3.4: SCF, HRSG and steam turbine configuration.........................................33
Fig. 3.5: FSF, HRSG, steam turbine, nitrogen heat exchanger and dryer configuration.........................................................................................................................37
Fig. 3.6: FSF, WHRB, superheat boiler and steam turbine configuration..............41
Fig. 4.1: Schematic of a Pachucas’ ferric reactor....................................................47
Fig. 4.2: Heat exchanger for Pachucas’ compressed air........................................48
Fig. 4.3: Electrolyte flow diagram for TLP............................................................50
Fig. 4.4: Electrolyte flow with envisaged heat exchangers....................................52
Fig. 4.5: Heat balance diagram for steam and electrolyte....................................53
Fig. 5.1: Electrolyte flow diagram for Nkana refinery electrolyte..........................60
Fig. 5.2: Heat balance diagram for electrolyte and steam for Nkana refinery.........61
Fig. A1: Schematic drawing of an electrolytic cell for copper deposition...............70
LIST OF TABLES

Table 3.1: Flue gas composition on wet and dry basis..................................................28
Table 3.2: A tabulated procedural calculation to determine number of moles for each element in the flue..............................................................................................................28
Table 3.3: Conversion of reactants to moles, where the results are based on the formulae in table 3.2.........................................................................................................................29
Table 3.4: Summary of gas composition after combustion in the incinerator..............30
Table 3.5: SCF gas composition and respective molar masses ..................................31
Table 3.6: A summary of design steam parameters across the HRSG.......................32
Table 3.7: Summary of results for potential steam parameters in the envisaged HRSG.................................................................33
## NOMENCLATURE

<table>
<thead>
<tr>
<th>Symbol</th>
<th>Description</th>
<th>Units</th>
</tr>
</thead>
<tbody>
<tr>
<td>$m_{\text{gas}}$</td>
<td>mass flow of flue gas</td>
<td>kg/s</td>
</tr>
<tr>
<td>$\overline{c}_{p, \text{gas}}$</td>
<td>average specific heat of gas</td>
<td>kJ/kgK</td>
</tr>
<tr>
<td>$T_{1, \text{gas}}$</td>
<td>inlet temperature of gas into the HRSG</td>
<td>°C</td>
</tr>
<tr>
<td>$T_{4, \text{gas}}$</td>
<td>exit temperature of gas from the HRSG</td>
<td>°C</td>
</tr>
<tr>
<td>$m_{\text{s}}$</td>
<td>mass flow rate of steam</td>
<td>kg/s</td>
</tr>
<tr>
<td>$h_{\text{sh}}$</td>
<td>enthalpy of superheated steam</td>
<td>kJ/kg</td>
</tr>
<tr>
<td>$h_{\text{fw}}$</td>
<td>enthalpy of liquid water</td>
<td>kJ/kg</td>
</tr>
<tr>
<td>$[O_2]$</td>
<td>number of moles of oxygen per unit mass of substance</td>
<td>mols/kg</td>
</tr>
<tr>
<td>$[N_{2, \text{air}}]$</td>
<td>number of moles of nitrogen per unit mass of air</td>
<td>mols/kg</td>
</tr>
<tr>
<td>$[N_2]$</td>
<td>number of moles of nitrogen per given mass of substance</td>
<td>mols/kg</td>
</tr>
<tr>
<td>$I_o$</td>
<td>molar amount of air needed for stoichiometric combustion</td>
<td>mols/kg</td>
</tr>
<tr>
<td>$g_o$</td>
<td>stoichiometric amount of gas</td>
<td>mols/kg</td>
</tr>
<tr>
<td>$m_{\text{air}}$</td>
<td>mass flow rate of air</td>
<td>kg/s</td>
</tr>
<tr>
<td>$RMM$</td>
<td>relative molecular mass of substance</td>
<td>g/mol</td>
</tr>
<tr>
<td>$a$</td>
<td>percentage composition of a gas</td>
<td>-</td>
</tr>
<tr>
<td>$V$</td>
<td>volume flow rate of flue gas</td>
<td>$m_n^3/h$</td>
</tr>
<tr>
<td>$m_{\text{tot}}$</td>
<td>total mass flow rate of gas</td>
<td>kg/s</td>
</tr>
<tr>
<td>$T_{\text{ave}}$</td>
<td>average temperature of gas across the HRSG</td>
<td>°C</td>
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<tr>
<td>$T_{\text{fw}}$</td>
<td>temperature of liquid water</td>
<td>°C</td>
</tr>
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<td>$T_{\text{sh}}$</td>
<td>temperature of saturated steam</td>
<td>°C</td>
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<tr>
<td>$h_{\text{fw}}$</td>
<td>enthalpy of liquid water</td>
<td>kJ/kg</td>
</tr>
<tr>
<td>$Q_{\text{gas}}$</td>
<td>energy content of gas</td>
<td>kW</td>
</tr>
<tr>
<td>$P_T$</td>
<td>power output of a steam turbine</td>
<td>MW</td>
</tr>
<tr>
<td>$\eta_{\text{mech}}$</td>
<td>mechanical efficiency of a steam turbine</td>
<td>-</td>
</tr>
<tr>
<td>$\eta_{\text{gen}}$</td>
<td>efficiency of electrical generator</td>
<td>-</td>
</tr>
<tr>
<td>$CO_{2, \text{emission}}$</td>
<td>carbon dioxide emissions, usually annual</td>
<td>tco$_2$/year</td>
</tr>
<tr>
<td>$GEF$</td>
<td>grid emission factor</td>
<td>tco$_2$/MWh</td>
</tr>
<tr>
<td>$\rho$</td>
<td>density of substance</td>
<td>kg/m$^3$</td>
</tr>
<tr>
<td>$\Delta T$</td>
<td>temperature difference</td>
<td>°C</td>
</tr>
<tr>
<td>Acronym</td>
<td>Definition</td>
<td></td>
</tr>
<tr>
<td>---------</td>
<td>------------</td>
<td></td>
</tr>
<tr>
<td>ACC</td>
<td>Air Cooled Condenser</td>
<td></td>
</tr>
<tr>
<td>CBU</td>
<td>Copperbelt University</td>
<td></td>
</tr>
<tr>
<td>CDM</td>
<td>Clean Development Mechanism</td>
<td></td>
</tr>
<tr>
<td>CER</td>
<td>Certified Emission Reductions</td>
<td></td>
</tr>
<tr>
<td>CFBC</td>
<td>Circulating Fluidized Bed Combustion</td>
<td></td>
</tr>
<tr>
<td>CO</td>
<td>Carbon Monoxide</td>
<td></td>
</tr>
<tr>
<td>CRF</td>
<td>Cobalt Recovery Furnace</td>
<td></td>
</tr>
<tr>
<td>DCS</td>
<td>Distributed Control System</td>
<td></td>
</tr>
<tr>
<td>DEH</td>
<td>Direct Electric Heater</td>
<td></td>
</tr>
<tr>
<td>EEO</td>
<td>Energy Efficiency Opportunities</td>
<td></td>
</tr>
<tr>
<td>ESP</td>
<td>Electrostatic Precipitator</td>
<td></td>
</tr>
<tr>
<td>ETS</td>
<td>Emissions Trading Scheme</td>
<td></td>
</tr>
<tr>
<td>FSF</td>
<td>Flash Smelting Furnace</td>
<td></td>
</tr>
<tr>
<td>FTSE</td>
<td>Financial Times and Stock Exchange</td>
<td></td>
</tr>
<tr>
<td>GEF</td>
<td>Grid Emission Factor</td>
<td></td>
</tr>
<tr>
<td>GHG</td>
<td>Greenhouse Gas</td>
<td></td>
</tr>
<tr>
<td>GWP</td>
<td>Global Warming Potential</td>
<td></td>
</tr>
<tr>
<td>HFO</td>
<td>Heavy Fuel Oil</td>
<td></td>
</tr>
<tr>
<td>HRSG</td>
<td>Heat Recovery Steam Generator</td>
<td></td>
</tr>
<tr>
<td>IBU</td>
<td>Integrated Business Unit</td>
<td></td>
</tr>
<tr>
<td>ID</td>
<td>Induced Draft</td>
<td></td>
</tr>
<tr>
<td>JI</td>
<td>Joint Implementation</td>
<td></td>
</tr>
<tr>
<td>IPCC</td>
<td>Inter-Governmental Panel on Climate Change</td>
<td></td>
</tr>
<tr>
<td>KBC</td>
<td>Kabundu Cathode (generic name for the cathode produced at Nchanga TLP)</td>
<td></td>
</tr>
<tr>
<td>KCM</td>
<td>Konkola Copper Mines</td>
<td></td>
</tr>
<tr>
<td>LHV</td>
<td>Lower Heating Value</td>
<td></td>
</tr>
<tr>
<td>LME</td>
<td>London Metal Exchange</td>
<td></td>
</tr>
<tr>
<td>KPI</td>
<td>Key Performance Indicator</td>
<td></td>
</tr>
<tr>
<td>kWh</td>
<td>Kilo-Watt hour</td>
<td></td>
</tr>
<tr>
<td>MES</td>
<td>Manufacturing Execution Systems</td>
<td></td>
</tr>
<tr>
<td>MVA</td>
<td>Mega Volt Ampere</td>
<td></td>
</tr>
<tr>
<td>MW</td>
<td>Mega Watt</td>
<td></td>
</tr>
<tr>
<td>Nm³</td>
<td>Normal-meter cubed</td>
<td></td>
</tr>
<tr>
<td>ORC</td>
<td>Organic Rankine Cycle</td>
<td></td>
</tr>
<tr>
<td>PDD</td>
<td>Project Design Document</td>
<td></td>
</tr>
<tr>
<td>PoA</td>
<td>Program of Activities</td>
<td></td>
</tr>
<tr>
<td>RMM</td>
<td>Relative Molecular Mass</td>
<td></td>
</tr>
<tr>
<td>Acronym</td>
<td>Description</td>
<td></td>
</tr>
<tr>
<td>---------</td>
<td>-------------</td>
<td></td>
</tr>
<tr>
<td>RTH</td>
<td>Refinery Tank House</td>
<td></td>
</tr>
<tr>
<td>SEE</td>
<td>Sustainable Energy Engineering</td>
<td></td>
</tr>
<tr>
<td>SCF</td>
<td>Slag Cleaning Furnace</td>
<td></td>
</tr>
<tr>
<td>SIDA</td>
<td>Swedish International Development Agency</td>
<td></td>
</tr>
<tr>
<td>SX</td>
<td>Solvent Extraction</td>
<td></td>
</tr>
<tr>
<td>$tco_{2e}$</td>
<td>tonnes of Carbon dioxide equivalent</td>
<td></td>
</tr>
<tr>
<td>TLP</td>
<td>Tailings Leach Plant</td>
<td></td>
</tr>
<tr>
<td>tpa</td>
<td>Tonnes per annum</td>
<td></td>
</tr>
<tr>
<td>tpd</td>
<td>tonnes per day</td>
<td></td>
</tr>
<tr>
<td>TPH</td>
<td>Tonnes per hour</td>
<td></td>
</tr>
<tr>
<td>UNFCCC</td>
<td>United Nations Framework Convention on Climate Change</td>
<td></td>
</tr>
<tr>
<td>UK</td>
<td>United Kingdom</td>
<td></td>
</tr>
<tr>
<td>UNZA</td>
<td>University of Zambia</td>
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</tr>
<tr>
<td>WHRB</td>
<td>Waste Heat Recovery Boiler</td>
<td></td>
</tr>
<tr>
<td>WPGC</td>
<td>Worley Parsons Gas Cleaning</td>
<td></td>
</tr>
</tbody>
</table>
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This report should not be interpreted as representing the views of those that contributed to its successful completion. The views expressed in this report are those of the author and do not represent the views of KCM. All errors, inaccuracies in the data collected and analysis remain the responsibility of the author, all data herein presented was chiefly for analytical purposes wherein the underlying principles were of the essence.

Wilson Miti
January 2014
ABSTRACT

Copper processing plants involved in smelting, electro-refining and electro-winning are heat-intensive undertakings that provide extensive challenges for attainment of high energy efficiency. Literature has shown that most of these plants, especially smelters, operate at low overall energy efficiency due to the seemingly complex energy scenario where heat and electricity as forms of energy are treated distinctly from each other. Many copper processing plants have not yet explored both available and emerging waste heat recovery technologies hence remain operating at lower energy efficiencies.

In the copper processing plants under study in particular the Nchanga tailings leach plant (TLP), plant operators hinted that some of the processes that ought to operate in heated environments operate at ambient temperatures because of lack of a heating mechanism. The project discusses possible heating mechanisms from available local resources and applicable technologies. As the competing options for providing the required heat at the Nchanga TLP present different carbon emission scenarios, the carbon emissions associated to the recommended installations shall be quantified against a suitable baseline. Flue gas waste heat from the nearby Nchanga smelter has been taken as the available local energy source on which the applicable heating scenarios at TLP are analyzed.

The project analyzed waste heat scenarios for three furnaces at Nchanga smelter where it has been established that flue gases from the furnaces contain 37.31 MW of waste heat. Analysis for channeling the waste heat into heat recovery steam generators gave the steam turbine power generation potential of 7.06 MW. The project also demonstrated how energy efficiency undertakings can be used as a driver for carbon emission reduction measures and for participation to the available carbon trading mechanisms such as CDM. Selection of suitable baseline scenarios revealed a lot of potential for carbon finance undertakings in the three case study plants. At the Nchanga smelter, the 7.06 MW power generation capacity has an associated potential of 61,820 tCO₂/year emission reductions that can be monetized through the available carbon trading markets. The research established that Nchanga TLP has a heating demand of 10.87MW. If this heating demand was to be met by using the smelter waste heat, the undertaking can be taken as CDM activity or other carbon trading platform with an associated potential of 95,183 tCO₂/year.
CHAPTER 1
INTRODUCTION

This chapter presents the scope of work that is to be undertaken in the research work. It lays out the existing energy scenarios in the case of three of Konkola Copper Mines (KCM)'s main copper processing plants. It further brings into view those areas that this research identified as opportunities for energy efficiency improvement and carbon emission reductions to align them with applicable emerging technologies for possible deployment.

1.1 Background

Konkola Copper Mines (KCM) Plc is Zambia’s second largest mining company after Lumwana copper mine. The company is owned (via majority shareholding) by Vedanta Resources, a globally diversified natural resources group with wide-ranging interests in aluminum, copper, zinc, lead, silver, iron ore, oil, gas and energy. Vedanta owns and operates units in Zambia, Namibia, South Africa, Liberia, India, Ireland and Australia [1].

In Zambia, KCM operates three main mines comprising Konkola, Nchanga underground and open pits, and Nampundwe. The company mines copper at both Nchanga and Konkola mine sites while at Nampundwe pyrite is mined. The company also runs mineral processing plants that include a smelter at Nchanga; concentrator plant at Nchanga; a tailings leach plant at Nchanga and an electro refinery at Nkana. All the areas of operation are situated on the Copperbelt Province of Zambia, except for Nampundwe mine which is located near Lusaka [2].

The processing plants, i.e. the Nchanga smelter, TLP and Nkana refinery were taken as the three case studies in this thesis research project. The research project undertook to investigate energy scenarios in these three mineral processing units of KCM discussed in the ensuing sections. For each possible installation recommended, the associated carbon emissions are discussed and quantified.

1.1.1 Nchanga smelter

The Nchanga smelter has a production capacity of 311 tons per annum and is part of the Nchanga integrated business unit (IBU) that also comprises the TLP, concentrator plant, Nchanga open pits and underground mines. The smelter plant was commissioned in September 2008 and replaced the old Nkana smelter on account of the latter’s heavy reliance on fossil fuels and high energy inefficiency [3]. The smelter was commissioned out of the company’s desire to contribute towards environmental protection by reducing
its carbon footprint. Figure 1.1 depicts a schematic outlook of KCM’s mining and processing operations at Nchanga.

**Fig 1.1:** Outlook of KCM’s Nchanga site operations. Processing of copper ore from both open-pits and underground mines begins at the primary crusher, the ore is then fed into the concentrator where separation of acid-soluble and insoluble copper occurs for input into TLP and the Nchanga smelter respectively.

The Nchanga smelter uses the Outokumpu flash smelting technology, a process used for sulfur-containing ores [4]. This copper smelting technology was developed by Outokumpu in Finland, and Outotec, formerly the technology division of Outokumpu now holds patents to the technology and licenses it worldwide.
The Outokumpu process employs the autogenic principle by using the energy contained in the sulfur and iron for melting the ore [5]. The technology is acclaimed for high energy efficiency as heat required for smelting is derived from the exothermic reactions and latent heat contained in the concentrates. The Outokumpu direct-to-blister technology is a one-step copper smelting process in which the metal (copper blister) is directly produced from mineral concentrate [6].

Combustion of sulfur and iron is exothermic and thus produces enough heat to sustain the reactions. This reduces demand and dependence on fossil fuels and hence minimizes on energy-related greenhouse gas emissions. Advocates of this technology claim that it is the cleanest method available and the most feasible way of producing copper and nickel [7]. The largest quantity of the needed heat comes from the feed material although minimum burning of fossil fuel is required to compensate for the losses through the furnace walls.

1.1.1.1 Flash smelting furnace

The major waste heat generation source is the flash smelting furnace (FSF). Under normal operation, the FSF releases $45,800 \text{m}^3/\text{h}$ of flue gases at temperatures of up to $1450^\circ\text{C}$. The FSF channels its flue gases into a waste-heat recovery boiler (WHRB) where steam is generated at $277^\circ\text{C}$, 60 bar, and mass flow of 60 tons per hour. The gases contain, among other constituents, 41% sulfur dioxide which is channeled to an acid plant where it is used for production of sulfuric acid. The acid production eliminates sulfur emissions into the atmosphere, giving the technology an environmental merit. The installed production capacity of the sulphuric acid plant is 1850 tons per day. [8]

Only a fraction of the steam generated in the WHRB goes to productive usage for drying of raw copper concentrate in the multi-coil steam dryer and a recently installed heat exchanger for heating of process nitrogen. The rest goes to an air cooled condenser (ACC) unit for cooling, where the resulting condensate is passed through a heat exchanger for further cooling and later pumped back to the WHRB for another cycle of heat absorption from the flue gases. The condensation of the excess steam at the ACC is achieved by use of eighteen electric fans. Although most of the steam is condensed back to water this way, at many instances part of the steam is seen vented out into the atmosphere through the dearator. It can be noted that an externally-sourced form of energy through the air cooled condenser is expended to dispose of the waste heat whose exergy is evidently high.

1.1.1.2 Other furnaces and heat points

Apart from the flash smelting furnace, the other sources of flue gases whose heat content can be considered for extraction include; slag cleaning furnace (SCF), cobalt
recovery furnace (CRF), acid plant and anode furnaces. All these are associated with them cooling towers for gas cooling.

1.1.2 Distribution of steam from the WHRB

Currently, the steam generated in the WHRB is distributed as follows:

1.1.2.1 Steam dryer

Two dryers are installed for drying of wet copper concentrate before it is conveyed into the flash smelting furnace. At full operation capacity of 80 tons per hour of concentrate feed, each dryer requires steam mass flow of 15 tons per hour. Steam feed to the dryer is regulated depending on the concentrate feed rate and moisture content. Pressure is used as the control parameter done on a distributed control system (DCS) in the control room.

1.1.2.2 Steam heat exchanger nitrogen heating

Recognizing the high availability of waste heat in the generated steam, installation of a steam heat exchanger for nitrogen heating was undertaken to replace an electric heater. The Electric heater had power rating of 1.35MW, used to heat nitrogen from 14°C to 240°C, operating for 6 hours in four cycles a day. The design steam mass flow through the nitrogen heat exchanger is 3 tph. [9]

1.1.2.3 Air cooled condenser

The air cooled condenser (ACC) was installed for purposes of cooling down and condensing the portion of steam from the WHRB that is not utilized. The daily energy consumption of this unit is around 4500 kW.hr [10].

1.1.3 Nkana Refinery

The Nkana Refinery utilizes direct electric heaters (DEH) for heating and maintaining electrolyte temperatures at optimum levels. Heating of the DEH is achieved by using grid electricity. The refinery plant has nine DEH circuits, each with power rating of 2.5 MW. These are required to maintain electrolyte temperatures in the range 63±5 °C. Each circuit has a standby heat exchanger that is run during downtime of the DEH. Furthermore, the plant has a 10 ton/hr electrode boiler on standby, also used for or for boosting of electrolyte temperatures whenever the DEH cannot sustain the required temperature levels. Other installations requiring heat at Nkana refinery are heating of de-mineralized water in two large water tanks containing submersible electric heaters. The heated water is used for making of the re-agents (glue and ligno sulphate) [11].
1.1.4 Nchanga tailings leach plant

The TLP treats acid-soluble copper. Acid-insoluble copper is treated at the Nchanga smelter. The acid-soluble copper is transmitted through pipes from the Nchanga concentrator in the form of slurry. This acid-soluble copper slurry also contains fractions of acid-insoluble copper that remained as impurity from the extraction process at the concentrator [12]. A guided tour of Nchanga TLP yielded operational parameters and other information with regard the heating demand as in the ensuing paragraphs [13].

At TLP, the slurry is fed into twelve fluidized bed reactors (cylindrically shaped, conical at the bottom) each with 833 m$^3$ effective volume. The fluidization effect is achieved by injecting a jet of compressed air at the bottom of the reactor at 3.1 to 4.0 bar pressure which bubbles-up the reactor, in the process keeping the reactants in suspended form. For this purpose, three compressors are installed, each producing 17,000 m$^3$/hr of compressed air, and a maximum of two are run at each given time.

Originally and currently, the temperature of the compressed air being fed into the reactor is at ambient levels. However, due to the presence of acid-insoluble copper, and also to improve recoveries of soluble copper, the company desires that the compressed air be heated to optimum reaction temperatures of 50 to 60 °C. The task is to devise a mechanism for heating-up the compressed air at minimum cost, possibly from available waste heat from the nearby Nchanga smelter.

Furthermore, in the refinery tank house at TLP are four dip tanks containing water for dip-washing of the copper cathodes as they are pulled from the cells upon maturity. Each of these tanks has four heating elements. The electric power rating of each individual heating element is 30kW, implying total installed heating capacity of 480kW for all the heater. The function of these heaters is to maintain water temperature in the dip tanks at 80 °C.

The temperature of the electrolyte in the refinery tank house is currently at ambient levels (30°C). TLP management desires to heat-up this electrolyte to 50°C to improve the dispatchability (percentage of good quality cathode) of the plant and reduce on reject cathodes.

To improve and maintain current efficiency at desired levels in the refinery tank house, all cell contacts are frequently washed using pressurized hosed water (to achieve good contact for flow of electricity thereby reducing losses through heat), this water is currently at ambient temperatures. TLP management further desires that even the water for hosing the cell contacts should be at higher temperature to effectively wash out any accumulating material deposits. Both TLP and Nchanga smelter management are considering the possibility of utilizing waste heat from smelter furnaces and the excess steam from the WHRB to meet the heating demand at TLP.
1.2 Problem Statement

From the foregoing, it can be noted that the KCM mineral processing plants studied under this thesis project are heat-intensive, and in the case of Nchanga smelter, potential does exist for improving overall energy efficiency that can have a positive spill-over effect to the nearby TLP, as well as in reducing energy-related greenhouse gas emissions. Some processes that otherwise require elevated temperatures are running at ambient temperatures thereby leading to poor product quality, as is the case at TLP. Utilization of grid electricity to meet heating demand of a heat-intensive installation as is the case at Nkana refinery can be argued not to be the best option when emerging technologies of heat and power, and other available energy and fuel sources such as biomass can be considered.

1.2 Aim of this work

The aim of this project is:

to take a holistic view to both heat and power scenarios and formulate applicable technological measures towards improving overall energy efficiency and reducing energy-related carbon emissions in the three copper processing plants as case study. In endeavoring to achieve this aim, the project will undertake to do the following:

(i) Formulate applicable measures that would improve the overall energy efficiency by taking a holistic view and beyond-boundary demand-side needs of both heat and electricity
(ii) Provide suggestions of possible measures that can be implemented to provide the required heat in sections that are running at ambient temperatures to improve product quality
(iii) Present possible technologies and measures that could be implemented to reduce dependence on grid electricity
(iv) Present strategies where energy and carbon management plans can be formulated in the three plants under study
(v) Discuss the linkage between energy and carbon management and the duo benefit of integrating strategies aimed at improving energy efficiency and reducing carbon emissions
(vi) Discuss and quantify greenhouse gas emissions associated with the several options of energy efficiency improvement

1.3 Methodology

Data will be collected from the company employees in operations and from management. Energy consumption data and other operations parameters will be
collected by review of daily consumption reports in specific sections in each of the three case study plants.

At the outset, analysis of waste heat scenario at the Nchanga smelter will be conducted. Designing of a system to meet the heat demand at the TPL will be anchored on available heat from the nearby Nchanga smelter. In order to design an integrated system for waste heat utilization, both demand side analysis in the TLP as well as steam generation capacity at the Nchanga smelter shall be discussed.

The analysis shall be conducted using fundamental theories of heat and power technology. The carbon emission scenarios associated with all the proposed technologies will be quantified using standard emission factors.

1.4 Significance of the study

In view of the scourge of global warming and climate change, corporate organizations are being compelled to undertake measures that will reduce their energy consumption and hence carbon footprint. This thesis project will contribute to the global agenda on climate change mitigation by formulating methods for possible implementation that reduce energy and carbon footprint for the three mineral processing plants under study and other similar mineral processing plants in general. All analyses of energy scenarios shall also provide quantitative analysis of the greenhouse gas emission benefits associated with the implementation of each undertaking.

1.5 Motivation to undertake the thesis subject

The interest to pursue the topic of energy efficiency and carbon removal strategies in mineral processing plants emanates from personal encounter with the heat-intensive installations analyzed in the project. From observation, efforts aimed at improving energy efficiency and savings are in most cases based on corrective actions to plant operations and switching off appliances when not in use, which are the obvious cases.

In general, industry operators seem to have overlooked the fact that to be able to adequately formulate tangible energy efficiency measures, a system ought to be viewed in its entirety, accounting for all available energy forms and losses. In the smelter plant, not much seems to have been done to incorporate heat in the equation to evaluate overall energy efficiency. Furthermore, disposal of waste heat is being done at a cost to the company yet modern technology provides leverage to harness the heat and re-channel it back into the system for productive utilization. It is these issues that motivated the current subject as project for this Master of Science dissertation.
CHAPTER 2

LITERATURE REVIEW

There exists a close relationship between energy efficiency and greenhouse gas emissions which leads to integration of the concepts of energy and carbon management. Energy efficiency upgrade undertakings have been severally documented as key strategy for greenhouse gas emission reduction in industry [14]. A survey of existing and emerging energy efficiency technology was conducted for this thesis project to recommend and justify those that would be suitable for implementation in the scenarios under study.

2.1 Energy management and carbon strategies in copper processing installations

Mineral processing is an energy-intensive industry whose energy demand is met from a diverse range of sources that include combustion of fuels and grid electrical energy. In some cases, the grid electricity is generated from fossil-fuel powered power plants. This industry has struggled to attain high levels of energy efficiency at various stages of the production chain to meet both economic and environmental obligations, the later arising from issues of global warming and climate change [15].

To attain significant levels of energy-efficiency improvement requires an integrated approach in a given environment. Various areas of operation offer different opportunities for energy efficiency improvement owing to the diverse nature of operations in mineral processing [16].

Owing to economic and environmental reasons, the copper processing industry faces pressure to reduce its energy consumption and these may become fundamentals that shall drive technological change in the mineral processing sector. These fundamentals represent significant challenges to the industry and create an environment supportive of technological innovation and business sustainability. Energy efficiency upgrade as key to meeting both economic targets as well as environmental sustainability has its own challenges. Accurate prediction of energy use remains a challenge for mining and mineral processing companies and this always leads to penalties from utility companies arising from the deviations of over or under forecasting. An important strategy for effective energy efficiency improvement undertakings would be to couple the relevant performance monitoring data to energy consumption data. Here literature cites lack of data availability on energy use as one of the major issues in formulating energy efficiency improvement measures, and in cases where data is available, it is hardly linked to production data [17].
Overall, literature has shown that research has been on-going with the aim of formulating ways of lowering the mineral processing industry’s energy consumption as a contribution towards climate change mitigation [18]. In the pre-global warming and climate change era, the desire to increase energy efficiency was associated with strategies to reduce production costs, however in the current scenario, energy efficiency reduction undertakings are also associated with the desire to participate in climate change mitigation as well as meet emerging government policies [19].

The existing and emerging carbon markets are providing further zeal for the mining and mineral processing companies to reduce their per-unit-of-production energy use to reduce their greenhouse gas emissions in return for revenue from carbon finance. Evidently, to achieve the desired results requires clear understanding of prevailing energy scenarios in a given environment, equipped with enough tools for decision making, as well as integrity of carbon management measures acceptable to the strict rules and standards of the carbon markets.

The news media report that governments across the world are now recognizing the need for energy efficiency, and emission reduction undertakings in energy-intensive installations, particularly the minerals industry. It has been reported that national governments are now encouraging companies to undertake carbon management as part of their integrated business strategies and to disclose their GHG emissions data alongside their financial reports [20]. The rationale behind is that many governments believe that increased uptake of cost-effective energy efficiency and renewable energy technologies is the right way to take towards a low carbon economy amidst the scourge of global warming and climate change.

### 2.2 Carbon management

Carbon management is a measurement process which promotes environmental effectiveness in organizations with the aim of reducing greenhouse gas emissions. It is designed to maximize efficiencies in the consumption of materials and resources with the aim of reducing greenhouse gas emissions to cushion global warming and climate change [21]. Organizational performance in terms of greenhouse gas emissions is usually measured in the form of a carbon footprint.

Carbon footprint refers to the total set of greenhouse gas emissions caused by an organization, event, product or person [22]. It is a measure of the total amount of carbon dioxide (CO$_2$), methane (CH$_4$) and other greenhouse gas emissions of a defined population, system or activity, considering all relevant sources, sinks and storage within the spatial and temporal boundary of the population, system or activity of interest [23]. All measurements are calculated as carbon dioxide equivalent (CO$_2$e) using the relevant 100-year global warming potential (GPW100). The GPW100 is a comparison of the amount of heat trapped by a given greenhouse gas to the amount of heat trapped by a similar mass of carbon dioxide over a period of hundred years. The choice of which
period to consider for GWP depends on how the gas concentration decays in the atmosphere over time. Other periods of time may be taken, the indicative 100 year period is usually the one considered by regulatory bodies such as the United Nations framework convention on climate change (UNFCCC) and California air resources board [24].

An organization’s carbon footprint can be measured by undertaking a GHG emissions assessment or other calculative activities denoted as carbon accounting. It is only once the size of the carbon footprint is known that strategies can be devised to reduce it such as by technological developments, better process and product management, carbon capture, and many others. The desire for corporate entities is to reduce their carbon footprints to competitive levels with other players in that particular industry [25]. Copper processing plants are perceived to have higher carbon footprints emanating from a direct linkage between energy usage and carbon emissions and their energy-intensive nature. It can therefore be seen that owing to the close relation between energy intensity and carbon emissions, players have found it imperative to manage the two in an integrated manner to achieve the desired results.

In order to promote participation to carbon emission reduction efforts at a global scale, market mechanisms that provide platforms of converting achieved carbon emission reductions by a project proponent into economic value have been devised. A number of carbon trading mechanisms have been formulated to provide exchange of emission reductions for economic returns as well as to meet mandatory obligations. The trading platforms include the voluntary markets and the regulated markets. Prominent among the regulated carbon trading platforms are those under the UNFCCC’s Kyoto Protocol. Under this platform are three mechanisms: Joint Implementation (JI); Emissions Trading Scheme (ETS); and Clean Development Mechanism (CDM). The first two are applicable to industrialized countries while the CDM allows less-industrialized countries to participate [26].

The unit of measure on the carbon market for sells of carbon credits, also called verified carbon units (VCUs) or certified emission reductions (CERs), is the tonne (tonne of carbon emissions equivalent):

\[ 1 \text{ VCU or CER} = 1 \text{ tonne of CO}_2 \text{ emissions or CO}_2 \text{ equivalent}. \]

### 2.3 Carbon management in the global village

At Rio +20, then British Deputy Prime Minister made a surprise announcement that effective 2013, all publicly listed companies would have to report on their carbon emissions. The British government further announced that it would consult on new regulations in 2016 to decide whether to implement the reporting requirements to all large companies in the UK. The pronouncement when realized makes the UK the first country in the world to require companies to include carbon data in their annual reports.
The legislation compels companies to report on their full organizational footprint [27]. Chadwick [Carbon Clear, 2012] stated that it is only once companies started to measure their carbon impact that they would realize the potential for cost savings and start to take real action to tackle their emissions. He further states that this would prevent companies from hiding poor carbon performance behind non-disclosure of their data.

The FTSE100 publication for 2012, based on selected criteria on carbon emissions performance for the listed companies, ranked the supermarket sector as the best performer, with the manufacturing, mining and metals and building materials sectors coming bottom of the league table. The criteria was based on measurement, reporting and verification of companies’ carbon footprints; their existing and planned strategies for reducing emissions; their actual carbon reductions and their work to engage stakeholders about their climate change programs [28]. This compels the mining and minerals' processing sector to formulate strategies for energy and carbon management if they are to improve their performance outlook.

2.4 Waste heat scenarios in copper smelters

2.4.1 Smelter flue gas heat recovery

Waste heat is heat generated in a process by way of fuel combustion or chemical reaction, which is then ‘dumped’ into the environment and not reused for useful and economic purposes. The essence is not the amount of heat that matters in the waste, but rather its value. The decision whether to recover waste heat in a given scenario or not, depends on the temperature of the flue gases and the economics involved [29].

Several literature on smelting energy scenarios and an observation of the Nchanga smelter reveal that copper smelters have a number of furnaces installed within a given plant to meet the several stages of the process. In all cases, these furnaces generate high flue gas flows at high temperatures and are of varying compositions. In these plants, flue gas heat recovery is cited as being critical to minimizing energy consumption and reducing operating costs [30]. Researchers are still endeavoring to develop technologies to improve heat recovery as well as meet the need to reduce energy-related greenhouse gas emissions.

In flash smelting furnaces, the process continuously generate flue gases at 1000-1200°C temperature and these are channeled into the waste heat recovery boiler which is typically close-coupled ahead with an electrostatic precipitator (ESP). The ESP consists of surfaces that trap the dust load by way of electrostatics, and the dust is collected at the bottom to where it drops by gravity. The Boiler cools down the off-gas to below 400°C before ejecting the gas into the ESP. Despite the thermal energy removal by the boiler, the gas still retains a lot of heat at ESP exit [31].
Flash furnace off-gas (flue gas) at 40,000 m\textsuperscript{3}/h can contain 5 to 6 MW of sensible heat at the ESP outlet and is relatively clean of dust. In installations containing systems for capture of sulphur, the off-gas is channeled to an acid plant after gas cleaning and cooling. The remaining heat after the ESP is expended in the wet gas cleaning plant, which has an attached operating cost to meet the cooling need. Downstream heat removal is conducted through cooling towers utilizing water. Flue gases from converters at 30,000 m\textsuperscript{3}/h blowing rate can contain 8-12 MW of sensible heat at the ESP outlet per blowing converter and is also relatively clean of dust. [32]. If hot ESP outlet gas could be further cooled through a heat recovery unit, additional heat recovery could be achieved, and expenses on water cooling tower would be reduced. The gas is cooled by water-spraying or radiatively to below 400\degree C for ESP; this could be replaced by a cooling unit with heat recovery vessel.

A typical copper smelter consists of electric furnaces, waste heat recovery units, and recently, sulphuric acid plant for conversion of sulphur dioxide to acid. The ensuing subsections briefly discuss the furnaces as the major sources of the waste heat and the boiler as recipient of the highest volume of hot flue gases. The quantitative figures of temperature and heat given were as obtained from process metallurgists at Nchanga smelter from monitoring of control and design parameters.

### 2.4.2 Electric furnaces

Electric furnaces are installed for purposes of slag-cleaning in sulfide smelting. They operate continuously with gas temperatures of 500 to 1000\degree C. The smelting off-gas is typically cooled by spray cooling and then mixed with secondary ventilation to a baghouse. Smelting furnace off-gas typically contains 12-15 MW of sensible heat before spray cooling. Approximately 6-18 MW of sensible heat could be recovered by a heat recovery unit instead of spray cooling. Alternatively, process gas could be sent directly to a dryer or kiln for full heat recovery. Electric smelting furnace off-gas at 310,000 m\textsuperscript{3}/h contains 19-24 MW of sensible heat at the baghouse outlet and is clean of dust. If baghouse outlet gas could be sent through a heat recovery unit, significant heat recovery could be achieved [33].

### 2.4.3 Anode furnaces

These are installed in sulfide smelters to refine blister copper for anode casting. Anode furnaces operate in batch process with each batch or cycle lasting about 3 hours or more. Most modern smelters have two or three anode furnaces, with two operating at a time. The Nchanga smelter complex has three anode furnaces.

The process gas flow rate generally is less than 10,000 m\textsuperscript{3}/h at temperatures of 1100 to 1200\degree C. During reduction, the gas can contain up to 25% combustibles which are burned. The gas is then cooled by spray cooling to 200\degree C and later channeled to a
baghouse which traps the dust component in the gas stream. The off-gas can still contain about 4.5 MW of sensible heat at the baghouse exit [34].

2.4.4 Waste heat recovery boilers

Waste heat recovery boilers (WHRB) are installed to receive hot flue gases emergent from the furnaces. They handle continuously flowing hot, contaminated, and potentially corrosive gas streams. A typical WHRB usually consists of radiative and convective sections as depicted in figure 2.1. The radiative section cools gas temperature down to 700°C and the convective section further cools the gas to 350-400°C. The boiler can be designed to produce saturated or superheated steam. The boiler is highly subject to impacts from thermal cycling and corrosion. WHRBs are said not to be effective at temperatures below 350°C due to temperature requirements of boiling at high pressure [35].

![Diagram of WHRB](image)

**Fig. 2.1:** Direction of flow of hot flue gas across the WHRB from the FSF, also showing placement of the radiation and convection sections. The gas first enters the radiation section of the boiler and exists through the convection section on its way to the ESP.

2.5 Other waste heat recovery equipment

Besides the WHRB which receives the largest volume of flue gas streams, several other waste heat recovery equipment currently exist and research is on-going to bring-in new technologies to cater for varying scenarios. This section discusses some of the existing and the emerging waste heat recovery technologies as covered in the surveyed literature [36].
2.5.1 Air-To-Gas Exchangers (Recuperators)

Recuperators are shell-and-tube type heat exchangers. They, like boilers, handle streams of dust-laden gases and are relatively cheaper to install. In these, heat is recovered in the form of preheated air suitable for use in dryers, kilns, burners, etc. The heat recovered is of low grade (less than 400°C). The drawback is that they have limited other uses of the recovered heat. The units are subject to impacts from thermal cycling and corrosion and are not effective at lower temperatures because of air content.

2.5.2 Organic Rankine Cycle (ORC) Heat Recovery and Power Generation

These are similar to boilers producing steam from water but use low boiling point fluid as heat transfer medium operating in closed circuit processes.

ORC is a well developed technology for low grade heat recovery in other applications. It works well at inlet temperatures above 175°C – suitable for installing downstream of baghouses or ESPs. Installing downstream of baghouses and ESPs reduces risk of fouling and plugging, and minimizes thermal cycling. ORC units can be installed downstream of boilers or air-to-gas heat exchangers for increased heat recovery. Figure 2.2 depicts a schematic of an ORC coupled to a steam cycle power plant while figure 2.3 shows an indirect heat exchange ORC with thermal oil as an intermediary heat exchange medium. In figure 22, the hot flue gas stream exchanges its heat with feed water/steam to a steam turbine through a heat exchanger.

In figure 3.3, the heat recovered from the flue gases is used for power generation through a steam turbine. However, in this case there is no direct contact between the water/steam side with the flue gas. The system has an intermediate heat exchanger consisting of thermal fluid which captures the heat from the flue gas and exchanges it with the steam cycle.

2.5.3 Thermal Oil Heat Exchangers

This closed-loop system is similar to waste heat boilers, but uses liquid oil as heat transfer medium. The oil can be heated up to 350°C. These installations are well suited to preheat air or other fluid, using two heat exchangers; and can be used as intermediate medium in an indirect ORC system. Key factors in selecting the oil include cost, expected operating range, quantities required, and availability.

Thermal oil heat exchangers could be used to replace spray coolers or radiative coolers in high temperature applications in conjunction with ORC. ORC units can produce enough electricity to run induced draft (ID) fans and additional equipment. Net energy conversion including parasitic loads is estimated at 10-20% of thermal energy recovered, or more for higher temperature applications.
Fig. 2.2: Organic Rankine Cycle – direct heat exchange.

Fig. 2.3: Organic Rankine Cycle Indirect heat exchange
Figure 2.4 depicts a thermal oil heat recovery system. In the figure, notice that thermal oil acts as an intermediate heat transfer medium. The first heat exchange is between the flue gas and the thermal oil through a heat exchanger. The second heat exchange is between the thermal oil and the air which captures the heat to final use. The drawback for this technology is that the system can be expensive, depending on the type of oil selected, and the need for a secondary heat exchanger to preheat air or other fluid.

2.5.4 Thermal Storage and Power Generation

In this technology (also close-loop type), heat is stored in the thermal fluid to generate steam or boil organic fluid for power generation or to preheat air for other processes. It is an economical heat recovery option in the batch processes. Figure 2.5 depicts a constant stream of thermal fluid drawn to produce power or preheat air.

**Fig. 2.4:** Emerging technologies: waste heat recovery techniques applicable in smelters. The figure depicts a typical thermal oil heat recovery system with air as the heated fluid. Air can be replaced with other suitable medium (e.g. steam cycle) if using the recovered heat for power generation.
When used for power generation, a turbine can be operated steadily without large variability in heat load.

**Fig. 2.5:** Emerging heat recovery technologies applications in smelters, the figure shows a typical thermal storage heat recovery system. Thermal fluid is used as the intermediate heat transfer medium. The second heat exchange is between the thermal fluid and the fluid taking the heat for selected use.

### 2.6 Benefits of process flue gas heat recovery

Design and adaptation of heat recovery and energy optimization practices in mineral processing can provide advancements in reducing energy consumption, recovering the otherwise wasted energy and reducing energy-related greenhouse gas emissions. Other benefits of flue gas heat recovery installations include [37]: reduction in energy and associated operating costs, increased production where there is higher dependence on energy availability, reduction in GHG emissions arising from reduced
consumption of fuels, and reduction in CO₂ emissions which can generate revenue from sell of carbon credits.

2.7 Challenges and solutions in recovering waste heat

The major challenge that needs to be addressed in waste heat recovery from smelter flue gases is the large dust load carried by the gas streams. This leads to high probabilities of fouling and plugging. Fouling and plugging reduce heat transfer, increase pressure loss, and can reduce fume capture efficiency at the source. Waste heat recovery units must be designed to allow streamlined gas flow through the unit to avoid dropout or impingement onto the heat transfer surfaces. It is often necessary to incorporate rappers or sonic horns to periodically knock accumulated dust off the heat transfer surfaces. In sulfide smelting, it is preferable to add sulfatizing air to the flue gas in order to oxidize sulfide dust to surfaces to reduce sticking to surfaces [38].

Corrosion is another challenge in waste heat recovery units. It usually arises when gas streams are cooled down to below dew points where water or acid condenses on the surfaces. To avoid or minimize corrosion, it is important to understand the dew points of the gas constituents and also the temperature profile so that optimization can be done to avoid operating closer to the dew point. Furthermore, thermal cycling can cause material fatigue, cracking and leakage from the expansion and contraction of the equipment. Heat recovery units should be designed with appropriate expansion allowance and operators should undertake regular maintenance to safeguard the material integrity of the equipment [39].

2.8 Example of waste heat recovery in a plant similar to the smelter case under study

2.8.1 The Kennecott Utah Copper Smelter

Kennecott Utah copper smelter is a subsidiary of Rio Tinto, located at Salt Lake County in Utah. The ensuing paragraphs give account of the features at Kennecott applicable to KCM Nchanga smelter [40].

Like the KCM Nchanga smelter, Kennecott employs the Outokumpu technology, processing copper concentrate through heat and oxidation in furnaces into 99.5% copper anodes which are later processed into copper cathodes ready for the market. The smelter production capacity is 300,000 tons of copper anodes per year. It is said that Kennecott smelter has the highest level of energy recovery of any smelter in the world. Based on forecasts of rising energy costs, Kennecott determined as a long-term investment to generate power from thermal energy that would otherwise be wasted. The custom-designed waste heat recovery system takes heat from several sources in the
smelting process and uses it to generate steam, which further runs a steam turbine unit for power generation.

Exhaust gases at 277°C from the flash smelting and flash converting furnaces pass through waste heat boilers that remove heat from the gases to generate steam. This steam is then superheated to 400°C and later merged with more steam from two other gas-fired boilers to drive the acid plant compressors. The compressors pull the emissions through the pollution control equipment upstream of the compressors and then push the gases through the downstream acid plant. Heat is captured from the exothermic conversion of sulfur to sulfuric acid to produce yet more steam. The steam generated in the acid plant is then superheated and combined with the low-pressure steam discharged from the acid plant compressors. The resulting steam drives a low-pressure steam turbine generator set to produce approximately 20-25 MW of electricity—about two-thirds of the smelter’s electricity needs and about 10 percent of Kennecott Utah copper’s total electrical need in 2011.

The size of the power generation system was determined based on the amount of waste heat available. Therefore, the amount of electricity generated depends on the amount of steam at hand, which depends on how much copper anode is being processed, which goes back to mine and concentrator production. Since the smelter runs 24/7, the waste heat recovery system also runs 24/7 at relatively constant rate. The logical philosophy behind waste heat recovery for power generation at Kennecott is that, “If you have waste heat which you can use to generate electricity, then go forward with the proper analysis. It can potentially be a great energy efficiency project with positive economic returns and environmental benefits”, and “When designing the system, focus upfront on getting the high level of reliability required by your plant”

2.8 other examples of waste heat recovery

2.8.1 Sterlite Industries India Limited (Siil), Tuticorin

Sterlite Industries, a subsidiary of KCM’s parent company Vedanta Resources, have an 11.2 MW steam turbine power plant that utilizes steam generated from a WHRB attached to the Isa smelt furnace. The WHRB was designed to recover waste heat from the hot gases emerging from the Isa Smelt furnace at temperature of 1040°C. This waste heat recovery project was taken as and is a registered CDM project owing to reduction of the smelter’s carbon footprint from the project [41].

2.8.2 Anhui Huai Bei Mining (Group) Cement Co. Limited, China

Another example of waste-heat recovery for power generation is the Anhui Huai Bei Mining (Group) cement company Limited in China. This is yet another example of such
undertakings that have been taken as CDM project, in this case, from a Cement plant [42].

2.9 Integrated efforts aimed at energy efficiency improvement and carbon management in copper processing

To integrate the sector efforts, the copper technology roadmap of November 2004 agreed on a quantitative goal of achieving 10 per cent improvement in energy efficiency and carbon footprint reduction through implementation of improved technology. This goal highlighted the importance of energy to copper mining and production, indicating a shared opportunity to lower costs and conserve energy. The industry goals agreed-on from the technology roadmap were to achieve the following [43]:

- Lower the cost of production
- Achieve the balance of acceptable economic, environmental and social effects
- Manage technological risk and investment
- Improve safety, health and industry hygiene
- Achieve a 10% improvement in energy efficiency and the associated greenhouse gas emissions through the implementation of improved technology
CHAPTER 3

NCHANGA SMELTER HEAT AND POWER ANALYSIS

In this chapter, we endeavor to capture current energy scenarios in the installations under study at Nchanga smelter. Quantification of available waste heat through thermodynamic analyses shall be done. Finally, the carbon emission reductions associated with the discussed potential installations of waste heat recovery units shall be quantified.

3.1 The Nchanga smelter complex

The Nchanga smelter processes raw copper into anode blister copper which goes for further refining into finished copper cathodes at the Nkana refinery and Nchanga TLP. The smelter processes a blend of copper concentrates of varying grades and composition from different sources that include the Nchanga open pits, Nchanga underground, Konkola mine, Nampundwe mine (pyrite), Lumwana mine, Kansanshi mine and Frontier mine. Figure 1.1 showed where Nchanga smelter sits in the neighborhood of other operations at the Nchanga IBU. Figure 3.1 depicts the smelter’s process flow and the main installations which include the flash smelting furnace (FSF), three anode furnaces, two cobalt recovery furnaces, slag cleaning furnace, anode casting wheel, waste heat recovery boiler, twin steam dryers, ESP, and the sulphuric acid plant.

3.2 Nchanga smelter waste heat scenarios

To be able to design waste heat recovery systems will require knowledge of the amount of heat available. The main sources of waste heat in the Nchanga smelter are: flash smelting furnace, slag cleaning furnace, two cobalt recovery furnaces, three anode furnaces and the acid plant.

3.2.1 Flash smelting furnace and the waste heat recovery boiler

The data and information in the ensuing paragraphs was obtained from the flash smelting furnace operators’ training manual [44].

The flash smelting furnace (FSF) produces hot flue gases at flow rate of 45800 m$_n^3$/hr (normal cubic meters per hour) and temperature of 1450°C. The flue gases are channeled into the waste heat recovery boiler (WHRB) which generates steam at 60 bar pressure, 277°C temperature and 60 tons per hour (60tph) mass flow. The WHRB is a single pressure vessel consisting of a radiation section (larger in volume) and
convection section (smaller). The flue gases from the FSF enter the boiler through the radiation section where heat transfer is mainly through radiation and the flue gases do not come into contact with the boiler tubes. In the convection bank, the gases come into contact with the membrane tube wall.

To dislodge the flue dust from the walls of the banks, a mechanical hammering system is installed. Both banks are equipped with dust hopper bottoms and drag conveyor for collecting and conveying the flue dust separated from the flue gas. The steam that is generated from the WHRB is channeled to the multi-coil dryer for drying of raw concentrate; the rest is channeled to an air cooled condenser (ACC) for condensation. Recently, a heat exchanger has been installed to utilize part of the steam for heating process nitrogen.

### 3.2.2 Slag cleaning furnace

The data and information in the following chapters and sections 3.2.3 and 3.3.1.1 was obtained from the operators training manual [45].

The slag cleaning furnace (SCF) was designed to process 1600 tons per day of FSF slag. The smelting effect in the SCF is met using heat from the exothermic reactions of materials in the furnace. To fully meet the heat demand, the SCF uses grid electricity to avail extra heat via electrical resistance between its three söderberg electrodes immersed in the molten metal, connected to a 20MVA transformer. The electricity compensates for heat needed for endothermic reduction reactions, furnace heat losses and for melting. The SCF is of circular type with a diameter of 13.5m, and height 4.7m.

The furnace is operated in batch mode, with cycles of 4 batches per day. Metallurgical coke is used for reduction of the highly oxidized FSF slag. This coke feeding is done via dosing through seven charging bins, and total feed is about 68 tons per day. Blister copper from SCF is tapped directly into the anode furnace at 235 tons per day while slag is tapped into CRF at 1300 tons per day. Flue gases from the SCF are combusted in an incinerator, thereafter cooled and later cleaned before being channeled to the stack. The flue gases consist of carbon monoxide (CO), carbon dioxide (CO$_2$), hydrogen (H$_2$), water (H$_2$O), and nitrogen (N$_2$).

The gas analysis during reduction at furnace outlet is as follows: CO: 49%; CO$_2$: 12%; H$_2$: 5%; H$_2$O: 4%; N$_2$: 30%, and the flue gas amount is 12200 m$^3$/h at temperature 1330 °C. Gas incineration and cooling is met by using combustion air amount of 42000 m$^3$/h and cooling air amount of 40000 m$^3$/h respectively. After incineration, the gas analysis is: N$_2$: 73%; O$_2$: 15%; CO$_2$: 9%; H$_2$O: 3%. Note an increase in nitrogen content, this is due to the inert air nitrogen that has been added.

Flue gas incineration is done to attain complete combustion of CO so that it is not vented to the atmosphere as an environmental protection feature. The incinerator is a
**Fig. 3.1:** Nchanga Smelter general process flow. The steam dryer feeds the dried concentrate pneumatically into the FSF. Emergent from the FSF is blister copper (tapped into the anode furnaces for casting); slag (channeled into the slag cleaning furnace); and a stream of flue gases (channeled into the WHRB). From the WHRB, the flue gases are channeled into an ESP, and later to the sulphuric acid plant for acid production. The SCF channels its molten content into two cobalt recovery furnaces where copper/cobalt alloy is recovered while the slag is channeled into a granulation pond. Source: KCM Nchanga Smelter SCF and CRF Training manual

cylindrical shaft made of heat resistant steel and lined with refractory bricks, conically shaped at the bottom. It contains a water-cooled copper element in the lower part, BIC element in conical part and horizontal elements in the off-gas combustion area. It has inside diameter of 3m and its total height is 22m. The combustibles contained in the off-gas stream are CO and H₂.
3.2.3 Cobalt recovery furnace

The cobalt recovery furnace (CRF) is operated in batch mode, to process 1300 tons per day of SCF slag, in 4 batches a day. The total amount of cobalt tapped is 70 tons per day in two batches. It is a circular-type electric furnace with diameter of 11.9m and wall height 4.7m. It is equipped with three Sòderberg electrodes with transformer power rating of 15MVA. Coke is fed as a reducing agent at 13.7 tons per day, and revert is fed at 14.5 tons per day. From the furnace, flue gases are fed into an incinerator for burning and later cooling. Maximum off-gas temperature after combustion is 1200 °C, the slag temperature reaches maximum 1359 °C. The nominal flue gas flow is 4200 m₃/h at temperature 1200°C. The flue gas analysis (%) is H₂: 4.8; H₂O: 3.0; CO: 34.6; CO₂: 14.6 and N₂: 42.9. Flue gas incineration and cooling is met using combustion air amount of 6700 m₃/h, cooling air amount of 8000 m₃/h respectively. The total gas amount is 18100 m₃/h and temperature (after incineration and cooling) is 800°C. The final gas composition is CO₂: 11.5%; O₂: 12.1%; H₂: 3.4% and N₂: 73%.

The inside diameter of this particular incinerator is 2.2m and total height 17m. It is equipped with combustion and cooling air distributors through which the combustion and the cooling air is blown in. After the incinerator is an emergency stack and valve. The minimum temperature of the off-gas after combustion is 800°C and maximum is 1200 °C. The warmed spray cooling water from SCF and CRF is collected to a common water tank from where it is pumped to the cooling water tank to be cooled by the cooling towers, thereafter pumped to the SCF and CRF for a new cycle.

3.2.4 Anode furnaces

The Nchanga smelter complex has installed in it three anode furnaces where two operate at any given time with one on standby. The three can be approximated to have identical features and operating parameters. Unfortunately, design and operation data on the three furnaces could not be obtained at the time of this research, so was the case with the sulphuric acid plant which is another waste heat source from its exothermic reactions. Analysis on both the anode furnaces and sulphuric acid plant is not conducted due to this non-availability of data.

3.3 Available energy extractable from smelter waste heat

In any heat recovery situation, it is essential to know the quality and amount of heat recoverable and also how it can be used [46]. The following sections provide an analysis of the amounts of recoverable heat at Nchanga smelter and potential for power generation. At the outset, proposed heat recovery equipment that is considered in the heat balance calculations is discussed in brief.
3.3.1 Installation of heat recovery steam generator for SCF and CRF

The existing scenario in smelters that have installed waste heat recovery systems is that common to all, they have waste heat recovery boilers. The WHRB installed at Nchanga smelter is a single pressure vessel with radiation and convection section where saturated steam is generated. Heat recovery steam generators (HRSGs), usually common in combined cycle power plants, can be multiple-pressure and designed to contain three heat exchangers (economizer, evaporator and superheater).

It can be observed that the flue gas streams from both the SCF and the two CRFs are channeled into respective incinerators with purpose of achieving complete combustion of CO and H$_2$. These flue gases are at maximum temperature of 1300$^\circ$C, later cooled down to 800 $^\circ$C. The objective of this analysis will seek to design a system that will capture the waste heat in the flue gases emergent from the incinerators. It will be sought that flue gas temperature be lowered in the envisaged heat recovery unit, reducing the temperature from the incinerator emergent temperature to below 400$^\circ$C by generating steam. The evaluation will quantify the recoverable heat in the flue gas and the steam that would be generated by installing a single pressure HRSG after the incinerators. The steam thermodynamic parameters will be evaluated to recommend appropriate utilization.

The simplified version requires that each furnace would have a separate WHRB or HRSG. Analysis will be given for possibility of installing a single HRSG into which flue gases from all the installed electric furnaces will be channeled. A schematic diagram of the incinerator connected to the envisaged HRSG and steam turbine is shown in fig. 3.2. In the HRSG, the economizer receives the feed water and uses the radiation heat from the flue gases to heat-up feed water to its saturation temperature, the evaporator uses the heat from the gases to evaporate the saturated feed water to saturated steam; and the superheater uses the convection heat from the flue gases to superheat the steam.

In the approximate calculations, heat balances across specific heat exchangers in the HRSG are not conducted, but an overall heat balance across the entire HRSG is what has been done. As can be seen in fig. 3.2, the HRSG is placed after the incinerator. The steam line is in parallel flow to the flue gas to avoid the areas with the most heated fluid flow coming into contact to protect the material of the boiler tubes. The generated steam emerges from the HRSG and is channeled into a steam turbine for power generation. In the gas stream, flue gas enters the HRSG at temperature $T_{1,gas}$ and exits at $T_{4,gas}$.

The heat balance across the HRSG between the gas side and the water/steam side respectively is as follows:

$$m_{gas} \overline{c}_{p,gas} (T_{1,gas} - T_{4,gas}) = m_{st} (h_{sh} - h_{fw}) \quad (3.1)$$
where $m_{\text{gas}}$ is the mass flow of the flue gas, $\overline{c}_{p,\text{gas}}$ is the average specific heat of the gas, $T_{1,\text{gas}}$ is the HRSG gas entrance temperature, $T_{4,\text{gas}}$ is the HRSG gas exit temperature, $m_{st}$ is the water/steam mass flow, $h_{\text{sh}}$ is the enthalpy of superheated steam exiting the HRSG and $h_{\text{fw}}$ is the enthalpy of the water entering the HRGS.

The left hand side of equation (3.1) is the quantity of heat damped by the flue gas across the HRSG, re-written as equation (3.2).

$$Q_{\text{gas}} = m_{\text{gas}} \overline{c}_{p,\text{gas}} (T_{1,\text{gas}} - T_{4,\text{gas}})$$

(3.2)

**Fig 3.2:** Configuring the incinerator, HRSG and steam turbine. The HRSG receives the hot flue gas stream from the incinerator to evaporate feed water and generate superheated steam for power generation through a steam turbine and electrical generator (G).
Combining equations (3.1) and (3.2) gives the mass flow of steam $m_{st}$ as in equation (3.3):

$$m_{st} = \frac{Q_{gas}}{(h_{sh} - h_{fw})}$$

(3.3)

3.3.1.1 Slag cleaning furnace waste heat and HRSG installation

The gas flow rate (emergent from the furnace) is $12200 \text{ m}_3^3$/h, at temperature of $1330\degree$C and gas composition CO: 49%; CO$_2$: 12%; H$_2$: 5%; H$_2$O: 4%; and N$_2$: 30%. From the SCF, the flue gas is channeled into the incinerator for combustion with air.

The combustion air amount fed into the incinerator is $42000 \text{ m}_3^3$/h as depicted in figure 3.3. After complete combustion in the incinerator, the emergent flue gas has the following composition: N$_2$: 73%; O$_2$: 15%; CO$_2$: 9%; and H$_2$O: 3%.

![Diagram of flue gas stream from the slag cleaning furnace and combustion air being fed into the incinerator to attain complete combustion of CO and H$_2$. Note the reactions that occur in the incinerator in the abundance of inert nitrogen.](image)

The gas emergent from the incinerator will be a combination of the initial gas composition and reactants from the incineration of CO and H$_2$ and excess air. It can be speculated that the prevailing large volume of combustion air (42000 m$_3^3$/h) fed into the incinerator, apart from the purpose of attaining complete combustion, is to partially achieve some level of temperature reduction for the flue gas. Consideration is made to adjust the amount of combustion air that is fed into the incinerator. The analysis will
seek to establish the actual amount of air that is needed for complete combustion of the CO and H\(_2\). The combustion reactions in the incinerator are according to the following empirical reactions:

\[
2\text{CO} + \text{O}_2 \rightarrow 2\text{CO}_2 \quad (3.4)
\]

\[
\text{H}_2 + 0.5\text{O}_2 \rightarrow \text{H}_2\text{O} \quad (3.5)
\]

As there is no free O\(_2\) in the flue gas, no reactant O\(_2\) contribution will come from the flue gas. We consider converting the reactants into moles as table 3.1 [47]. The table gives the gas compositions where the third column takes account of the 4% moisture content and gives the gas compositions on dry basis.

Consider a fuel with the following composition for total substance (%-mass): a, b, c, d, e, f. then

\[
\begin{align*}
\text{Element} & \quad \text{Molar Mass (g/mol)} & \quad \text{Analysis (g/kg\text{fuel})} & \quad \text{Analysis (mol/kg\text{fuel})} \\
\text{A} & \quad \text{RMM}_A & \quad \text{RMM}_A \\
\text{B} & \quad \text{RMM}_B & \quad \text{RMM}_B \\
\text{C} & \quad \text{RMM}_C & \quad \text{RMM}_C \\
\text{D} & \quad \text{RMM}_D & \quad \text{RMM}_D \\
\text{E} & \quad \text{RMM}_E & \quad \text{RMM}_E \\
\text{F} & \quad \text{RMM}_F & \quad \text{RMM}_F
\end{align*}
\]
Table 3.2: A tabulated procedural calculation to determine number of moles for each element in the flue gas with results given in grams of substance per kilogram of flue gas (g/kg\text{fuel}), and moles per kilogram of flue gas (mol/kg\text{fuel}).

<table>
<thead>
<tr>
<th>Reactant</th>
<th>Molar Mass (g/mol)</th>
<th>Analysis (g/kg\text{fuel})</th>
<th>Analysis (mol/kg\text{fuel})</th>
</tr>
</thead>
<tbody>
<tr>
<td>CO</td>
<td>28</td>
<td>470.4</td>
<td>16.800</td>
</tr>
<tr>
<td>CO\textsubscript{2}</td>
<td>44</td>
<td>115.2</td>
<td>2.618</td>
</tr>
<tr>
<td>H\textsubscript{2}</td>
<td>2.02</td>
<td>48.0</td>
<td>23.762</td>
</tr>
<tr>
<td>H\textsubscript{2}O</td>
<td>18.02</td>
<td>40.0</td>
<td>2.220</td>
</tr>
<tr>
<td>N\textsubscript{2}</td>
<td>28</td>
<td>288.0</td>
<td>10.286</td>
</tr>
<tr>
<td>O\textsubscript{2}</td>
<td>32</td>
<td>unknown</td>
<td>Unknown</td>
</tr>
</tbody>
</table>

Table 3.3: Conversion of reactants to moles, where the results are based on the formulae in table 3.2.

From the first empirical reaction, equation (3.4), we deduce that the ratio of reactants CO and O\textsubscript{2}, and the product (CO\textsubscript{2}) is 2:1:2. Table 3.3 gives the result of the actual number of moles of CO as 16.8. Using the ratio of the reactants implies that the 16.8 moles of CO would require 8.4 moles of O\textsubscript{2} to produce 16.8 moles of CO\textsubscript{2}. Nitrogen does not take part in the reaction, retaining the number of moles in the product gas in addition to the air nitrogen. Similarly, in the second empirical reaction, equation (3.5), the ratio of the reactants H\textsubscript{2} and O\textsubscript{2}, and the product H\textsubscript{2}O is also 2:1:1. Table 3.3 gives the result of the actual number of moles of H\textsubscript{2} as 23.762. Using the ratio of the reactants implies that 11.881 O\textsubscript{2} is what is required to complete the reaction to produce 23.762 moles of H\textsubscript{2}O.

Therefore, oxygen needed for complete combustion of the flue gas is:

\[ [O_2] = 8.400 + 11.881 = 20.281 \text{mol/kg}_{\text{gas}} \]

The N\textsubscript{2} in the air fed into the incinerator (air nitrogen) can be determined from:

\[ [N_{2,\text{air}}] = 3.76[O_2] \quad (3.6) \]

Substituting the values yields

\[ [N_{2,\text{air}}] = 3.76[O_2] = 3.76 \times 20.281 = 76.257 \text{mol/kg}_{\text{gas}} \]

The molar amount of air needed for stoichiometric combustion per kg of the flue gas is obtained from:

\[ I_0 = [O_2] + [N_2] \quad (3.7) \]
Substituting the values yields:

\[ I_0 = 20.281 + 76.257 = 96.538 \text{mol/kg}_{\text{gas}} \]

The gas composition after the incinerator, based on the empirical reaction equations (3.4) and (3.5), and table 3.3 is as in table 3.4.

<table>
<thead>
<tr>
<th>Element</th>
<th>No. of moles (mol/kg\text{gas})</th>
</tr>
</thead>
<tbody>
<tr>
<td>CO\text{ }_2</td>
<td>16.8</td>
</tr>
<tr>
<td>O\text{ }_2</td>
<td>8.4</td>
</tr>
<tr>
<td>H\text{ }_2</td>
<td>23.762</td>
</tr>
<tr>
<td>N\text{ }_2</td>
<td>10.286</td>
</tr>
<tr>
<td>H\text{ }_2\text{O}</td>
<td>23.286</td>
</tr>
<tr>
<td>N\text{ }_2,\text{air}</td>
<td>76.257</td>
</tr>
</tbody>
</table>

**Table 3.4:** Summary of gas composition after combustion in the incinerator.

The Stoichiometric amount of flue gas \((g_0)\) is the sum of the number of moles obtained in the foregoing analysis:

\[ g_0 = [CO_2] + [O_2] + [H_2] + [N_2] + [H_2O] + [N_2,\text{air}] \]  

(3.8)

Substituting the values yields

\[ g_0 = 158.791 \text{mol/kg}_{\text{gas}} \]

We assume excess air of 20% implying an excess air factor \(m = 1.20\). The real amount of air needed is obtained from equation (3.9).

\[ I = mI_0 \text{(mol/kg}_{\text{gas})} \]  

(3.9)

Substituting the values yields:

\[ I = 1.2 \times 96.538 = 115.846 \text{(mol/kg}_{\text{gas})} \]

Combustion air can be computed from equation (3.10):

\[ m_{\text{air}} = I \times \text{RMM} \]  

(3.10)
where $m_{air}$ is the combustion air and RMM is the total molar mass of the gas composition in the previous result, i.e., CO$_2$, O$_2$, H$_2$, H$_2$O, and N$_2$.

$$RMM = (44 + 2.02 + 18.02 + 32 + 28) g = 124.04 g = 0.12404 kg/mmol$$

Substituting the values in equation (3.10) gives:

$$m_{air} = 115.846 mol \times 0.12404 kg/mmol = 14.370 kg/s$$

The foregoing is the actual amount of air required to achieve complete combustion in the incinerator. We now convert the 12200 m$^3$/h flue gas flow (before combustion in the incinerator) to kg/s to simplify further computations. To do this we assume ideal gas conditions for the flue gas at normal conditions (1 atm, 273 K), and molar volume = 0.0224 m$^3$/mol. We convert the m$^3$/h to kg/s using equation (3.11). Respective molar masses of the gases are given in table 3.5.

<table>
<thead>
<tr>
<th>Flue Gas component</th>
<th>Molar Mass (g)</th>
<th>% Composition</th>
</tr>
</thead>
<tbody>
<tr>
<td>CO</td>
<td>28</td>
<td>49</td>
</tr>
<tr>
<td>CO$_2$</td>
<td>44</td>
<td>12</td>
</tr>
<tr>
<td>H$_2$</td>
<td>2.02</td>
<td>5</td>
</tr>
<tr>
<td>H$_2$O</td>
<td>18.02</td>
<td>4</td>
</tr>
<tr>
<td>N$_2$</td>
<td>28</td>
<td>30</td>
</tr>
</tbody>
</table>

**Table 3.5:** Gas composition and respective molar masses, to be inputted into equation (3.11) for the conversion of flue gas flow to kg/s.

$$m_{gas} = \sum [RMM_i (kg) / 0.0224 (m^3/mmol)]aV (m^3/s)$$

(3.11)

where $a$ = percentage composition of respective gases, $V$ is the volume flow of the gas stream and $RMM_i$ is the molar mass of the respective gases.

Substituting the values gives:

$$\dot{m}_{gas} = [(0.028/0.0224) \times 0.49 + (0.044/0.0224) \times 0.12 + (0.00202/0.0224) \times 0.05 + (0.01802/0.0224) \times 0.04 + (0.028/0.0224) \times 0.3] \times (12200/60 \times 60) = 4.18 kg/s$$

The total flue gas emergent from the Incinerator will thus be:
For design purposes, it is estimated that the flue gas exit temperature from the HRSG will be 370 °C which is well above the H₂O dew point (a function of moisture content and temperature) and thus expect no condensation. The incinerator exit temperature in the prevailing scenario will be taken as entry temperature of gas into the envisaged HRSG. For purposes of the calculation, the flue gas exit temperature from the incinerator is taken as 1000 °C (in current scenarios this ranges between 800 – 1200 °C, hence 1000 is good default estimation).

The average specific heat \( \bar{c}_{p,\text{gas}} \) is determined as follows: consider flue gas characteristics between 1000°C and 370°C. The average specific heat is dependent on both temperatures and the gas content (x) and will be determined from the average temperature with estimation of the gas content at 30%.

\[
T_{\text{ave}} = \left( T_{1,\text{gas}} - T_{4,\text{gas}} \right) / 2
\]

where \( T_{1,\text{gas}} \) is the HRSG approach temperature (1000°C) and \( T_{4,\text{gas}} \) is the HRSG exit gas temperature (370°C). Substituting the values yields \( T_{\text{ave}} = 685°C \). The \( \bar{c}_{p,\text{gas}} \) value can be looked up from a \( c_p \)– temperature diagram as 1.12 kJ/kg.

To set design parameters for the single-pressure HRSG, it will be desired to generate superheated steam at 60 bar pressure, and 420°C temperature. Table 3.6 sows a summary of the parameters across the HRSG.

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Description</th>
<th>Quantity</th>
</tr>
</thead>
<tbody>
<tr>
<td>( m_{\text{gas}} )</td>
<td>mass flow of flue gas [kg/s]</td>
<td>17.28</td>
</tr>
<tr>
<td>( \bar{c}_{p,\text{gas}} )</td>
<td>average specific heat of the gas [kJ/kg]</td>
<td>1.12</td>
</tr>
<tr>
<td>( T_{1,\text{gas}} )</td>
<td>HRSG gas entrance temp [°C]</td>
<td>1000</td>
</tr>
<tr>
<td>( T_{2,\text{gas}} )</td>
<td>HRSG gas exit temp [°C]</td>
<td>370</td>
</tr>
<tr>
<td>( m_{\text{st}} )</td>
<td>mass flow of the generated steam [kg/s]</td>
<td>?</td>
</tr>
<tr>
<td>( T_{fw} )</td>
<td>Temperature of feed water [°C]</td>
<td>100</td>
</tr>
<tr>
<td>( T_{sh} )</td>
<td>Temperature superheated steam [°C]</td>
<td>420</td>
</tr>
<tr>
<td>( h_{sh} )</td>
<td>HRSG Steam exit enthalpy (turbine entry:60bar, 420°C) [kJ/kg]</td>
<td>3226</td>
</tr>
<tr>
<td>( h_{fw} )</td>
<td>Enthalpy saturated liquid water (@100°C), HRSG entry [kJ/kg]</td>
<td>419.1</td>
</tr>
</tbody>
</table>

Table 3.6: A summary of design parameters across the HRSG.

Substituting values into equation (3.2) yields the following:
\[ Q_{gas} = 18.55 \text{kg/s} \times 1.12 \text{kJ/kg} \times (1000 - 370) = 13088.88 \text{kJ/s} = 13.08 \text{MW} \]

Substituting values into equation (3.3) yields the steam mass flow as follows:

\[ m_{st} = \frac{13088.88}{3226 - 419.1} = 4.66 \text{kg/s} \]

3.3.1.2 Summary of steam generated from envisaged HRSG

Table 3.7 summarizes the steam parameters obtainable from the envisaged HRSG for slag cleaning furnace flue gas.

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Quantity</th>
</tr>
</thead>
<tbody>
<tr>
<td>Temperature (°C)</td>
<td>420</td>
</tr>
<tr>
<td>Pressure (bar)</td>
<td>60</td>
</tr>
<tr>
<td>Mass flow (kg/s)</td>
<td>4.66</td>
</tr>
</tbody>
</table>

*Table 3.7: Summary of results for steam parameters in the envisaged HRSG.*
**Fig. 3.4:** System configuration showing the incinerator, HRSG and steam turbine unit. Flue gas stream from the SCF is channeled through the envisaged HRSG. Superheated steam is generated and is emergent from the superheater and is channeled into a steam turbine for power generation.

### 3.3.1.3 Power generation from HRSG steam at SCF

Consider installation of a steam turbine unit to utilize the generated steam for power generation. Fig. 3.3 depicts a possible placement of the main components, i.e., incinerator, HRSG and steam turbine. The steam cycle is of recirculation type with a condenser after the turbine and the water is pressurized by a pump before entry into the HRSG. The water/steam passes through three heat exchangers in the HRSG, economizer, evaporator and superheater (denoted as Eco, Eva and Sup respectively in fig. 3.4).

The power generation potential of the steam turbine from the given parameters in the foregone analysis can be obtained from the following equation:

\[
P_T = m_{st} (h_{in} - h_{out}) \eta_{mec} \eta_{gen}
\]  
(3.13)

where: \( m_{st} \) is the steam mass flow, \( h_{in} \) is the turbine entrance enthalpy, \( h_{out} \) is the turbine exit enthalpy, \( \eta_{mec} \) is the mechanical efficiency of the turbine and \( \eta_{gen} \) is the electrical generator efficiency.

We will assume a steam turbine with both mechanical and generator efficiencies of 95% respectively. The standard steam tables give the value for \( h_{out} \) for saturated steam at 1bar (condenser pressure) as 2673.8 kJ/kg. We substitute the values into equation (3.13) to obtain the following result as the power generation potential for SCF flue gas through an HRSG and steam turbine unit.

\[
P_{SCF} = 4.66 \text{ kg/s} \times (3226 - 2673.8) \text{ kJ/kg} \times 0.95 \times 0.95 = 2322.36 \text{ kJ/s}
\]

\[
= 2.32 \text{ MW}
\]

### 3.3.2 Cobalt recovery furnace waste heat and power generation

The Nchanga smelter plant has two cobalt recovery furnaces whose fundamental operational parameters can be approximated to be the same. The analysis will be based on the given parameters of: temperature range after combustion (800-1200°C), gas amount (4200 m\(_3\)/h), combustion air amount (6700 m\(_3\)/h), cooling air amount (8000 m\(_3\)/h), total gas amount (18100 m\(_3\)/h) and gas composition ([CO]: 11.5%; [O\(_2\)]: 12.1%; [H\(_2\)]: 3.4%; and [N\(_2\)]: 73%). From the gas composition, notice that the gas has zero moisture content.
In the case of SCF, the gas composition after the incinerator was known from which it was easy to deduce the actual amount of air required for complete combustion in the incinerator to avert suspicion that the quantity in the prevailing scenario was too much. In the case of the CRF, this is not known. We proceed to conduct the analysis with prevailing amount of combustion air being fed into the incinerator. For flue gas (4200 m$^3$/h); we convert this to [kg/s] using equation (3.11) assuming near-ideal gas conditions so that the equation yields:

$$m_{gas} = 1.15\text{kg/s}$$

Combustion air amount is 6700 m$^3$/h. Approximating that gas composition is N$_2$ (78%; RMM=28) and O$_2$ (21%; RMM=32) and neglecting the rest, also assuming ideal gas conditions, equation (3.11) gives

$$m_{air} = 2.37\text{kg/s}$$

The total flue gas flow will be the sum of the foregoing two results;

$$m_{tot} = 3.525\text{kg/s}$$

Like in the previous case, we consider installation of a single-pressure HRSG and a steam turbine power plant unit. With all parameters remaining as in the SCF case, we obtain the recoverable heat from equation (3.2) as:

$$Q_{gas} = 2485.8kW = 2.49MW$$

Substituting this value of recoverable heat and the enthalpy parameters to compute the steam flow rate into equation (3.3) yields:

$$m_{st} = 0.89\text{kg/s}$$

Noting that there are two cobalt recovery furnaces in operation at any given time, flue gases from both of them will be channeled into a single HRSG and thus the total steam output will be double the above result, that is,

$$m_{st,tot} = 1.77\text{kg/s}$$

Passing this steam into a steam turbine unit with the relevant parameters as in the case of SCF steam, equation (3.13) will yield turbine output power of:

$$P_{CRF} = 0.88MW$$

This thesis project has not taken deep treatment of the combustion scenarios, ideal conditions of heat transfer and the isentropic parameters across the turbine but the analysis and results suffice for consideration prior to decision making.
3.3.4 Flash smelting furnace waste heat and power generation

Currently, enthalpy in the flue gases from the FSF is expended in the WHRB to generate steam. Only a portion of this steam goes to productive utilization in the two multi-coil steam dryers for drying of the wet copper concentrate. The rest of the steam is disposed-off through an air cooled condenser (ACC) for condensation at a cost using electric fans. The task here is to re-design the system, starting from the furnace outlet, to harness the enthalpy in the flue gas such that there is maximum utilization of the waste heat.

Installation of a HRSG as in the earlier cases that will generate superheated steam for power generation will be considered. The flue gas flow from the FSF is known to be 45800 m$^3$/h at temperatures of up to 1450 °C. The flue gas composition is SO$_2$: 14%; CO$_2$: 4%; H$_2$O: 3%, N$_2$: 78% and a dust load of 14tph.

Assuming ideal gas conditions, we convert the 45800 Nm$^3$/hr to kg/s using equation (3.11) as follows:

$$m_{gas} = [(64/0.0224) \times 0.14 + (44/0.0224) \times 0.04 + (18.02/0.0224) \times 0.03 + (28/0.0224) \times 0.78] \times 45800/(60 \times 60) g / s = 18800.47 g / s = 18.80 kg / s$$

No combustion reactions occur in the WHRB so that, to determine the specific heat of the gas, we approximate gas composition to zero (x=0). Consider a HRSG with entrance temperature of flue gas (emergent from the FSF) at 1350 °C and exit temperature of 370 °C. The specific heat of the gas across the HRSG is dependent on the average temperature and the gas content. Average temperature is (1350+370)/2 = 860 °C, at x=0 the specific heat from the (specific heat / temperature) graph is read-out as 1.18 kJ/kgK. Substituting into equation (3.2) yields the flue gas enthalpy drop across the HRSG as:

$$Q_{gas} = 18.80 kg / s \times 1.18 kJ / kgK \times (1350 - 370) = 21740 kJ / s = 21.74 MW$$

We assume that feed water into the envisaged HRSG is re-circulated and enters at 100 °C with enthalpy (as function of temperature) of 419.1kJ/kg. The emergent superheated steam is at 420 °C, 60 bars, and enthalpy of 3226kJ/kg. Substituting these values into equation (3.3) yields the steam mass flow as:

$$m_{st} = 21740 kJ / s / (3226 - 419.1) kJ / kg = 7.75 kg / s = 27.88 tph$$
Interpretation: At the prevailing conditions of flue gas from the FSF, it is possible to generate steam at 60 bars, 420°C; and mass flow of 7.75 kg/s (27.88 tph) through a heat recovery steam generator.

If we consider a steam turbine with both mechanical and generator efficiencies of 95%, electric power generation potential can be obtained using equation (3.13) as:

\[ P_T = 7.75\, \text{kg/s} \times (3226 - 2673.8)\, \text{kJ/kg} \times 0.95 \times 0.95 = 3.86\, \text{MW} \]

**Figure 3.5:** Envisaged configuration of FSF, HRSG, steam turbine (T) and electric power generator (G). Steam extractions to meet demand at the dryers and for nitrogen heating are done in the turbine exit. The steam amount to be channeled towards nitrogen heating and concentrate drying shall depend on demand; a by-pass route to the condenser has been created to take-up any excess steam.
The foregoing is the estimated amount of power (real conditions) that would be generated if all the steam generated in the envisaged HRSG were inputted into a steam turbine unit. However, it should be noted that part of the steam generated is needed for concentrate drying in the dryer and for heating of process nitrogen at the oxygen plant.

Since saturated steam suffices for these other needs and noting that the dryers are fed with steam at maximum pressure of 14 bars [48], steam extractions in the turbine exit (figure 3.4) shall be at a selected 20 bars and prevailing temperature of 277°C.

In figure 3.4, two routes for steam from the turbine can be noted. One route takes the saturated steam to the nitrogen heat exchanger and the steam dryers while the other one by-passes the heat exchangers to take steam direct into the condenser. The two routes are for regulation purposes and uptake of steam to the heat exchangers shall depend on demand.

The design maximum wet concentrate feed rate for the twin dryers is 160 tons per hour. When on full load, the steam has to evaporate 15 tones per hour of water from the wet copper concentrate. To determine the amount of enthalpy damped across the steam dryers and the nitrogen heat exchanger (HE, in figure 3.5), information on parameters of the steam in the return lines from these heat exchangers was not available. However, we note that currently, steam to the twin dryers from the boiler is channeled at a regulated maximum pressure of 14 bars. For allowance purposes, the turbine exit steam shall be at 20 bars and the current temperature which is 277°C. Saturated steam at 20 bars and 277°C has enthalpy of 2798.7kJ/kg (as a function of pressure). Turbine power output will be computed from equation (3.13) where in this case, $m_{st} = 7.75$, $h_{in} = 3226$ kJ/kg, $h_{out} = 2798$ kJ/kg, $\eta_{mec} = 0.95$, $\eta_{gen} = 0.95$. So that substituting the values yields:

$$P_{FSF} = 2994kW = 2.99MW$$

### 3.3.5 Anode furnaces

Due to non-availability of data at the time of this research, a thorough analysis on the anode furnaces is not conducted; however, the treatment here would be similar to that done on the SCF and CRF.

### 3.3 Achievable energy efficiency and carbon emission reductions

The foregone analysis discussed possible installation of HRSGs to harness waste heat contained in the flue gases from the respective furnaces. In three cases, potential for power generation was calculated. Installation of these systems would entail a localized power generation system that would displace equivalent amount of power from the grid,
this would result into overall energy efficiency improvement for the smelter plant. Furthermore, this localized power generation system would emanate into carbon emission reductions which were originally associated to consumption the displaced grid power.

### 3.4.1 Energy efficiency improvement

Results from the analysis done on SCF, CRF and FSF gave a combined power generation potential as follows:

\[
P_{\text{tot}} = P_{\text{SCF}} + P_{\text{CRF}} + P_{\text{FSF}}
\]

\[
P_{\text{tot}} = 2.32 + 0.88 + 2.99 = 6.19\, \text{MW}
\]

Power demand for Nchanga smelter is about 40 MW, these proposed installations would thus offset the plant’s power demand by about 15 % or higher when all the waste heat points are considered.

Historically, at the decommissioned Nkana smelter, flue gas from the reverberatory furnace was emergent at 900 - 1200°C. The flue gas was channeled into a heat recovery boiler that had an in-built superheater. The boiler used to generate 18 tph of steam at temperatures ranging 340 – 400°C. The plant had a steam turbine unit for power generation that became obsolete with time and was shut down several years before the whole plant was decommissioned in 2008.

### 3.4.2 Associated carbon emission reductions

These emission reductions associated with the proposed installations will be quantified using standard emission factors. Noting that the installations would have a bearing on consumption of grid electricity, the associated grid emission factor for electricity consumed at Nchanga shall be used. A grid emission factor (GEF) is the specific quantity of emissions associated to generation of the electricity being consumed. It is here given in tons of carbon dioxide equivalent per mega watt-hour (tco$_2$/MWh).

The nitrogen steam heat exchanger at Nchanga smelter oxygen plant was recently installed to replace a 1.35MW electric heater and was implemented by KCM as a CDM project. Zambia meets almost 100 percent of its electricity demand from hydropower generation. Hydropower is a renewable energy source and has very minimal greenhouse gas emissions associated with it. The national GEF for Zambia is a paltry 0.007 tco$_2$/MWh [49]. This low GEF for Zambia made implementation of grid-related CDM projects over the years unattractive. However, the UNFCCC has allowed use of cross-border GEFs to allow those countries that have near-zero emissions such as Zambia to participate under CDM [50]. Zambia sits on the hub of the Southern African
power pool (SAPP) grid whose GEF is higher than its national grid owing to countries such as South Africa that are part of this grid that meet the bulk of their power demand from coal. KCM took advantage of this fact and embarked on CDM projects that are grid-related. The GEF for the SAPP is 0.9996 tCO$_2$/MWh [51] and for purposes of demonstrating the value of implementing energy-efficient undertakings, quantification of all carbon emission reductions shall be done using the SAPP GEF.

Recall the result for the combined power generation potential from the SCF, CRF and FSF. The combined power output from the three furnaces was found to be:

\[ P_{tot} = 6.19 \text{MW} \]

The associated annual carbon emission reductions assuming 100% plant availability (8760 hours) is obtained from:

\[ CO_{2, \text{emissions}} = 8760 P_{tot} GEF_{SAPP} \quad (3.15) \]

Substituting the values into equation (3.15) yields:

\[ CO_{2, \text{emissions}} = 8760h \times 6.19 \text{MW} \times 0.9996 tCO_{2,e} / \text{MWh} = 54202 tCO_{2,e} \]

Note that these installations aimed at utilizing current excess boiler steam will also displace the ACC unit. The energy consumption for the ACC unit is 4500 kWh/day [52]. This energy will be saved as the ACC will not be operational when the envisaged installations are implemented. Again assume 365 days of operation per year so that:

\[ CO_{2, \text{em, ACC}} = 4.5 \text{MWh/day} \times 365 \text{days} \times 0.9996 tCO_{2,e} / \text{MWh} = 164 tCO_{2,e} \]

Combining the carbon emission reductions from the foregoing two results yields the carbon emissions reductions which is the potential certified emission reductions attainable:

\[ CO_{2, tot} = 54202 + 164 = 55843 tCO_{2,e} / \text{year} \]

3.5 Other waste heat utilization scenarios

The foregoing section gave an analysis of overhauling the existing WHRB and replacing it with an HRSG that would generate superheated steam from the prevailing FSF flue gas conditions. This section considers installation of a separate smaller boiler in addition to the existing WHRB to superheat the steam from its prevailing temperature of
270°C to 420°C and 60bars. Consider a boiler that will be fired by an option of available fuels, i.e., HFO and biomass.

3.5.1 Power generation potential from installation of superheat boiler

Specific enthalpy for superheated steam at 420°C and 60bars as deduced in the foregone sections is 3226 kJ/kg. Noting that the boiler generates 60tph, and that the dryer at full load consumes 30tph, and the nitrogen heat exchanger 3tph, the excess steam available for power generation is therefore:

\[
m_{st} = 60 - 30 - 3 = 27tph = 7.5kg/s
\]

Figure 3.6: FSF, WHRB, steam superheat boiler and turbine system configuration. Steam is generated in the WHRB and channeled into a boiler for superheating and thereafter into a steam turbine for power generation.

Assume a turbine with design exit conditions of 100°C and 1bar pressure in the condenser. At these conditions, the enthalpy in the turbine exit is 2673.8kJ/kg. Turbine mechanical and generator efficiencies are both assumed to be 95% each. We estimate
the power generation potential at these steam superheat conditions using equation (3.13) as follows:

\[ P_T = 7.5 \text{kg/s} \times (3226 - 2673.8) \text{kJ/kg} \times 0.95 \times 0.95 = 3737.4 \text{kJ} = 3.74 \text{MW} \]

Figure 3.5 depicts a configuration showing the gas flow from the FSF into the WHRB and then to the ESP. The steam cycle is a recirculation type, from the WHRB where steam is generated at 277°C, 60bars; it is channeled into a superheat boiler from where it emerges for inlet into the turbine at 420°C, 60bars.

### 3.5.1 Fuel Analysis for superheat boiler

This research identified HFO and biomass as readily available fuels that can be used to fire the superheat boiler.

#### 3.5.1.1 HFO as boiler fuel

Consider HFO with lower heating value of 42 MJ/kg as the fuel to be used in the superheater. Records show that the allowable range for HFO at the smelter quality laboratory is 42-44 MJ/kg. Taking the boiler combustion efficiency to be 90% and heat transfer efficiency of 30%, we seek to quantify the amount of HFO required for superheating and the associated greenhouse gas emissions. Noting that in determining the quantity of fuel required we need to take account of the 90% and 30% combustion and thermal efficiencies respectively, (in other words 10% or 70% inefficiencies) the boiler output power \(Q_b\) is computed as in equation (3.16).

\[
Q_b = \dot{m}_{st} (h_{sh} - h_{in}) \times (1 + (1 - \eta_c)) \times (1 + (1 - \eta_t))
\]

\[ (3.16) \]

where \(h_{in}\)is the specific steam enthalpy at 277°C and 60 bar, \(h_{sh}\) is the exit enthalpy of the superheated steam at 420°C, 60bars, \(\eta_c\) and \(\eta_t\) are the combustion and thermal efficiencies respectively. Here, \(h_{in} = 2790 \text{kJ/kg}\) and \(h_{sh} = 3226 \text{kJ/kg}\). Substituting these values into equation (3.16) we get:

\[
Q_b = 7.5 \text{kg/s} \times (3226 - 2790) \text{kJ/s} \times 1.1 \times 1.7 = 6114.9 \text{kJ} = 6.11 \text{MW}
\]

The mass flow of fuel is calculated using equation (3.17):

\[
m_{fuel} = \frac{Q_b}{LHV}
\]

\[ (3.17) \]
In equation (3.17), \( LHV \) is the lower heating value of the fuel. Substituting the values gives:

\[
m_{\text{fuel,HFO}} = \frac{6114.9\text{kW}}{42000\text{kJ/kg}} = 0.1456\text{kg/s} = 0.524\text{tph}
\]

### 3.5.2.2 Associated Carbon Emissions to HFO Boiler

The carbon emission factor for HFO can be looked-up from the 2006 IPCC guidelines for national greenhouse gas inventories as \( 77.4\text{tCO}_2/\text{TJ} \). Assuming 100% operation and availability of the boiler, the annual carbon emission is obtained from equation (3.18).

\[
\begin{align*}
CO_2,HFO &= tP_{\text{tot}}EF_{HFO} \\
(3.18)
\end{align*}
\]

where \( t \) is the yearly time of operation and \( EF_{HFO} \) is the emission factor for HFO. Substituting the values yield:

\[
CO_2,HFO = (8760 \times 60 \times 60)s \times 6114.9\text{kW} \times 77.4tCO_2/\text{TJ} = 14925tCO_2/\text{year}
\]

Recall that in section 3.5 it was deduced that removal of the ACC implies reduction of Nchanga smelter carbon emissions by 1,641 tCO\(_2\)/year. The current analysis of a superheat boiler also implies removal of the ACC operation. The previous result shows that installation of an HFO boiler leads to carbon emission reductions of 14,925 tCO\(_2\)/year. The net effect is therefore an increase in carbon emissions by:

\[
CO_2,HFO,\text{net} = 14925 - 1641 = 13284tCO_2/\text{year}
\]

### 3.5.2.3 Biomass as fuel for superheated steam generator

Biomass was identified as a fuel for consideration owing to vast potential from sawmills scattered on the Copperbelt province. The Copperbelt province has a total of about 587 Sawmills, all holding concession from the Zambia Forest and Forestry Industries Corporation (ZAFFICO). It is estimated that 86,000 m\(^3\) or 20% of the logs cut are never collected from the bush, 206,000 m\(^3\) or 60% of the logs harvested are wasted as follows [53]: wood barks (10%) or equivalent to 20,640 m\(^3\) per annum, slabs (20%) or equivalent to 41,280 m\(^3\) per annum, sawdust (20%) or equivalent to 41,280 m\(^3\) per annum, off-cuts (10%) or equivalent to 20,640 m\(^3\) per annum. The foregoing is the Copperbelt province waste-biomass potential into which the company can tap into as fuel source for the envisaged biomass-fired superheat boiler.

For computation purposes, we consider biomass with lower heating value of 15 MJ/kg as the fuel to be used in the superheat boiler. Furthermore, we consider the boiler to
have combustion efficiency of 90% and thermal efficiency 30%. We seek to quantify the amount of biomass required for superheating. In section 3.5.1.1 we calculated the boiler heat load as:

\[ Q_b = 6114.9kW = 6.11MW \]

The mass of fuel is computed from equation (3.17) as follows:

\[ m_{\text{fuel, biomass}} = \frac{6114.9kW}{15000kJ/kg} \times \frac{0.4077kg}{s} = 1.47tph \]

3.5.2.4 Associated carbon emissions with biomass as boiler fuel

When a tree is alive, it takes-in carbon dioxide from the air through the process of photosynthesis. When the tree is cut and taken as biomass fuel and combusted, its complete combustion produces carbon dioxide as one of the products. It can be noted that the carbon dioxide produced from combustion of woody biomass is that which was already in the atmosphere, hence no new CO\(_2\) would have been generated. The carbon finance mechanisms do not therefore, consider carbon emissions from biomass where the combustion is said to generate zero carbon emissions. However, identification of a suitable baseline scenario for the demonstration of additionality is what qualifies biomass projects as CDM or other carbon trading undertakings. A baseline scenario is one that would occur in the absence of the implemented (biomass) project.

In the current case, we note that the biomass is to be used as fuel to meet a heating demand. The first competing baseline scenario would be to meet this heat demand using HFO as boiler fuel, which then entails a CDM project with potential of 13,284 tCO\(_2\)/year (i.e. 13,284 certified emission reductions, or carbon credits) as deduced from section 3.5.1.2. The second point is to note that the objective of this biomass combustion is to generate electricity. In the second case, the baseline scenario becomes the quantity of carbon emissions associated with the amount of electricity being generated that would be avoided if this electricity were consumed from the grid. Recall that the power generation potential from installation of this superheat boiler was deduced in section 3.5.1 as 3.74MW.

Using the SAPP GEF, the associated emissions if this power were from the grid is calculated from equation (3.15), assuming 100% annual availability as:

\[ CO_{2,\text{biomass}} = 8760h \times 3.74MW \times 0.9996tCO_2/MWh = 32749tCO_2/year \]

Noting that the ACC also gets decommissioned in this scenario, the net emission reduction becomes:
\[ CO_{2,\text{biomass, tot}} = 32749 + 1641 = 34390 \text{tCO}_2 \text{ / year} \]
CHAPTER 4

ANALYSIS OF TLP HEATING DEMAND BASED ON NCHANGA SMELTER WASTE HEAT

4.1 Process scenarios requiring heat at TLP

The Nchanga tailings leach plant (TLP) consists of several stages in processing of the acid-soluble copper received from Nchanga concentrator. A review of steps and processes at TLP that are critical to heating demand is presented in the subsequent sections.

4.1.1 Compressed air for Pachucas’

The TLP treats acid-soluble copper from the concentrator plant and spent slurry from old dams that still contains a bit of recoverable copper. The slurry from the concentrator plant also contains impurities of acid-insoluble copper that remain from the extraction process [54].

The slurry is fed into twelve fluidized bed reactors (cylindrically shaped, conical at the bottom) each with effective volume of 833 m$^3$. The fluidization of the material in the reactor is achieved by injecting a jet of compressed air at the bottom of the reactor at pressure range (3.1 - 4.0 bars) which bubbles-up the reactor, in the process keeping the reactants in suspended form. For this purpose, three compressors are installed, each producing 17,000m$^3$/hr of compressed air, with maximum of two being run at each given time.

In the prevailing scenario, the temperature of the compressed air being fed into the reactor is at ambient levels, (25 – 30°C). However, due to the presence of acid-insoluble copper, and also to improve recoveries of the dissolved copper, the company desires that the compressed air be heated to optimum reaction temperature of 80°C. The task is to devise a mechanism for heating-up the compressed air at minimum cost, preferably using available heat generated at the nearby Nchanga smelter.

Figure 4.1 depicts a Pachucas’. The figure shows the main materials being inputted at the top of the reactor. The main materials inputted into the Pachucas’ include sulphuric acid, copper slurry from Nchanga concentrator, spent slurry from the old dams and raffinate (an organic compound). Compressed air is fed at the bottom of the reactor and bubbles up through the height of the reactor to the open top. The figure also shows the rubber lining inside the metallic casing of the reactor.
Fig. 4.1: A two-dimensional view of a Pachucas’ ferric reactor, showing the main feed materials. Compressed air is injected at the bottom of the reactor. The metallic casing of the reactor is lined with rubber material inside.
We proceed with demand side analysis of the required heat and base the calculations on the available waste heat and excess steam at the nearby smelter. Consideration will be made comparing the alternative case of installing an electric heater, and a fuel-powered boiler to benchmark quantification of carbon emission reductions.

4.1.1.1 Installation of steam heat exchanger

Chapter 3 gave a detailed analysis of waste heat scenarios associated with the waste heat recovery boiler (WHRB). Recall that currently, the WHRB generates steam at 60 bar pressure, 277°C temperature and mass flow of 60 tons per hour (60tph). It was noted that about 50 percent of the steam generated at the WHRB goes to waste. Figure 4.2 shows compressed air from the two compressors and steam from the WHRB being channeled into a heat exchanger. The heated compressed air from the heat exchanger is then channeled into the Pachucas’ reactor.

![Diagram](image-url)

**Fig. 4.2:** Heat-up of agitation air for the Pachucas’. Compressed air from the two compressors and steam from Nchanga smelter WHRB being channeled into a heat exchanger.

We seek to determine the quantity of steam at the prevailing parameters required to heat-up the compressed air to the desired temperature of 80°C. The heat balance across the heat exchanger is as follows:
\[ m_{\text{air}} \cdot \bar{c}_{p,\text{air}} (T_{\text{in}} - T_{\text{out}}) = m_{\text{st}} (h_1 - h_2) \]  

(4.1)

where \( m_{\text{air}} \) is the mass flow of the compressed air, \( \bar{c}_{p,\text{air}} \) is the average specific heat for air across the heat exchanger, \( T_{\text{in}} \) is the air inlet temperature, \( T_{\text{out}} \) is the air exit temperature from the heat exchanger, \( m_{\text{st}} \) is the steam mass flow, \( h_1 \) is the steam inlet enthalpy (at 60bar, 277°C) and \( h_2 \) is enthalpy of saturated liquid water at outlet (100°C)

Here \( m_{\text{air}} = 34,000 \text{m}^3/\text{h} \), \( T_{\text{in}} = 25^\circ\text{C} \), \( T_{\text{out}} = 80^\circ\text{C} \), \( h_1 = 2790 \text{kJ/kg} \) and \( h_2 = 419.1 \text{kJ/kg} \)

The left hand side of equation (4.1) is the enthalpy gain (\( Q_{\text{air}} \)) by the compressed air. We re-write the equation as:

\[ Q_{\text{air}} = m_{\text{air}} \cdot \bar{c}_{p,\text{air}} (T_{\text{in}} - T_{\text{out}}) \]  

(4.2)

Combining equations (4.1) and (4.2) gives the steam mass flow as follows:

\[ m_{\text{st}} = \frac{Q_{\text{air}}}{(h_1 - h_2)} \]  

(4.3)

We convert the volume flow rate of compressed air (\( V \)) from \( \text{m}^3/\text{h} \) to \( \text{kg/s} \) by taking the density (\( \rho \)) of air at 25°C as 1.18 kg/m\(^3\) as follows:

\[ m = \rho V \]  

(4.4)

Substituting the values into equation (4.3) yields:

\[ m_{\text{air}} = 1.18 \text{kg/m}^3 \times 34000 \text{m}^3/\text{h} = 40120 \text{kg/h} \times 11.14 \text{kg/s} \]

The specific heat of air is taken as 1.005 kJ/kg. Substituting all the values into equation (4.2) yields:

\[ Q_{\text{air}} = 11.14 \text{kg/s} \times 1.005 \text{kJ/kg} \times (80 - 25) = 616.01 \text{kJ/s} \]

The steam requirement is gotten from equation (4.3) by substituting the values as follows:

\[ m_{\text{st}} = \frac{616.01 \text{kJ/s}}{(2790 - 419.1) \text{kJ/kg}} = 0.2598 \text{kg/s} = 0.94 \text{tpk} \]
The heat gain by the air in raising its temperature from 25 to 80°C was found to be 616 kW. Assuming 90% thermal efficiency, the heat to be expended by an electric heater would be 678kW. If the heating were to be done via an electric heater, it can be achieved by an approximately 700kW rated electric heater.

4.1.1.2 Material challenges in pachucas’ lining

The metallic casing of the Pachucas’ reactor is lined with rubber material. The temperatures of the reactants must be at levels that will not compromise the integrity of this rubber material. However, consideration should be made to replace this rubber lining with one that would withstand higher temperatures, such as steel.

![Electrolyte flow diagram for TLP Tank house. Note: Unit 1 contains 80 cells, Unit 2 contains 160, and Unit 3 contains 80. D1, D2 and D3 are the respective dearation tanks for each unit, each connected to a pump. Line A contains electrolyte from the extraction tank and line B returns spent electrolyte from the electro-winning units to the tank.](image)
4.2 Heating of TLP Tankhouse Electrolyte

The TLP tankhouse uses the process of electro-winning for production of copper cathodes. For effective copper recoveries and production of good quality copper, the electrolyte temperature has to be at optimal levels of \((50 – 60^\circ C)\) and correct mineral composition. Currently, prevailing electrolyte temperatures are at ambient range \((25 – 30^\circ C)\). It is believed that the quality of the produced Kabundi cathode (KBC) would improve if the electrolyte temperatures were elevated and this would reduce the prevalence of cathode rejects.

Figure 4.3 shows the electrolyte circuit in the tankhouse. Electro-winning occurs in the units 1, 2 and 3. Electrolyte continuously circulates around a closed loop between the units, the dearators and the solvent extraction tank (SX). Unit 1 and 3 each contains 80 electro-winning cells while unit 2 is the largest, containing 160 cells. Each unit has an electrolyte circulation pump and a dearator (liquid/vapor separator) \((P1, P2, P3\) and \(D1, D2, D3)\) respectively. The electro-winning process is exothermic in the cells where there is a \(10^\circ C\) temperature gain by the electrolyte on its return (line B) to the solvent extraction tank for replenishing. Electrolyte in line B is at a temperature \(10^\circ C\) higher than in line A.

The desire is to have electrolyte temperature of \(50 - 60^\circ C\) in the electro-winning cells. The challenge is that the electrolyte must return to the solvent extraction tank at a temperature not higher than \(40^\circ C\). This project has the agony therefore, of designing a system that will raise the temperature in the electrolyte feed line (A) to \(50^\circ C\) and extract heat from it in the return line (B) to a temperature not higher than \(40^\circ C\). Due to this challenge, the plant operators would want to raise the temperature in the electrolyte feed line (A) to \(50^\circ C\) which would be raised to maximum \(60^\circ C\) from the exothermic reactions across the electro-winning units.

A heat exchanger in line A is required to raise the temperature from that emergent from the SX to \(50^\circ C\). Another heat exchanger will be required in line B to reduce the electrolyte temperature from the level at which it emerges from the electro-winning units to the required \(40^\circ C\) on return of the spent electrolyte to the advance tank.

Figure 4.4 depicts the placement of the envisaged heat exchangers in both the electrolyte feed and return lines. In the heat exchanger VW, the temperature of electrolyte is raised to \(50^\circ C\). The heating medium \((N)\) is taken as steam from Nchanga smelter. In the heat exchanger XY, the electrolyte has to dump its heat to maximum \(40^\circ C\) to the colder medium \((M)\).
4.2.1 Analysis of applicable heating scenarios

4.2.1.1 Heat exchanger in electrolyte feed line

Noting the availability of steam from smelter, the current task is to determine the quantity of steam required to raise the electrolyte temperature from that at SX exit to 50°C as entry temperature into the electro-winning units.
Figure 4.5 depicts the heat balances across the heat exchanger VW in the electrolyte feed line (line A in figure 4.3). Note the parameters in either medium across the heat exchanger.

![Diagram](image)

Figure 4.5: Heat balance diagram. Steam from the WHRB at smelter is channeled into the heat exchanger for electrolyte heating.

The heat balance equation is as follows:

$$m_{st} (h_{in} - h_{out}) = m_{el} \cdot c_{p, el} (T_{out, el} - T_{in, el})$$

(4.5)

The right hand side of the heat balance equation gives the electrolyte heat gain ($Q_{el}$) as follows:

$$Q_{el} = m_{el} \cdot c_{p, el} (T_{out, el} - T_{in, el})$$

(4.6)

The left hand side of the equation (4.5), combined with equation (4.6) gives the mass flow of steam.

$$m_{st} = \frac{Q_{el}}{(h_{in} - h_{out})}$$

(4.7)

where in the foregoing three equations, $m_{st}$ is the mass flow of steam, $h_{in}$ is the enthalpy of steam at heat exchanger (HE) entry (at 60 bar, 277°C) $h_{out}$ is the enthalpy of steam at HE exit, $m_{el}$ is the mass flow of electrolyte, $c_{p, el}$ is the specific heat of steam.
electrolyte, $T_{\text{out,el}}$ is the electrolyte temperature at HE exit and $T_{\text{in,el}}$ is the electrolyte approach temperature.

Here, $h_{\text{in}} = 2790 \text{ kJ/kg}$, $h_{\text{out}} = 419.1 \text{ kJ/kg}$, $T_{\text{out,el}} = 50^\circ\text{C}$ and $T_{\text{in,el}} = 40^\circ\text{C}$.

The maximum electrolyte flow rate around the closed loop is $1000 \text{ m}^3/\text{h}$. The specific heat $c_{p,\text{el}}$ of the electrolyte is $3.45 \text{ kJ/(kg.K)}$ and has density of $1020 \text{ kg/m}^3$. To simplify computations, the electrolyte flow is converted to kg/s using equation (4.4).

$$m_{\text{el}} = 1020 \text{kg/m}^3 \times 1000 \text{m}^3/\text{h} = 1020,000 \text{kg/h} = 283.33 \text{kg/s}$$

Substituting the values into equation (4.6)

$$Q_{\text{el}} = 283.33 \text{kg/s} \times 3.45 \text{kJ/kg.K} \times (50 - 40) = 9774.89 \text{kJ/kJ/s} = 9.77 \text{MW}$$

The required steam mass flow is calculated from equation (4.7) as follows:

$$m_{st} = \frac{9774.89}{(2790 - 419.1)} = 4.12 \text{kg/s} = 14.84 \text{tph}$$

This is the quantity of steam required at all times (steady-state operations) in the steam/electrolyte heat exchanger since the temperature drop into the SX is a process requirement.

Initial start-up of the heating process over the same period of time as the steady-state case implies raising the electrolyte temperature through $(35 - 50^\circ\text{C})$. In this case, we substitute the respective temperatures in the heat balance equation (4.5). Repeating the foregoing computations yields the required steam mass flow for initial heat-up as $6.18 \text{kg/s}$ or $22.26 \text{tph}$. Note that the initial temperature here is taken as $35^\circ\text{C}$, this is so since the $10^\circ\text{C}$ temperature gain across the electro-winning units (in addition to the $25^\circ\text{C}$) is a continuous process. The real case though is that due to heat losses to the environment the process is considered as occurring at ambient temperatures.

4.2.1.1 Heat exchanger in the electrolyte return line

The heat exchanger in the electrolyte return line (line B in fig. 4.4) has the envisaged function of reducing electrolyte temperature from $60^\circ\text{C}$ to $40^\circ\text{C}$. To dump the excess heat from the electrolyte in line B, this project considers available options of heat exchange media.

It should be noted that there is limitation on the heat exchanger to be installed arising from the second law of thermodynamics. The limiting temperature will be that one to
which the electrolyte requires to be dropped (40°C). This will be the maximum temperature to which the cooling medium can go. This rules-out possibility of utilizing this waste heat to meet other heating demands in the tankhouse. Possible cooling media for use across the heat exchanger on the cold side are considered in the following subsections.

4.1.1.2 Water as heat exchange medium

The analysis will determine the amount of water required to extract the excess heat from the electrolyte in line B and proceed to determine the quantity of steam required to raise the electrolyte temperature by 10°C in line (A), figure (4.4). In the first case, the heat balance equation is as follows:

\[ m_{el}c_{p,el}(T_{in,el} - T_{out,el}) = m_wc_{p,w}(T_{out} - T_{in}) \]  

The left hand side of equation (4.8) gives the heat extracted from the electrolyte re-written in equation (4.9) and equation (4.10) gives the required mass flow of water.

\[ Q_{el} = m_{el}c_{p,el}(T_{in,el} - T_{out,el}) \]  

\[ m_w = \frac{Q_{el}}{c_{p,w}(T_{out} - T_{in})} \]

where in equations (4.8), (4.9) and (4.10), \( T_{in} \) is the entry temperature of water (25°C), \( T_{out} \) is the water exit temperature (here set at 38°C), \( c_{p,el} \) is the specific heat of electrolyte, \( m_{el} \) is the mass flow of electrolyte and \( T_{in,el} \) is the electrolyte HE approach temperature (60°C), and \( T_{out,el} \) is the electrolyte HE exit temperature (40°C). Substituting the values into equations (4.9) and (4.10) gives the heat extraction and the mass flow of water respectively as follows:

\[ Q_{el} = 283.33kg/s \times 3.45kJ/kg(60 - 40) = 19549.77kg/s = 19.55MW \]

\[ m_w = \frac{19549.77kJ/s}{3.45kJ/kg(38 - 25)} = 435.9kg/kg = 1569tph = 1569m^3/h \]

4.2.3 Heating of water in dip tanks

When the copper cathodes are pulled from the cells after maturity, they are dipped into rectangular tanks to wash away impurities that may be deposited on the sides, such as
copper sulphate crystals. For this purpose, the tankhouse has four dip tanks, each with four electric heaters. The power rating of each individual heating element is 30kW, implying total installed heating capacity of 480kW with the purpose of maintaining water temperature in the dip tanks at 80°C. The plant operators would like to heat-up this water in the dip tanks by using steam from the smelter.

The current task is to conduct thermodynamic analysis and establish the quantity of steam required to displace the currently installed electric heater. It should be noted that the water in the dip tanks is static and is only changed when it has been used for a period of about four days. This will require emersion of the heat transfer surface (steam-carrying coils) into the dip tanks. The calculation to determine the steam mass flow shall be based on the capacity of the already installed electric heaters to be displaced.

The mass heat rate of the water \( \dot{m}_w \) by steam is obtained from equation (4.13). In the equation, \( P_{el} \) is the rated electrical power (480kW), \( c_p \) is the specific heat of water (4.184 kJ/kgK) and \( T_1 \) and \( T_2 \) are the initial and final temperatures of water in the dip tanks, 25 and 80°C respectively.

\[
\dot{m}_w = \frac{P_{el}}{c_p (T_2 - T_1)} \quad (4.13)
\]

Substituting the values yields the following:

\[
\dot{m}_w = \frac{480 \text{kJ/s}}{4.184 \text{kJ/kg} \times (80 - 25)} = 2.09 \text{kg/s}
\]

The heat dissipated across the dip tanks is given by the following:

\[
Q_w = \dot{m}_w c_p (T_2 - T_1) \quad (4.14)
\]

Substituting the values yields the following:

\[
Q_w = 2.09 \text{kg/s} \times 4.184 \text{kJ/kg} \times (80 - 25) = 480.95 \text{kJ/s} = 0.48 \text{MW}
\]

The required mass flow of steam is obtained from the following equation:

\[
\dot{m}_{st} = \frac{Q_w}{(h_{in} - h_{out})} \quad (4.15)
\]

Here \( h_{in} \) and \( h_{out} \) are inlet and exit enthalpy of steam as defined under equation (4.7). Substituting the values yields the following:
\[ m_{st} = \frac{480.95 \text{kJ} / \text{s}}{(2790 - 419.1) \text{kJ} / \text{kg}} = 0.2029 \text{kg} / \text{s} = 0.73 \text{tph} \]

The foregoing result is the combined steam requirement for the four dip tanks. For decision purposes, the steam demand is here rounded-off to 1tph to give allowances to transmission losses, as the actual delivered steam shall be regulated through actuators. The steam demand for each of the four dip tanks is therefore 0.25tph.

### 4.3 Pump capacities

It can be observed in earlier analyses that the pump is a vital element in bringing the fluids to required pressure levels. When pumping a liquid medium, enthalpy change across the pump is negligible since liquid is incompressible and does not cause significant change of temperature and entropy.

The first law of thermodynamics can be defined in terms of equation (4.15) as follows:

\[ P_{out} = m(h_{out} - h_{in}) \]  

(4.16)

where \( h_{in} \) is the inlet enthalpy and \( h_{out} \) is the outlet enthalpy. Equation (4.15) implies that inlet enthalpy is equal to outlet enthalpy. We can reasonably approximate that pumping electrolyte to another pressure level would cause a temperature increase of 0.3°C. The power rating of the pump that would do the work would be as follows:

\[ P_{pump} = m_{el} c_p \Delta T \]  

(4.17)

where \( m_{el} \) is the mass flow of the electrolyte at the TLP, earlier noted as 283.33kg/s, \( c_p \) is the specific heat of the electrolyte (3.45 kJ/kgK) and \( \Delta T \) is the temperature difference (here 0.3K). Substituting the values into equation (4.17) yields the following:

\[ P_{pump} = 283.33 \text{kJ} / \text{kg} \times 3.45 \text{kJ} / \text{kgK} \times 0.3 \text{K} = 293.2 \text{kJ} / \text{s} \]

The above can be rounded off to 300kW, which is the power rating for any pump that would be placed on the electrolyte line.

### 4.5 Associated carbon emissions

Carbon footprint for both Nchanga smelter and TLP is not known as no audit has been done. Here, quantification of carbon emissions will be based on emission factors for competing scenarios. It will be assumed that the entire heating demand at TLP will be met from the Nchanga smelter excess steam generated from the smelter waste heat.
The total heating demand for TLP will be the sum of all the process areas requiring heat. The total heating demand \( Q_{TLP,\text{tot}} \) according to the foregoing sections is computed as follows:

\[
Q_{TLP,\text{tot}} = Q_{\text{pac}} + Q_{\text{el}} + Q_{w}
\]  

(4.18)

where \( Q_{\text{pac}} \) is the heat demand at the Pachucas', \( Q_{\text{el}} \) is the heat demand for electrolyte and \( Q_{w} \) is the heat demand in the water dip tanks. Substituting the value obtained earlier yields the following:

\[
Q_{TLP,\text{tot}} = 616.01 + 9774.89 + 480.95 = 10871.85kJ / s = 10.87MW
\]

To be able to quantify the carbon emission scenarios if the heating installations discussed were implemented, we set the baseline scenarios against which the reductions or increase in emissions shall be calculated.

### 4.5.1 Meeting heating demand via grid electricity

If the 10.87MW heating demand at TLP were to be met via grid electricity, and assuming 100 percent operation around the year, the associated emissions can be calculated using equation (3.15) as follows:

\[
CO_{2,\text{emissions}} = 8760\text{hrs} \times 10.87\text{MW} \times 0.9996tCO_{2} / \text{MWh} = 95,183tCO_{2} / \text{year}
\]

Selecting grid electricity to meet the TLP heating demand would therefore lead to 95,183 tCO\(_2\)/year emission reductions, which is the carbon trading potential.

### 4.5.2 Meeting heating demand via HFO fired boiler

The carbon emissions associated to the scenario where an HFO fired boiler is used to generate steam and meet the heating demand can be calculated from equation (3.18) with consistency of units to yield the following:

\[
CO_{2,HFO} = 8760h \times 10871.85kJ / s \times 77.4tCO_{2} / TJ = 26,537tCO_{2} / \text{year}
\]
CHAPTER 5

NKANA REFINERY

5.1 Overview of operations and energy scenarios

The Nkana refinery produces Rokana electro-refined copper (REC) cathodes through the refinery tankhouse. The refinery processes blister anode copper from the Nchanga smelter to produce copper cathodes ready for the market. The plant production capacity is 300 tons per annum. To produce the copper cathode to London metal exchange (LME) standards, the electrolyte must be at optimum conditions of temperature, mineral composition, and correct quantity of reagents. Optimum temperature for the electrolyte is 63±5 °C. The maximum electrolyte flow rate in the electro-refining sections is 300m³/hr. Note that information and data presented in this section was obtained from interaction with plant operators, induction and operators’ training manuals [55][56].

The refinery tankhouse consists of 72 sections containing 1936 cells in total. Each section consists of 26 cells, except for a few giant sections that are made up of 34 cells each. The sections are grouped into nine heating circuits which are treated distinctively. Of the 72 sections, nine are dedicated to producing starter sheets while the rest produce copper cathodes. Starter sheets are a thinner copper plate which is inserted into the electro-refining cells for growing of copper cathodes by deposition (electroplating).

The prevailing scenario is that each of the nine heating circuits has an electric heating system known as the direct electric heater (DEH). Each DEH has a power rating of 2.0MW, implying combined electric heating capacity of 18.0 MW. Each of these heating circuits has a steam heat exchanger attached to it on standby, operated when the electric system is down.

Figure 5.1 shows the electrolyte circuit in the refinery tankhouse. The main electrolyte flow is between the electro-refining cells and the header tanks. Each of the nine heating circuits has its own header tank. Addition of sulphuric acid, re-agents and other electrolyte constituents are added to the electrolyte through the header tanks. Temperature maintenance is done via another close-loop circuit where electrolyte flows between the header tanks and the DEH. The total electrolyte flow into the header is 600 m³/hr and the temperature difference between the incoming electrolyte from the cells and the header tank is 1.5°C. The temperature in the return line from the cells to the header tank is 59.4°C. The density of the electrolyte is 1230 kg/m³ and the specific heat is 3.643 kJ/kg. Before the DEHs were installed, steam that was used in the refinery heat exchangers was generated from the waste heat boiler at the Nkana smelter (decommissioned in 2008).
Recently, management desired to install a boiler to generate steam for the refinery heat exchangers to replace the electric heating system. To participate to the global agenda of climate change mitigation, management considered installation of a biomass fired boiler for possible utilization of the locally generated sawdust and other waste from timber processing plants. The undertaking was to be considered as additional to the company’s portfolio of CDM projects.

The project team gave 20tph as the total steam requirement for this purpose. This steam would be generated at a temperature of 206°C and pressure of 17 bars. The feed water temperature into the boiler would be 60 °C.

Fig. 5.1: System outlook for the electrolyte loop in the Refinery tankhouse. In the outer loop, electrolyte flows between the header tanks and the electro-refining circuits. The inner circuit shows electrolyte flow between the header tank and the heat exchanger. Note the two routes for electrolyte heating. Electrolyte heating is done either by the DEH or the steam heat exchanger (HE).
This thesis project takes a holistic view of the refinery electrolyte heating demand and will provide analysis that simplifies data and information that could be used as guiding tool for decision making. Furthermore the refinery tankhouse has dip tanks for washing of copper cathodes when they are pulled from the sections to wash away light deposits of crystals, such as copper sulphate. The combined volume water change for the dip tanks in the tankhouse is 57.024 m$^3$/day.

### 5.2 Refinery electrolyte heating

#### 5.2.1 Energy and Fuel analysis

This analysis shall conduct its calculation based on the parameters of steam as given by the project team. The current task is to determine the quantities of fuel required for boiler firing to meet the electrolyte heating demand. Recall that the steam mass flow was given as 20 tph which is to be distributed between the nine heating circuits. The calculation will be conducted on the combined effect of the heating circuits. Figure 5.2 depicts the two media exchanging heat across each of the nine heat exchangers and the heat balances across each, leading to the heat balance equation (5.1).

![Heat balances across the respective heat exchangers in the Nkana refinery tankhouse.](image)

**Fig. 5.2:** Heat balances across the respective heat exchangers in the Nkana refinery tankhouse.

Equation (5.2) gives the enthalpy gain by the electrolyte across the heat exchangers.
\[ m_{el} c_{p,el} (T_{in,el} - T_{out,el}) = m_{st} (h_{in} - h_{out}) \]  

(5.1)

\[ Q_{el} = m_{el} c_{p,el} (T_{in,el} - T_{out,el}) \]  

(5.2)

where \( m_{el} \) is the electrolyte mass flow, \( c_{p,el} \) is the specific heat of the electrolyte, \( T_{in,el} \) is the electrolyte temperature on entry into the heat exchanger, \( T_{out,el} \) is the electrolyte exit temperature, \( m_{st} \) is the steam mass flow, \( h_{in} \) is the steam enthalpy on entry into heat exchanger and \( h_{out} \) is the steam exit enthalpy from heat exchanger.

Here \( m_{el} = 300 \) m³/h (with density given as 1030 kg/m³, equation (4.4) gives \( m_{el} = 85.83 \) kg/s), \( c_{p,el} = 3.643 \) kJ/kg and \( (T_{out,el} - T_{in,el}) = 1.5K \). Substituting the values into equation (5.2) yields the enthalpy drop across the heat exchanger as:

\[ Q_{el} = 85.83 kg/s \times 3.643 kJ/kg \times 1.5K = 469 kJ/s = 0.47 MW \]

We consider installing a fluidized bed boiler with combustion efficiency 90% and heat transfer efficiency of 30%. The actual boiler load can be obtained from the foregoing result as follows:

\[ Q_b = Q_{el} (1 + (1 - \eta_c))(1 + (1 - \eta_{th})) \]  

(5.3)

where \( \eta_c \) and \( \eta_{th} \) are the combustion and heat transfer efficiencies, \( Q_b \) is the boiler heat load and \( Q_{el} \) is the heat gain by the electrolyte. Substituting the values into equation (5.3) yields the following as boiler heat load:

\[ Q_b = 469.04 kJ/kg \times 1.1 \times 1.7 = 877.1 kJ/s = 0.88 MW \]

The available fuels for boiler firing are HFO and biomass (sawdust). We proceed to compute the quantity of fuel that would be required in either scenario.

5.2.2 Fuel quantities

5.2.2.1 HFO as boiler fuel

We consider heavy fuel oil (HFO) with default low heating value (LHV) of 42,000 kJ/kg. We obtain the fuel mass flow from equation (3.17) as follows:
\[ m_{HFO} = \frac{877.10 \text{kJ/s}}{42000 \text{kJ/kg}} = 0.02088 \text{kg/s} = 0.0518 \text{tph} \]

The associated annual carbon emissions to the foregoing quantity of HFO consumption can be deduced from equation (3.18), assuming 100% operation follows:

\[ CO_{2,HFO} = (8760 \times 60 \times 60) s \times 877.1 kW \times 77.4 t CO_2 / TJ = 2,140 t CO_2 / \text{year} \]

5.2.2.2 Biomass as boiler fuel

Recall that the available biomass potential was discussed in section 3.5.2.3. We assume the biomass low heating value (LHV) as 15,000 kJ/kg and compute the fuel mass flow from equation (3.17) as follows:

\[ m_{biomass} = \frac{877.10 \text{kJ/s}}{15000 \text{kJ/kg}} = 0.05847 \text{kg/s} = 0.2105 \text{tph} \]

Assuming 100% operation, we compute the associated carbon emission reductions from equation (3.15) with all entries as defined in section 3.5.2.4 as follows:

\[ CO_{2,biomass} = 8760 h \times 0.877 \text{MW} \times 0.9996 t CO_2 / MWh = 7,679 t CO_2 / \text{year} \]

The choice of the fluidized bed combustor emanates from its flexibility to use a wide range of fuels without need for alterations to the installed hardware. The quantities of fuel in the foregoing analysis are for steady-state operation. Initial heat-up of the electrolyte is to be met from the DEHs.
CHAPTER 6
DISCUSSION, RECOMMENDATIONS AND CONCLUSIONS

6.1 Nchanga Smelter

6.1.2 Discussion

An overview of waste heat scenarios of three furnaces, the flash smelting furnace (FSF), slag cleaning furnace (SCF) and the cobalt recovery furnaces (CRFs) was given. Possible installation of heat recovery steam generators (HRSGs) to harness waste heat in the gas streams emerging from each of the furnaces was analyzed. In each case, a thermodynamic analysis was conducted to determine the quantity of energy available in the flue gases and the quantity of steam that can be generated. The results from the analyses gave the relevant parameters that ascertain the power generation potential of the steam from the envisaged HRSGs.

Recall that coupled to each of the electric furnaces is an incinerator that has the function of bringing combustible gases remaining in the gas stream (carbon monoxide and hydrogen) to complete combustion which is achieved by introducing air into the incinerator. As there is need to reduce the temperature of the flue gas, more air is introduced to meet this purpose. The first analysis on the SCF endeavored to establish the actual amount of combustion air is actually required to bring the gases to complete combustion. This was done as the envisaged installation would require capture of all the heat in the flue gases where the HRSG would act as the cooling vessel for the flue gases.

The analysis revealed that the SCF gas stream contains 13.08MW of heat and that installation of an HRSG to capture this waste heat would result into generating superheated steam at 420°C temperature, 60 bars pressure and 4.66kg/s mass flow. Installation of a steam turbine unit to utilize the enthalpy in this steam would result into generating 2.32MW of power. A similar thermodynamic analysis was done on the CRF flue gas where it was revealed that the gas contains 2.49MW of heat. Channeling the CRF flue gas into an HRSG would result into generating 1.77kg/s of steam at the same parameters as in the SCF case. The power generation potential from the steam generated from the CRF flue gas was found to be 0.88MW.

The research noted that the highest single volume of flue gases from the furnaces was that from the flash smelting furnace. A thermodynamic analysis of the flue gases revealed that the energy contained therein is 21.74MW. Channeling this flue gas stream into an HRSG would generate steam at 420°C temperature, 60 bar pressure and 7.75kg/s mass flow. The steam has a power generation potential of 2.99MW through a steam turbine unit. Besides the 2.99 MW power, the steam generated would still suffice for the other operations in the current scenario. Analyses on the other waste heat points
such as the anode furnaces and the sulphuric acid plant were not conducted due to non-availability of data at the time of the research. The combined power generation potential from the three furnaces analyzed from installation of HRSGs and steam turbine units was found to be 6.19MW. The selected emission baseline scenarios in the analyses gave a combined carbon trading potential of 55,843 tCO$_2$/year (potential CERs).

The other possible installation that has been discussed in the quest to maximize usage of waste heat associated with the FSF was installation of a separate boiler to superheat the steam from the currently installed WHRB. HFO and biomass were the two fuels that were analyzed for possible usage in firing this boiler. The analysis took account of the other needs and current utilization of part of the steam that is generated from the FSF. The computations were conducted only on the portion of steam that is by design regarded as excess. Recall that the excess steam mass flow was established to be 27tph (7.5kg/s). This was the steam that was envisaged to be channeled into the superheat boiler and for power generation. The boiler heat load was deduced to be 6.11MW. To attain superheat temperature of 420°C and pressure of 60 bars, it was established that the associated HFO consumption would be 0.524tph. Similarly, the biomass fuel consumption was established to be 1.47tph.

The analysis revealed that if the superheat boiler were to be fired by HFO, the associated emissions in this case would be 13,284 tCO$_2$/year. Similarly, carbon emissions analysis for a biomass-fired boiler revealed that its installation would culminate into 34,390 tCO$_2$/year emission reduction if grid electricity were taken as baseline, and of course 13,284 tCO$_2$/year if an HFO-fired boiler were considered as baseline scenario.

6.1.3 Recommendations

Based on the findings in the analysis conducted on the units considered, this project recommends that a single HRSG be installed into which all flue gases from the SCF, CRF and anode furnaces should be channeled. Installing a single HRSG would be less expensive than installing as many incinerators or HRSGs per furnace. The steam generated should be channeled into a steam turbine unit for power generation. Apart from availing locally generated power and hence reducing dependence on grid power, these installations would cut-on operating costs for cooling towers.

As for the FSF flue gas, this project recommends that a separate superheat boiler be installed to superheat the prevailing excess steam for power generation. This option would be cheaper than the alternative of overhauling the currently installed boiler and replacing it with an HRSG. It can be noted from the analysis that power generation potential from the HRSG would be 2.99MW while that from superheated steam through a superheat boiler would be 3.74MW. Consideration of the economics and convenience associated with each of the two scenarios favors installation of a separate superheat
boiler. Furthermore, if the later were considered and biomass selected to be the boiler fuel has the benefits of bringing-in more revenue for the company from the carbon finance associated with the emission reductions.

6.1.4 Conclusion on Nchanga smelter heat and power scenarios

The smelter has vast potential of waste heat which can be harnessed and utilized productively to reduce dependence on grid electricity by more than 15%. A single heat recovery steam generator can be installed to capture the waste heat in flue gases emergent from the electric furnaces to generate steam and run a steam turbine unit. Excess steam in the current scenario from the flash smelting furnace can be superheated through a separate biomass-fired boiler to generate 3.74MW of power. Installation of the recommended waste heat recovery and power generation units has high potential for carbon finance.

6.2 Tailings Leach Plant

6.2.1 Discussion and recommendations for Nchanga TLP

Three areas requiring heating at the TLP were analyzed. Two of these are currently occurring at ambient temperatures. Air from two compressors is fed into the Pachucas' ferric reactor to keep the reactants in suspended form. It is believed that there would be significant improvement to the reactions in the Pachuca's if this compressed air which is currently being fed at ambient temperature were heated. This project analyzed the possibility of utilizing the excess steam from the nearby Nchanga smelter to heat-up this compressed air to 80°C. The objective of the analysis was to establish the quantity of steam required to raise the temperature of compressed air to the required temperature. The analysis revealed that the mass flow of steam required for this purpose is 0.94tph. This translated into an enthalpy gain by the compressed air of 616.01 kJ/s. It was further estimated that if the heating were done using electricity would require a 700kW rated electric heater.

The second process area currently at ambient temperature and requiring heat is the tankhouse electrolyte. Thermodynamic analyses revealed that to elevate the electrolyte temperature through the required profile would require steam mass flow of 14.84tph at its current parameters from the WHRB. In the process, the electrolyte would undergo enthalpy gain of 9.77MW. The electrolyte however offers the major challenge where on its return to the advance tank it must not exceed 40°C temperature to avoid denaturing the organic compounds being replenished in the tank. This project discussed one way of overcoming this challenge by installing a second heat exchanger with water as the cold medium in the electrolyte return line to extract excess heat from it to the required 40°C. The quantity of water required to achieve this purpose was found to be 1569m³/hr.
The third area requiring a steam heat exchanger were the dip tanks containing water used for washing of cathodes. This heating demand is currently met via grid power through electric heaters. The desire is to replace these electric heaters with steam heat exchangers. Analysis in this case revealed that the combined quantity of steam from the smelter for all the four dip tanks is 0.73 tph, rounded-off to 1 tph.

This project established that the combined heating demand for the three process areas analyzed at TLP is 10.87MW. To be able to quantify the carbon trading potential if the heat exchangers discussed were installed, two baseline scenarios were considered. The first one would be to meet this heating demand through the grid, via electric heaters. It was established that this case avails a carbon trading potential of 95,183 tCO$_2$/year. The second baseline scenario considered was to meet the heating demand via an HFO-fired boiler. The second case availed carbon trading potential of 26,537 tCO$_2$/year.

This project recommends installation of heat exchangers to utilize Nchanga smelter excess steam for the TLP operations discussed to improve the process. Apart from improved copper recoveries at TLP, the installations further avails opportunities for the company to generate revenue through the sale of carbon credits.

6.2.3 Conclusion on Nchanga TLP

The combined heat demand to heat-up the processes that require heat but are currently occurring at ambient temperature, and to replace the electric heaters in the dip tanks is 10.87MW. The combined quantity of steam required if this heating demand were met from the Nchanga smelter steam is 17 tph. Current excess steam from the Nchanga smelter WHRB is 27tph by design. The carbon trading potential of these installations can be as high as 95,183 tCO$_2$/year.

6.3 Nkana refinery

6.3.1 Discussion and recommendations

The main heat-demanding process at the Nkana refinery is the electrolyte. Optimum electrolyte temperature is required to be able to produce the copper cathode to required standard of quality. The process requires that there be very minimal or no fluctuations in electrolyte temperature. Currently, electrolyte temperature is maintained within the required range via direct electric heaters (DEH). The desire at Nkana refinery is to replace these electric heaters with steam heaters. The temperature drop between the cells and the replenishing tanks was given as 1.5$^\circ$C. The heating system is required to re-heat the electrolyte on its return from the header tanks to the electrolytic cells. The main task in the treatment of the refinery heating scenario was to establish the quantity of fuel that would be used in a boiler to generate the required steam. The two fuels
considered for possible use as boiler fuel were HFO and biomass. It was established that the electrolyte enthalpy gain across the heat exchangers to compensate for the $1.5^\circ$C temperature drop across the cells is 0.47 MW. Furthermore, the project established that the fuel feed would have to provide boiler power of 0.88MW. The quantity of HFO required to provide this much power was found to be 0.0518 tph with associated carbon emissions of 2140 tCO$_2$/year. The quantity of biomass that would be required if this heating demand were to be met via a biomass-fired boiler would be 0.2105 tph. In this case, grid electricity qualifies to be the best candidate for baseline scenario because of the DEH. Here, the carbon trading potential is 7679 tCO$_2$/year.

**6.4 FINAL DISCUSSION AND CONCLUSION**

An overview of the Nchanga smelter plant has revealed great potential for energy efficiency improvement from vast quantities of waste heat. The broader view taken of the current waste heat scenarios at Nchanga smelter shows the absence of an integrated approach towards energy and carbon management that incorporate all forms of energy (including heat). This project has shown that several options and technologies for productive utilization of smelter waste heat at Nchanga smelter are available for implementation. The several ways and processes by which waste heat at Nchanga smelter within and at the nearby TLP can be utilized have been holistically analyzed. Apart from the installed WHRB, it has been shown that waste heat from other furnaces can also be captured and be rechanneled back into the system for further energy efficiency improvement and for reduction of production cost.

The Nchanga smelter and the TLP provide great opportunity for application of the principles of industrial ecology. Industrial ecology seeks to design the flow of materials, energy, products and processes that lead to attainment of sustainable manufacturing strategies [57]. In industrial ecology, waste output from one production process becomes input to the next. The Nchanga smelter and TLP plants sit adjacent to each other and the two can be taken as players in one industrial ecological loop. The project has demonstrated how energy efficiency improvement measures can be used as a driver for carbon emission reduction undertakings for participation to the carbon trading mechanisms. The project has further shown that energy efficiency improvement undertakings can be implemented without any alteration to the metallurgical processes.
APPENDICES

APPENDIX 1

Production of blister copper

The Nchanga smelter processes raw copper concentrate received from the Nchanga Concentrator into blister copper anode. The smelting process requires vast amount of heat to melt the concentrate for recovery of the output product.

The largely-simplified chemistry of copper smelting (as done at Nchanga Smelter) is as follows:

- The copper ore is dried in a multi-coil steam dryer before being pneumatically fed into the flash smelting furnace
- The concentrate is heated with silicon dioxide (silica) and air/oxygen in the series of furnaces
- The copper (II) ions in the furnace reactions are reduced to copper (I) sulphide
- The iron is converted into iron (II) silicate slag which is removed
- Most of the sulphur in the smelting contents turns into sulphur dioxide gas (which can be captured and used for manufacturing sulphuric acid)
- The main reactions are as in the equations:
  - $2\text{CuFeS}_2 + 2\text{SiO}_2 + 4\text{O}_2 \rightarrow \text{Cu}_2\text{S} + 2\text{FeSiO}_3 + 3\text{SO}_2$
    The copper (II) sulphide produced is converted to copper with a final blast of air
  - $\text{Cu}_2\text{S} + \text{O}_2 \rightarrow 2\text{Cu} + \text{SO}_2$
    The product in the last equation is what is called copper blister, a porous form of copper, about 98 – 99.5% pure. The copper so produced from sulphide ores according to the above chemistry is impure. The blister copper has to be treated further to remove any remaining sulphur (trapped as bubbles of sulphur dioxide in copper – hence the name *blister copper*) and then cast into anodes for refining by electrolysis.
APPENDIX 2

Electrowinning and Electrorefining

The Nchanga TLP produces copper cathode using the electro-winning process while the Nkana Refinery uses the electro-refining process.

Electrowinning is also called electroextraction, it is the electro-deposition of metals from other ores that have been liquefied. A current is passed from an inert anode through a liquid leach solution containing the metal so that the metal is extracted as it is deposited in an electroplating process onto a cathode [58].

Electro-refining is the electrolytic deposition of a metal from an anode source onto a cathode plate. In copper processing, it is used to purify the metal by removing the impurities that remained during smelting.

The anodes consist of unrefined impure metal, and as the current passes through the acidic electrolyte the anodes are corroded into the solution so that the electroplating process deposits refined pure metal onto the cathodes.
Both electro-winning and electro-refining involve deposition of the pure copper onto a metal plate. The electrolysis process is basically the same as in figure A1 below. Note the size of the copper anode as compared to the cathode, in figure A1. The copper cathode is initially thicker but as deposition on the cathode progresses, the anode diminishes in size. The anode sludge is the impurity that falls-off from the anode. The sludge is collected as a cake after each cycle and consists of precious metals and is sold separately.

Cathode reaction: copper(II) ions are deposited as copper according to the reaction

\[ \text{Cu}^{2+} \text{(aq)} + 2e^- \rightarrow \text{Cu} \text{ (s)} \]

Anode reaction: Copper goes into solution as copper(II) ions according to

\[ \text{Cu} \text{ (s)} \rightarrow \text{Cu}^{2+} \text{(aq)} + 2e^- \]

The deposition occurs in an electrolyte of copper(II) sulphate solution, impure copper anodes, and starting sheets of pure copper for growing of the copper cathodes.

The concentration remains at steady-state composition since for every copper ion that is deposited at the cathode, another one goes into solution at the anode. The cathode gets bigger and bigger as more pure copper is deposited and the anode gradually disappears.

All metals in the impure anode are below copper in the electrochemical series and thus there will be no displacement reaction that would result into generation of ions. These impurities in the copper anode fall to the bottom of the cells. At KCM refinery these impurities are called slimes; they contain precious metals such as gold, silver, bismuth, etc. The slimes are collected and dried, for the export market.
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