Using Geothermal Energy for Raffine Heating in Copper Production

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30 ECTS thesis submitted in partial fulfillment of a
Magister Scientiarum degree in Mechanical Engineering

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Abstract

The aim of this work was to study the feasibility of using geothermal energy for heating raffine (raffinate) solution in the process of copper production. Small-scale experiments have indicated that copper extraction levels can be improved significantly by adding heat to the solution. Two thermal energy sources were considered, namely the cooling water sourced from an adjacent geothermal power plant and low-temperature geothermal brine produced in the vicinity of the mine. These two alternatives were considered for Collahuasi copper mine in Chile due to the fact that the option of using geothermal energy for industrial heating was available on site, namely drilling for low temperature fluid and using cooling water from a prospective power plant.

According to the resulting high profits generated from the adaptation of either one of the two proposed process enhancements, it would be advisable to commence drilling in the low-temperature geothermal area as soon as possible. If however, the low-temperature reservoir cannot withstand the proposed utilization, it would be possible to re-evaluate the feasibility of the cooling-water approach after the commencement of the prospective power plant, which would in any case better than abandoning the project entirely.

To Íris

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Abbreviations

CMCC	Compania Minera Cerro Colorado
CMDIC	Compañía Minera Dona Inés de Collahuasi
Cp	Specific heat [kJ/kg-K]
CS	Carbon steel
Cu	Copper
CW	Cooling water
D/d	Diameter [m]
δ	Insulation thickness [mm]
DSC	dissolved silica content [mg/kg]
ER	Evaporation ratio [%]
η	Efficiency [%]
EUR	Euros (currency)
f	friction factor
g	Gravitational force [m^2/s]
GRP	Glass reinforced plastic
h	Enthalpy [kJ/kg]
$\mathbf{h_L}$	Head loss [m]
HL	Heap Leaching
HW	Hot water
HX	Heat exchanger
i_r	Interest rate [%]
IC	Investment cost [EUR]
IM	Insulation model
K	Head loss coefficient
k	Heat transfer coefficient [$W/(m^2K)$]
L	Length [m]
λ	Thermal conductivity [W/m°C]
LT	Low-temperature
$\dot{m}$	Mass flow [kg/s]
μ	kinematic viscosity [m^2/s]
N	Years of evaluation
n	height [mm]
NPV	Net Present Value [EUR]
OC	Operational cost [EUR]
OD	Outside diameter (of a pipe) [m] or [mm]
OI	Operational income [EUR]
ω	Specific humidity [kg/kg]
P/p	Pressure [bar], [kPa] or [Pa]

PE	Polyethylene
POM	Pipeline optimization model
Q/q	Heat transfer [MW_{th}]
R	Thermal resistance [m°K/W]
r	radius [m] or [mm]
Re	Reynolds number
rel	relative humidity [%]
ρ	Density [kg/m^3]
RS	Raffine solution
s	Entropy [kJ/ (kg*K)]
SF	Silica formation
SS	Stainless steel
SSI	Silica saturation index
T/t	Temperature [°C], [K] or [F]
TSF	Tower size facor
v	Velocity [m/s]
$\dot{V}$	Volumetric flow rate [m^3/s]
W	Work [MW] or [kW]
X	Steam quality [%]
Δ	Difference

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1 Introduction

Copper (Cu) is a chemical element, a metal that is relatively easy to shape, mold or stretch and its corrosion resistance along with excellent heat conduction ability and electrical efficiency makes it highly desirable for various applications. Copper is produced worldwide with Chile as the world's largest copper producer (Wikia, 2012) (MBendi, 2012).

The principal aim of this work was to conduct a feasibility study on the use of geothermal energy for the purpose of heating raffine (raffinate) solution (RS) in the process of copper production. Two thermal energy sources have been considered, namely the cooling water sourced from an adjacent geothermal power plant and low-temperature geothermal brine produced in the vicinity of the mine. Collahuasi Copper Mine in Chile was chosen as the case study for this analysis, due to the fact that the option of using geothermal energy for industrial heating was present on site and prospects for a power plant to commence in the area. At Collahuasi mine, copper is leached from crushed ore by "heap leaching", a process that involves spraying raffine on the heaps. Currently, only about 40% of the copper in the heaps is extracted. Small-scale experiments have indicated that extraction levels can be improved significantly by adding heat to the solution, resulting in associated economic benefits.

The first possible source of energy for the raffine heating is the cooling water from a geothermal power plant. The operation of the plant necessitates the use of an appropriate cooling system to condense the steam coming out of the turbine. Typically, cooling water is used for the condensing process. The cooling water heats up as it flows through the condenser and subsequently it must be cooled back down before it can be allowed to re-enter the cooling cycle. The alternative scenario is to consider the use of this excess heat to heat up the raffine solution. Heat capture from cooling water does present a number of challenges. For example, the cooling water, having circulated through the cooling tower, is rich in oxygen. All pipes and auxiliary equipment in contact with the water must therefore be constructed of materials that have excellent corrosion resistance properties. Another challenge is that in order to accomodate the thermal energy load of the raffine, a great quantitiy of cooling water is required due to its relatively low temperature. This is a critical point in the analysis, as despite the on-site availability of the thermal energy source, the overall cooling capacity of the proposed system may not be sufficient to render it a feasible design option.

Another possible source of heat for the proposed raffine heating system is low-temperature geothermal fluid. This type of fluid is typically located within zones of geothermal activity, where temperatures of 50-200°C can be found at depths of 1 km. The option of utilizing such fluid directly in heating the raffine solution is also considered. The temperature of the geothermal fluid is higher than that of the cooling water and therefore in order to accomodate the thermal energy load of the raffine, less quantitiy of low-temperature water is required. In spite of this fact, the cooling range of the low-temperature fluid is further limited by its dissolved silica content. As it cools down, dissolved silica begins to precipitate from the fluid and forms scaling, causing problems in pipes and heat exchangers. Additional constraints stem from the fact that the utilization of low-temperature geothermal fluid requires drilling and the set-up of a system that would be entirely dedicated to the heating of the raffine solution. Moreover, production of low-temperature fluid is only possible if an adequate productive reservoir is located, therefore the inherent risk of not locating one must also be considered.

A feasibility study of using either cooling water or low temperature fluid for the purpose of heating a raffine solution involves various processes which have to be studied. These processes involve; (a) the thermal exchange between the raffine fluid and the heating fluid in a heat exchanger, (b) transportation of the heating fluid and (c) the power plant. The first two are simulated for both approaches, while the last mentioned is solely for the cooling water approach. The processes were simulated by models developed using MATLAB. To begin with the theoretical background of the processes was analysed. Then, necessary assumptions were made. Following the results of the preliminary model results, one implementation of each approach is proposed for the case study at Collahuasi. A detailed cost analysis is carried out for each of the approaches and the net present value of the investments calculated, based on the potential gain of selling the additional copper produced for the sake of the raffine heating. Furthermore, based on results from the feasibility study and the comparison of the two approaches a set of recommendations of which approach should be chosen for the raffine heating project is presented.

2 Background

Heap leaching for gold and silver recovery is fairly similar to copper but instead of a sulfuric acid used to dissolve copper from its ores, a dilute sodium cyanide solution is delivered to the piles, typically by sprinkling or drip irrigation. It has been shown that the addition of heat to the cyanide dissolution process accelerates the chemical reaction (Trexler, Flynn, & Hendrix, 1990). Altogether, 10 producing gold, silver or gold/silver mines in operation were identified in Nevada with geothermal resources on the mine property or in close proximity to the leaching facilities. It was further determined that gold and silver recovery could be enhanced by 5-17 percent in an experiment that simulated heating of the cyanide solution with geothermal brine. An important benefit of using enhanced geothermal heap leaching is that it allows year round operation, independent of weather conditions, but the heap leaching process is not carried out if the temperature of the solution goes below 4°C. This last study was the first research contribution to the utilization of geothermal energy for leaching and currently two mines in Nevada are using geothermal fluids in their heap leaching operations to extract gold and silver from crushed ore: namely Round Mountain Gold and the Florida Canyon Mine. The first mine began operation in 1906 and has since produced over 283 thousand tons of gold. (Hanson, 2006). At the mine almost 95,000 ton of ore were loaded on the heap (2001). The mine uses geothermal fluids from two shallow wells (~300 m), producing 82°C of geothermal hot water with an average flow rate close to 70 liters per second. The heat from the geothermal fluid is transferred to the cyanide leach solution by a counter flow heat exchanger. The geothermal fluid exits the heat exchanger at 27°C and is re-injected in its entirety back to the geothermal reservoir through a 322m deep injection well located 1.2 km north-northwest of the production well site. Annual averages of energy usage near 42 TJ and the system has an installed capacity of 14.1 MW_{th} (Lund, 2003). At the Florida Canyon mine close to 13,000 tons of ore are loaded onto the heap daily. Currently the mine produces 2.5 kg of gold per day, resulting in annual production levels of 905 kg of gold and almost 800 kg of silver (Driesner & Coyner, 2007). Geothermal fluids are produced and transported from a 177 m deep well to a tube-and-shell heat exchanger, where heat is transferred to the barren cyanide solution used in the heap leaching process. The operating temperature is higher than the one at Round Mountain mine, at 99°C with the maximum temperature in the well reaching 114°C, which has been associated to frequent calcium carbonate precipitation problems in the production well (Trexler, Flynn, & Hendrix, 1990). Annual energy usage from the geothermal fluid is estimated to 42 TJ and the total installed capacity is 1.4 MW_{th} (Lund, 2003). The similarity of the two projects to the proposed utilization of geothermal low-temperature fluids at the Collahuasi site for raffine solution heating are evident, however different geological settings, geothermal activity, fluid chemistry and temperature differentiate these projects. As mentioned above, the raffine solution used for copper heap leaching is also different from the cyanide solution used in these projects. The raffine solution is mostly a mixture of water and sulphuric acid, with slight amounts of some auxiliary chemicals such as copper, iron, manganese and chlorine. The pH of this solution is estimated between 1.5 and 2, i.e. the solution is extremely corrosive and heating this fluid will present considerable challenges. The primary challenge therefore lays in selecting a suitable material for the heat exchangers.

In 1995, a raffine heating system for the Compania Minera Cerro Colorado (CMCC) copper mine was installed. The mine is situated only a two hour drive east of Iquique, Chile, at an altitude of 2,500 m, only about 100 km away from Collahuasi. The task there was to heat an acidic copper-barren solution used in the extraction process at the solvent extraction plant from 15°C up to 35°C, prior to re-circulating the solution to the heap pads. The objective was to improve the kinetics of the leaching process. Geothermal fluids were not considered as a prospective energy source for this heating system. The primary source of energy for the mine is diesel-oil and propane was available on site in small quantities during the planning stage of the project. As a result, a 3.93 MW_{th} submerged combustion system for heating of the solution was installed using diesel-oil as the main fuel. The biggest challenge was sourcing proper materials of the system components and metallurgical tests were conducted for this purpose. Carpenter 20Cb-3 alloy was selected for all combustion chamber components. The tank, vent stack and piping were made of fibreglass using an acid resistant Derakane 411-45 resin, Durco valves were chosen to withstand the corrosive raffine solution, while the pumps were made out of CD4MCu alloy. The system turned out to be greatly successful and was followed by the installation of a second identical raffine heater at the CMCC, commissioned a year later (Inproheat, 2009).

Geothermal energy has never been used on a large scale for the purpose of heating raffine solution in a copper processing mine. The success of these projects serves as a strong indication of the potential of heating the raffine solution at Collahuasi using geothermal. Not only have these projects plans turned out to be a success, but they have also validated that a project can be considered feasible when using oil for the purpose of heating a raffine solution to enhance the heap leaching.

Geothermal energy is defined as the part of earth's heat that can be exploited by man (Dickson & Fanelli, 2004). On average, the temperature gradient in the rock is about 30°C/km, but in areas associated with tectonic plate boundaries it can exceed 200°C at 1 km depth and as a result, these sites or zones are classified as high-temperature (HT) areas. Areas within the zone of geothermal activity whose fluid‘s temperature is below 200°C are classified as low-temperature (LT) areas (Pálmason, 2005). This thermal energy can be accessed through drilling and collecting water or steam from geothermal reservoirs.

Geothermal energy is utilized for electricity production, space heating and in various industrial applications. A successful application requires the existence of a geothermal reservoir able to yield adequate amounts of high-temperature fluids that can be extracted for a long period of time without the risk of jeopardizing the long-term capacity of the field. The type of application that geothermal fluids can be utilized for depends on their temperature and phase. High-pressure steam, sourced directly from dry-steam wells or generated through flashing of high-temperature brine sourced from liquid-dominated wells, can be used directly to generate electricity. The brine can be further flashed to produce steam for a turbine operating at a lower pressure or it can be used to heat a secondary fluid or other direct-use applications. At the end of its use cycle, the brine is typically re-injected back into the geothermal reservoir. A cooling system is needed for operating a power plant in order to condense the steam from the turbine. In most cases cooling water is used in either an open- or closed-loop system. The cooling water heats up as it flows through the condenser and therefore the heat must be extracted from the cooling water, usually in a cooling tower. The thermal energy is lost unless it is captured and put to further use.

Another source of geothermal energy for utilization is low-temperature geothermal fluids; mainly used in district heating, green houses, aquaculture and soil warming, while those at the higher 150-200^{o}C range can be used in industrial applications such as pulp, paper and coal drying. The temperature and the chemical composition of the geothermal fluids are the two primary design factors that determine the selection of an appropriate application.

Chile is situated on the west coast of South-America, where it extends 4270km from its northern border with Peru and Bolivia to Cape Horn in the south. This unique shape and geographical location makes the country highly favorable in terms of geothermal activity, as it lies on the plate boundaries of the Nazca and South-American plates on the so-called "Ring-of-fire". More than 300 locations for potential geothermal activity have been identified throughout the entire country, making Chile the largest undeveloped geothermal province worldwide (Sepúlveda, Lahsen, Dorsch, Palacios, & Bender, 2005) (Lahsen, Muñoz, & Parada, 2010) (Woodhouse, 2011).

3 System Modeling

Simulation of the raffine solution heating, transporting of geothermal fluid and operation of a power plant was conducted, by creating a joint model. The model is divided into three smaller individual models:

- Raffine solution heating model
- Pipeline optimization model
- Power plant model

The heating of the raffine solution was selected as a starting point for the modeling analysis. The pipeline is optimized in terms of cost using NPV (net present value). The power plant was optimized as to achieve the maximum power output and was simulated to create a fixed point for the cooling water (CW) analysis. The cooling range of the cooling tower was fixed to be the same as the raffine heating range. One model was additionally constructed to calculate the minimum temperature of the fluid based on dissolved silica content of the fluid in order to determine the cooling range of the fluid used for heating of a secondary fluid.

All the models can operate individually and therefore can be used and arranged in a bigger model, which involves the whole process. The models all need different input variables to operate and give the solution in an output-array, which than can be used as an input for the next model. At the end all models are linked and solved together. This allows the user to adjust variables in the model and see the affect it has on the final outcome. The investment cost, operational cost and operational income (revenue) is then calculated based on the results of the models. The net present value of the projects is calculated.

The NPV method is used to evaluate an investment over a period of time. The limitations of using this method is that it uses fixed interest rates but it gives a good estimate of the loss or gains of an investment. For the selection of the most suitable method, the highest project NPV is taken as an indication of profitability. The NPV can be presented as:

$$NPV = \sum_{k=0}^{N} (OI - OC) * \left(\frac{(1 + \mathrm{i_r})^{\mathrm{k}} - 1}{\mathrm{i_r} * (1 + \mathrm{i_r})^{\mathrm{k}}} \right) - IC \tag{3.1}$$

where OI, OC and IC [EUR], is annual operational income, operational cost and investment cost, i_r [%] interest rate and N [-] number of years for evaluation.

3.1 Raffine Solution Heating Model

The first model returns the mass flow of water needed to be transported and used in a heat exchanger to heat up the raffine solution (RS) for a given conditions.

The energy equation is used to describe the thermodynamic process for the heat exchangers (Moran & Shapiro, 2008). Because the RS is not undergoing a phase change a constant specific heat can be assumed. The total rate of heat transfer between the RS and the hot water is therefore:

$$q_{raf} = \dot{m}_{raf} * C_{p_{raf}} \left(T_{raf_{out}} - T_{raf_{in}}\right) \tag{3.2}$$

where $\dot{m}_{raf}$ [kg/s] is mass flow, $C_{p_{raf}}$ [kJ/kg-K] specific heat, $T_{raf_{out}}$ and $T_{raf_{in}}$[°C] temperatures of the RS, exiting and entering the heat exchanger. Using q_{raf} it is possible to determine the amount of hot water needed for the heating of the RS:

$$\dot{m}_{hw} = \frac{q_{raf}}{\left(h_{hw_{in}} - h_{hw_{out}}\right)} \tag{3.3}$$

where $h_{hw_{in}}$ and $h_{hw_{out}}$[kJ/kg] is the enthalpy of the hot water entering and exiting the heat exchanger. The purpose of the raffine heating model is simply to return the mass flow of water required for the raffine heating based on the cooling range of the heating fluid.

3.2 Pipeline Optimization Model

For simulating the transportation of the fluids a pipeline optimization model was created. It selects the optimal pipe material, thicknesses and diameter, with regards to cost, using inputs such as; pipe length, elevation, mass flow and temperature. The modeling is based on finding the optimal diameter with regards to NPV (equation 3.1) therefore the installation and operational cost of both pumps and pipeline is calculated along with pressure drop in the pipe. To compute the power needs for pumping the pressure drop in the pipeline transporting the fluid needs to be calculated. The pressure between two points is given by (White, 2008):

$$\mathrm{P_B} - \mathrm{P_A} = \rho \mathrm{g}(\mathrm{z_A} - \mathrm{z_B} - \mathbf{h_L}) \tag{3.4}$$

Where P_B and P_A [bar] is the pressure and $z_A$and z_B [m] are the elevation at points A and B, respectively; g [m/s^2] is the gravitational force and ρ [kg/m^3] is the density of the water. The head loss,h_L [m], is the combination of friction head ($h_{L_{fric}}$) and the local head ($h_{L_{loc}}$), (White, 2008):

$$\mathbf{h_L} = \sum \mathbf{h}_{\mathbf{L}_{\mathrm{fric}}} + \sum \mathbf{h}_{\mathbf{L}_{\mathrm{loc}}} \tag{3.5}$$

The friction head is the mechanical energy that gets converted into heat due to the contact the fluid has with the inner surface of the pipe given by (White, 2008):

$$\mathbf{h}_{\mathbf{L}_{\mathrm{friction}}} = f \frac{\mathrm{Lv}^2}{2\mathrm{D_{in}g}} \tag{3.6}$$

Where f [-] is the friction factor, L [m] is the total length of the pipe, D_{in} [m] is the inner diameter of the pipe and v [m/s] is the velocity of the fluid. To calculate the friction head first the velocity is calculated based on the flow rate and the inner diameter of the pipe, (White, 2008):

$$v = \frac{\dot{V}}{\frac{\pi D_{in}^2}{4}} \tag{3.7}$$

where $\dot{V}$ [m^3/s] is the volumetric fluid flow. In order to determine the friction factor the Reynolds number [-] must be found, (White, 2008):

$$Re = \frac{\rho v D_{in}}{\mu} \tag{3.8}$$

where μ [m^2/s] is the kinematic viscosity of water. Then the Darcy-Weisbach friction factor can be approximated with the Swamee-Jain (Swamee & Jain, 1976) equation which is based on the Colebrook-White equation where ϵ [m] is the absolute roughness of the pipe walls, (White, 2008):

$$f = \frac{0.25}{\left(\log_{10}\left(\frac{\frac{\epsilon}{D_{in}}}{3.7} + \frac{5.74}{R_e^{0.9}}\right)\right)^2} \tag{3.9}$$

The local head is the mechanical energy that is converted into heat when the fluid is flowing through valves, bends, connections, and expansion units, (White, 2008):

$$\mathbf{h_L}_{\text{local}} = K\frac{V^2}{2g} \tag{3.10}$$

as K [-] is the constant combined head loss coefficient.

The work that the motor driving the pump has to deliver is therefore, (White, 2008):

$$W_p = \frac{\dot{m}\rho g(z_A - z_B - \mathbf{h_L})}{\eta_p \eta_{p_m}} \tag{3.11}$$

where $\dot{m}$ [kg/s] is the mass flow of fluid, η_p and η_{p_m} are the pump and motor efficiencies, respectively.

The temperature drop in the pipeline can be presented as (Bjornsson, 1980):

$$T = T_{amb} + (T_{in} - T_{amb})\, e^{-\frac{kL}{\dot{m}c_p}} \tag{3.12}$$

and gives the temperature T [°C] at a distance L [m] from the intake, or as it enters the control volume, where the temperature of the fluid is T_{in}, $\dot{m}$ [kg/s] is the mass flow and c_p [kJ/kg-K] specific heat of the fluid, k [W/(m K)] is the heat transfer coefficient and T_{amb} [°C] is the outside design temperature.

Equation 3.12 is valid for both pipes above ground and buried. What differs is the evaluation of the heat transfer coefficient (k) which is dependent on the thermal resistance

(R) of the materials intervening thermal exchange between the fluid and the surroundings (Bjornsson, 1980):

$$k = \frac{1}{R_{total}} \tag{3.13}$$

R_{total} [m°K/W] is the combined thermal resistance of pipe (R_{pipe}), plastic cover (R_{pe}), insulation (R_{pu}) and also the sand (R_{sand}) and soil (R_{soil}) covering the pipe (Bjornsson, 1980):

$$R_{total} = R_{pipe} + R_{pu} + R_{pe} + R_{soil} + R_{sand} \tag{3.14}$$

where the thermal resistance coefficient for each individual component are calculated as follows (Bjornsson, 1980):

$$R_{pipe} = \left(\frac{2 * \pi * \lambda_{pipe}}{ln\left(\frac{r_p}{r_{p_i}}\right)} \right)^{-1} \tag{3.15}$$

$$R_{pu} = \left(\frac{2 * \pi * \lambda_{insul}}{ln\left(\frac{r_p + \delta}{r_p}\right)} \right)^{-1} \tag{3.16}$$

$$R_{pe} = \left(\frac{2 * \pi * \lambda_{pe}}{ln\left(\frac{r_p + \delta + b_{cover}}{r_p + \delta}\right)} \right)^{-1} \tag{3.17}$$

$$R_{sand} = \left(\frac{2 * \pi * \lambda_{sand}}{ln\left(\frac{r_{PEH} + n_1}{r_{PEH}}\right)} \right)^{-1} \tag{3.18}$$

$$R_{soil} = \left(\frac{2 * \pi * \lambda_{soil}}{ln\left(\frac{2 * (r_1 + n_2)}{r_1}\right)} \right)^{-1} \tag{3.19}$$

where λ_{pipe}, λ_{insul}, λ_{pe}, λ_{sand} and λ_{soil} [W/m°C] are thermal conductivities of the pipe material, insulation, plastic cover, sand and soil, r_p and r_{p_i} [mm] refers to the radius of the pipe up to the outer surface and inner surface, respectively. Insulation thickness is denoted with δ [mm], b_{cover} [mm] thickness of the plastic cover, n_1 and n_2 [mm] height of the sand and soil, respectively. The variable r_{PEH} [mm], is the combined radius of the pipe, insulation and the plastic cover, namely (Bjornsson, 1980):

$$r_{PEH} = r_p + \delta + b_{cover} \tag{3.20}$$

In order to calculate the resistance of the soil the pipe and sand is analyzed as buried cyclone and with r_1[mm] as radius, denoted as *(Bjornsson, 1980):*

$$r_1 = 2 * (r_{PEH} + n_1) \tag{3.21}$$

3.3 Power plant model

In order to determine the feasibility of utilizing the cooling water (CW) for the raffine solution (RS) heating; a power plant model is developed and linked to the RS heating model and pipeline optimization model. Modeling of the power plant is necessary since a power plant has not been initiated at Collahuasi. Modeling the power plant helps in the understanding of how much CW is available for the purpose of heating the RS and the minimum required size of a power plant to provide enough CW. It also highlights the effect the RS heating has on the operation of the power plant that could affect decision making whether CW approach is chosen over low-temperature (LT) approach.

The design of the power plant is optimized to get the maximum net power out of a single flash system (figure 3.1 & 3.2). The high-temperature fluid is a mixture of liquid and steam when exiting a well and is lead to a separator. In the separator the brine is separated from the steam and the steam is lead to the turbine. The brine is usually re-injected into the geothermal reservoir. The steam expands in the turbine, rotating the turbine shaft which is connected to a generator. The steam is then lead to a condenser, where cooling water is used to condense the steam into liquid. Both closed (figure 3.1) and open (figure 3.2) condensers are considered. The CW then returns to a cooling tower where the heat is extracted from the CW and it is then reused in the condenser. The idea is to utilize this CW for heating the RS so special emphasis is on the condensing and cooling water process.

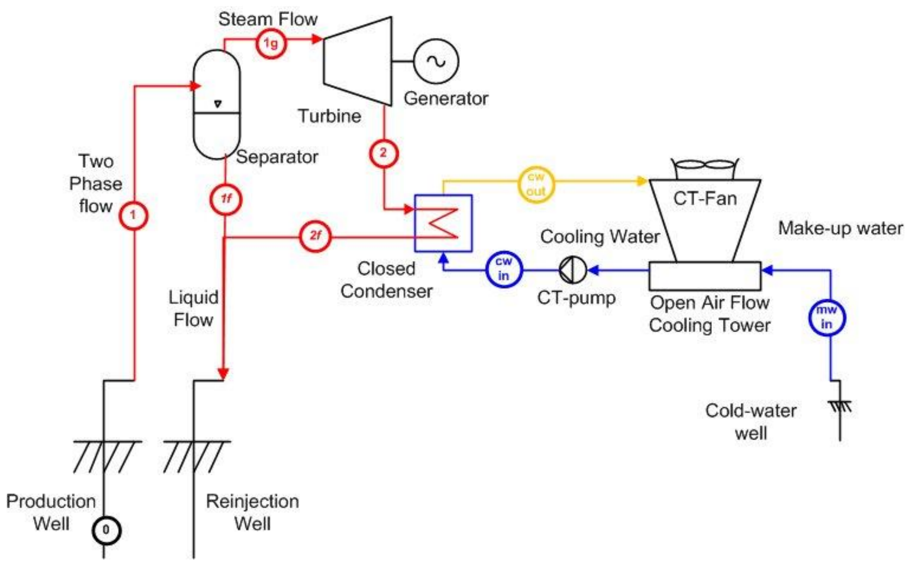


Figure 3.1 Single flash system, closed condenser

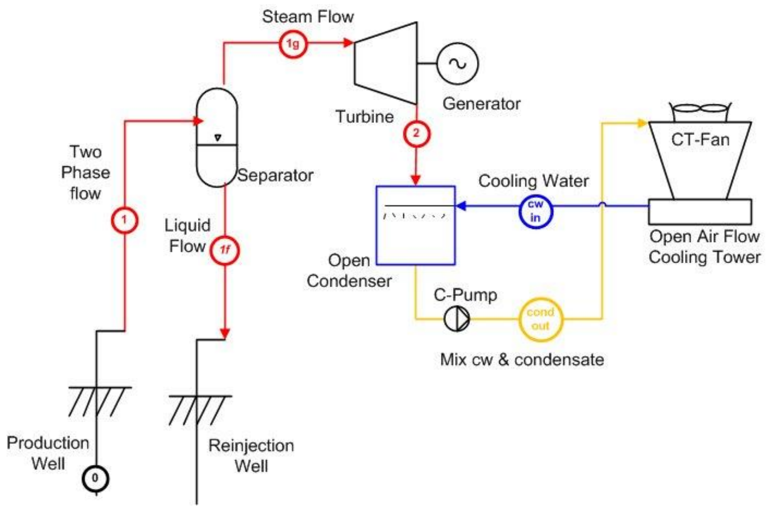


Figure 3.2 Single flash system, open condenser

The flow of fluid through the well is assumed to be isenthalpic, thus $h_0 = h_1$ (DiPippo, 2008), where h_0 and h_1 [kJ/kg] are the enthalpy of the saturated fluid at the bottom of the well, and wellhead, respectively (point 0 & 1, figure 3.1 & 3.2).

The transportation of the mixture from the wellhead to the separator is assumed to be an isobaric process, so the wellhead pressure is the operating pressure of the separator. The total amount of steam in this mixture by weight is denoted with X [-] and can be calculated using this formula (DiPippo, 2008):

$$X_1 = \frac{h_1 - h_{1_f}}{h_{1_g} - h_{1_f}} \tag{3.22}$$

where h_{1_f} and h_{1_g} [kJ/kg] is the enthalpy of saturated fluid and saturated steam at the separation pressure. The steam then flows to the turbine and expands there.

The amount of dry steam entering the turbine can be found by (DiPippo, 2008):

$$\dot{m}_{1_g} = X_1 * m_0 \tag{3.23}$$

where $\dot{m}_{1_g}$ [kg/s] is the mass flow of steam exiting the separator and entering the turbine. To calculate the actual power output the isentropic expansion process is first assumed, with $s_{1_g} = s_{2_s}$ where $s_{1_g}$and s_{2_s} [kJ/ (kg*K)] are the entropy values. The isentropic turbine efficiency is the ratio between the actual (W_t) and isentropic (W_s) work, given in MW_e (DiPippo, 2008):

$$\eta_t = \frac{W_t}{W_s} = \frac{h_{1_g} - h_2}{h_{1_g} - h_{2_s}} \tag{3.24}$$

where h_{1_g} [kJ/kg] is the enthalpy of saturated steam entering the turbine, h_2 and h_{2_s} [kJ/kg] the enthalpy of steam/liquid exiting the turbine in actual (h_2) and ideal (h_{2_s}) process. As the steam expands in the turbine it exits at a lower pressure stage. The condensing pressure is usually 0.1-0.2 bar_a, but it is kept below atmospheric pressure to enhance the power output of the turbine. Part of the steam saturates as it expands in the turbine, but the steam fraction should not go below 85% at the turbine exit to minimize the risk of damaging the turbine blades. In order to calculate the actual output, the enthalpy of the ideal process is first calculated (DiPippo, 2008):

$$h_{2_s} = h_{2_f} + \left(h_{2_g} - h_{2_f}\right) * X_{2_s} \tag{3.25}$$

where h_{2_g} and h_{2_f} [kJ/kg] are enthalpies of the saturated steam and liquid at the 2nd pressure step in the cycle, namely the operating pressure of the condenser. The quality of the steam, $X_{2_s}$is computed with (DiPippo, 2008):

$$\left(\frac{s_{1_g} - s_{2_f}}{s_{2_g} - s_{2_f}}\right) = X_{2_s} \tag{3.26}$$

using the entropy at the turbine inlet s_{1_g} and the entropies at the saturated liquid and saturated steam at the 2nd pressure step namely, s_{1_f} and s_{1_g}, respectively.

Next, the enthalpy from the actual process can be calculated using the ideal enthalpy and the pre-determined isentropic turbine efficiency, η_t [%]:

$$h_2 = h_{1_g} - \eta_t(h_{1_g} - h_{2_s}) \tag{3.27}$$

Finally the actual power output from the turbine is:

$$\mathrm{W_t} = \dot{m}_{1_g}\left(h_{1_g} - h_2\right) \tag{3.28}$$

In the condenser the steam flowing from the turbine is condensed into liquid. The thermal energy that is extracted during this process can be calculated using:

$$Q_c = \dot{m}_2\left(h_2 - h_{2_f}\right) \tag{3.29}$$

where $\dot{m}_2 = \dot{m}_{1_g}$[kg/s] is the mass flow of steam flowing through the turbine. For condensing the steam in an open-condenser, cooling water is sprayed on the steam and a mixture of condensate and cooling water is drained from the condenser. The mass balance equation from this process is as follows:

$$\dot{m}_2 + \dot{m}_{cw_{in}} = \dot{m}_{c_{out}} \tag{3.30}$$

with $\dot{m}_{cw_{in}}$ [kg/s] as the mass flow of the cooling water used in the condensing process, $\dot{m}_2$ [kg/s] the mass flow of steam being condensed, and $\dot{m}_{c_{out}}$ [kg/s] the saturated mixture.

The energy balance is for the same process is:

$$h_2\dot{m}_2 + h_{cw_{in}}\dot{m}_{cw_{in}} = h_{2_f}\dot{m}_{c_{out}} \tag{3.31}$$

where $h_{cw_{in}}$[kJ/kg] is the enthalpy of the cooling water and h_{2_f} [kJ/kg] the enthalpy of the mixture, which is simply the enthalpy of a saturated liquid at the condensing pressure.

For condensing in a closed-condenser, no mixing occurs between the condensate and the cooling water from the cooling tower, so the energy balance is calculated as for a heat exchanger:

$$\dot{m}_2\left(h_2 - h_{2_f}\right) = \dot{m}_{cw_{in}}\left(h_{cw_{in}} - h_{cw_{out}}\right) \tag{3.32}$$

where $h_{cw_{in}}$ and $h_{cw_{out}}$ [kJ/kg] are the enthalpies of the cooling water entering and exiting the condenser, respectively and the $\dot{m}_{cw_{in}}$ [kg/s] the mass flow. h_2 and h_{2_f} [kJ/kg] are the enthalpies of the steam entering the condenser and exiting the condenser as a saturated liquid, with the mass flow denoted by $\dot{m}_2$ [kg/s].

For the condensing process, cooling water (CW) is used to condense the steam coming from the turbine. The CW is in a loop circulating from the condenser to a cooling tower were the heat is extracted from the CW and is then lead back to the condenser.

Modeling the power plant is necessary to estimate its minimum size to provide enough cooling water for the heating process. Another purpose of the power plant modeling is to figure out the affect that the operation of the power plant has on the raffine solution (RS) heating, where cooling water is used for heating. If the CW approach is to be chosen, less CW is lead to the towers and it is otherwise used for RS heating. This affects the operation of the power plant in such a way that less power is needed for cooling resulting in increased efficiency of the power plant. With the CW approach in operation less water is lead to the towers and consequently less water evaporates in the towers. This affects the utilization of the geothermal reservoir as less water is lost by evaporation and can therefore be re-injected to the geothermal reservoir. This will have a positive effect of the capacity of the reservoir in the long term and potentially lengthen the reservoir “life-time” as a resource for geothermal energy. For this reason a special cooling tower model was created within the power plant model to compute the mass of water evaporating in the cooling tower and power consumption of the cooling tower

Figure 3.1 illustrates a cooling tower operating with a closed condenser. In that process set-up the mass flow lost out of the system by evaporation is constantly made up by additional makeup water. The condensate in this set-up is directly re-injected into the reservoir. This requires availability of cold water on site but in geothermal power plant applications the condensate can be used as makeup water. This requires the condensate to be mixed with the cooling water and since therefore the mixture is cooled down in the cooling tower. Mixing between the condensate and the cooling water is therefore inevitable. Open condensers are frequently used where mixing of the cooling water and condensate happens inside the condenser. Figure 3.2 illustrates cooling tower operating with an open condenser. The mass flow of evaporation may not exceed the mass flow from the turbine if no makeup water is to be used. The rest, which is not lost due to evaporation or used again for the condensing process, can be re-injected to the geothermal reservoir.

The aim of the cooling tower (CT) model is to figure out the mass flow of water that evaporates in the CT-process. The wet bulb temperature is the key factor for the CT design. Dry bulb temperature and relative humidity are used to compute the wet bulb temperature. The wet bulb and dry bulb temperatures are then used to determine the specific humidity of the air, ω [kg/kg], which can be reviewed in the Appendix.

The mass balance equations for the CT are (Kröger D. G., 2004):

For dry air:

$$\dot{m}_{a_{in}} = \dot{m}_{a_{out}} = \dot{m}_a \tag{3.33}$$

For water:

$$\dot{m}_{w_{in}} = \dot{m}_{w_{out}} + \dot{m}_e \tag{3.34}$$

Where $\dot{m}_{a_{in}}$ and $\dot{m}_{a_{out}}$ are the mass flow of air entering and exiting the cooling tower, respectively. There is a conservation of dry air in the system; hereafter $\dot{m}_a$ will be used for the mass flow of dry air [kg/s]. There is also conservation of water, where $\dot{m}_{w_{in}}$ [kg/s] is the mass flow of water entering CT, $\dot{m}_e$ is the mass flow of water that evaporates in the process and the rest; $\dot{m}_{w_{out}}$ [kg/s] is returned to the condenser.

The mass flow of water evaporating in the CT is determined based on the assumption that the air exiting is saturated with water vapor. This assumption is made for this study based on Merkel's (1925) assumption that the air exiting the CT is saturated air with vapor. It is however possible for the air exiting the CT to be either under-saturated or super-saturated (Kröger D. G., 2004). For such conditions complex partial differential equations solving is required which is a not within the scope of this study.

The evaporation mass flow is therefore (Kröger D. G., 2004):

$$\dot{m}_e = \dot{m}_a(\omega_{out} - \omega_{in}) \tag{3.35}$$

and must be smaller than the mass flow from the turbine, or else makeup water is needed for the condensing process. To compute ω_{out} energy change of the air entering and exiting the CT is used, as it should be equal to the energy change happening in the water in the cooling tower. Thus (Kröger D. G., 2004);

$$\dot{m}_w \Delta h_w = \dot{m}_a \Delta h_a \tag{3.36}$$

or:

$$\dot{m}_w(h_{w_{out}} - h_{w_{in}}) = \dot{m}_a(h_{a_{out}} - h_{a_{in}}) \tag{3.37}$$

Note that during the cooling process dry air exchanges heat and gets heated up. Another assumption is made that the mass flow of air and water is the same, thus:

$$\frac{\dot{m}_w}{\dot{m}_a} = 1 \tag{3.38}$$

this is an assumption made for this case but it can vary for other cases, dependent on the CT design, but it has to be made in order to calculate the mass flow of cooling water that evaporates in the towers. The enthalpy of the water can be calculated as the cooling capacity of the CT is pre-determined and so the enthalpies of water entering ($h_{w_{in}}$) and exiting ($h_{w_{out}}$) the CT is known. The enthalpy of the air entering the CT ($h_{a_{in}}$) can be calculated using:

$$h_{a_{in}} = Cp_a * T_a + \omega_{in} * (h_{lat} + Cp_v * (T_a)) \tag{3.39}$$

Where h_{lat} is the latent heat at 0°C and Cp_a and Cp_v [kJ/ (kg*K)] are specific heat of dry air and vapor at atmospheric conditions.

The enthalpy of air exiting the cooling tower can be found by

$$h_{a_{out}} = h_{a_{in}} + \Delta h_w * \frac{\dot{m}_w}{\dot{m}_a} = Cp_a * T_x + \omega_{out} * (h_{lat} + Cp_v * (T_x)) \tag{3.40}$$

using the assumption that the mass flow of air and water is the same and the energy change is the same, consequently the enthalpy change is the same. Therefore the enthalpy of the dry air exiting the CT, $h_{a_{out}}$[kJ/kg], is the enthalpy of the ambient air ($h_{a_{in}}$) with enthalpy change in water (Δh_w) added. With all these parameters known it is possible to compute the specific humidity ω_{out} [kg/kg] of the air exiting by guessing the temperature value T_x [°C], which is the temperature of the dry air exiting the CT. Each value of T_x gives new values of ω_{out} using same equations as before using the assumption that saturated air exits the CT.

Further it is possible to determine the energy demand of the motor driving the fan in the cooling tower, using the assumption that the mass flow of air and water is the same in the process. This directly affects the operation of the power plant and can therefore have an effect on decision making.

Using the mass flow of air, $\dot{m}_a$ [kg/s] and the density of air, ρ_a [kg/m^3] gives the volumetric flow:

$$\dot{V}_a = \frac{\dot{m}_a}{\rho_a} \tag{3.41}$$

in [m^3/s]. Then the power required from the fan is:

$$W_f = \frac{\dot{V}_a \Delta P_f}{\eta_f} \tag{3.42}$$

where ΔP_f [kPa] is the assumed value of pressure drop for the air travelling through the CT, assumed as a constant value and η_f is the mechanical efficiency of the fan. The power required to drive the motor:

$$W_m = \frac{W_f}{\eta_m} \tag{3.43}$$

The temperature difference between the wet bulb temperature, T_{wb} [°C], and the temperature of the cold water from the tower, $T_{w_{out}}$ [°C], is called the approach temperature, T_{app} [°F]. The size of the cooling tower is determined using T_{app} by computing the tower size factor (TSF) where TSF=1 gives the average size of cooling tower where the approach temperature is 15°F (or 8.3°C). Bigger approach temperature difference results in smaller cooling towers needed and vice versa (Hensley, 1985). This is illustrated in figure 3.3. This is calculated to determine the size of cooling towers needed and has direct affect on the cost analysis.

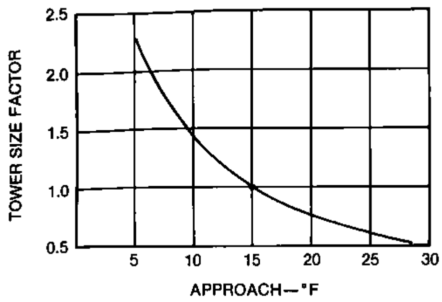


Figure 3.3 Tower Size Factor as a function of approach temperature (Hensley, 1985)

Supplementary model was designed to compute whether and consequently where scaling may occur in the system. The model calculates the minimum temperature that geothermal fluid can be cooled down before scaling occurs. The method was introduced by Stefán Arnórsson (Arnórsson, 1976). Silica forming is a function of the fluid temperature, T_α [K], such:

$$SF\{T_\alpha\}$$

Silica scaling in the fluid happens when the Silica Saturation Index (*SSI*) is equal to one, calculated from the Dissolved Silica Content (*DSC*) and the Silica Formation (*SF*):

$$SSI = \frac{DSC}{SF} \tag{3.44}$$

If *DSC* $\geq$ *SF*; *SSI*$\geq$*1* and silica falls out as scaling, if however the concentration in the fluid is less than *SF* at that temperature all the silica is dissolved in the fluid. Consequently by reducing the temperature of a fluid, still containing the same amount of dissolved silica the *SF* of the fluid changes until *SSI* becomes equal to one. The temperature where this happens is the silica formation temperature of the fluid, and therefore the utilization of the fluid is necessitated by this temperature. The *SF* is a function sole of the temperature of the fluid, T_α[K], and can be computed using:

$$SS_s = 10^{C_0 + C_1 * T_\alpha + C_2 * T_\alpha^2 + C_3 * T_\alpha^3} \tag{3.45}$$

where C_0, C_1, C_2 and C_3 are constants given as:

$$C_0 = -1.34959 \tag{3.46}$$

$$C_1 = 1.625 * 10^{-2} \tag{3.47}$$

$$C_2 = -1.758 * 10^{-5} \tag{3.48}$$

$$C_3 = 5.257 * 10^{-9} \tag{3.49}$$

This summarises the system modeling for both low-temperature and cooling water approaches. In next chapter the models are used to analyze the processes. This helps make one implementation of both approaches for a case study at Collahuasi site using geothermal at Olca.

4 Preliminary Model Results and System Implementation

4.1 Copper processing at Collahuasi, Chile

Chile holds approximately 24% of the known copper reserves in the world, and it is the world's largest copper producer: 5390 thousand metric tons were produced in 2010, accounting for almost one-third of all foreign trade. Copper is the most consistent and major export in Chile, where almost the entire economy is based on mineral export (Wikia, 2012) (MBendi, 2012). On the roots of the Olca volcano lays the Collahuasi Copper Mine. The mine site is located close to the border with Bolivia in the northeast Chile, 180 km southeast of Iquique at an altitude of 4000 m. Collahuasi is the fourth largest copper mine worldwide and it is operated by a joint venture company, Compañía Minera Dona Inés de Collahuasi SCM. The processing area consists of both oxide- and sulphide plants. The oxide ore is crushed by three-stage crushing and copper is leached from the crushed ore by "heap leaching", which involves spraying raffine solution on the heaps. The raffine solution consists of sulphuric acid and water, which percolates through the crushed rocks, partially dissolving them and extracts copper by forming a "pregnant solution". Copper is recovered from the pregnant leach solution by striping with an organic solution in the solvent-extraction plant. The raffine fluid returns to the leach paths and re-enters the process. The end product is copper cathodes and they are produced by electro-winning. In the year 2010, approximately 38,836 tons of copper cathodes were produced in Collahuasi by heap leaching although currently only about 40% of the copper can be recovered from the heaps (Mining-Technology, 2011). There are presently up to 25 heap piles for leaching that cover a surface of about 48 ha (480,000 m^2) reaching a height of 20 m at the mine. During the heap-leaching process, 48,000 m^3 of raffine solution is sprayed over the piles each day in a closed re-circulating system – this corresponds to a flow rate of almost 800 liters per second. The temperature of the circulating fluid is typically between 10 and 15°C but small-scale experiments at the Collahuasi's site have indicated that the extraction ratio of copper can be increased, amounting to 1.2% increase per degree Celsius. The optimal temperature is assumed around 30°C based on the results from the small-scale experiments (Monardes, et al., 2011). The idea behind the heating is that the overall efficiency of the heap leaching process will be directly enhanced, possibly resulting in considerable economic benefits. The concept involves heating the raffine solution as it exits the Solvent-Extraction (SX) plant and prior to it re-entering the circulation cycle, where it is sprayed on the heap piles. This concept is illustrated in figure 4.1. The heat added to the raffine solution is harvested by a heat exchanger (HX). The first possible source of energy for the raffine heating is considered the cooling water from a prospective geothermal power plant, planned on the Collahuasi site, while the second is low-temperature geothermal fluid. Geothermal exploration has begun on the Collahuasi's site as a prospect for utilizing high-temperature fluids for power production. Shallow slim wells have been drilled in an area located 7 km from the mine and they indicate that potential for fluids in temperatures above 70°C at 700 m depth. This area is considered to have potential for low-temperature production and according to conceptual models built from resistivity measurements, the prospective high-temperature zone, suitable for a geothermal power plant, is thought to exists few kilometers further from the mine southeast of the low-temperature zone (Monardes, et al., 2011).

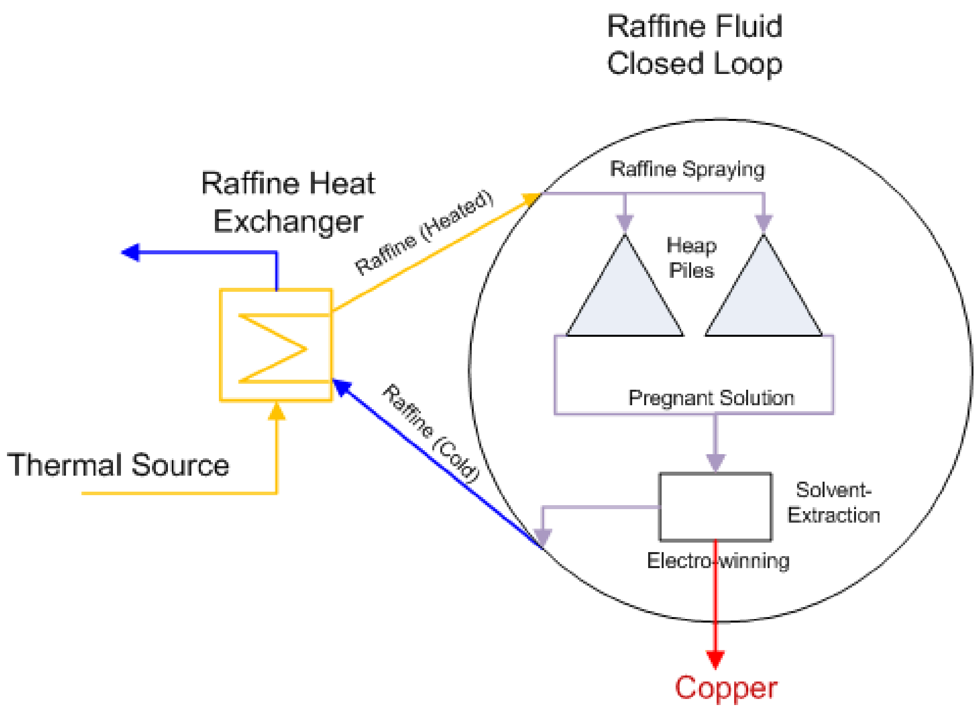


Figure 4.1 The enhanced heap leaching system at Collahuasi

One implementation will be made for the process for utilizing the cooling water for the raffine heating at Collahuasi. This approach will be referred to as the CW-approach. For this study it is assumed that electricity production has initiated at a power plant, which serves the purpose of providing cooling water for the heating process.

Another implementation will be made for the heating process, which involves utilizing low-temperature fluid from a nearby low-temperature area. This approach will be referred to as the LT-approach. The low-temperature area is assumed to provide geothermal fluid of 70°C to use directly for the heating.

4.2 Assumptions for the raffine heating at Collahuasi

The assumptions that were made for the raffine heating at Collahuasi are presented in tables 4.1 through 4.9.

The selection of correct heat exchangers is one of the most critical factors for a reliable analysis of the raffine heating project. The optimal temperature is assumed to be 30°C for the RS as small scale experiments have indicated, but currently the average temperature of the raffine solution in the heap leaching process is 10°C (Monardes, et al., 2011). The temperature difference (ΔT) of the RS in the heat exchanger (HX) is therefore 20°C. The assumed flow requirements of the hot water (HW), temperatures and density is presented in table 4.1, along with the heat exchanger's pinch point, which was selected 10°C to allow sufficient cooling capacity of the heating fluid while keeping the size within suitable range (Verkís, 2012).

Table 4.1 Numerical assumption for raffine solution heating model (Monardes, et al., 2011), (Verkís, 2012)

Optimal temperature of the raffine solution (RS)	30	°C
Current temperature of the RS	10	°C
Flow rate of RS	800	l/s
Density of RS	1.139	kg/m^3
Capacity of each heat exchanger (HX) unit	100	l/s
HX pinch point	10	°C

The cost of each heat exchanger unit depends on the size and material chosen. Based on current knowledge of the chemical composition of the RS and that the similarities of the RS used in CMCC are not known, titanium or other metal with suitable corrosion resistance properties is suggested as material for the HX. Two types of HX are set to be used in the model and presented in table 4.2

Table 4.2 Investment cost assumption for heat exchangers (Schmidt-Bretten, 2011)

Type of heat exchanger	ΔT of HW	Cost of each 100 l/s unit [EUR]
Type A	20	150,000
Type B	40	50,000

Four types of pipes were analyzed for transporting the cooling water and the low-temperature fluid; carbon steel pipes (CS), glass reinforced plastic (GRP) pipes, polyethylene (PE) plastic pipes and stainless steel (SS) pipes.

Pressure, temperature and chemical composition of the fluid being transported are the main design factors for selecting the correct pipe material. It is assumed that the cooling water (CW) has at some point in the system been lead to the cooling tower and been air-rated. This means that the water is sprayed over an up-flow of air and turns to oxygenated fluid which is highly corrosive so that CS pipes are not applicable for transporting cooling water. SS pipes were considered as they can meet the corrosion criteria but were found to be too expensive. Both GRP and PP can meet the design pressure and corrosion criteria but PP pipes are really expensive for large diameter pipes, and the wall thickness becomes very large when designed for high pressure. The GRP pipes were found to be applicable for both transporting the low-temperature fluid from the low-temperature wells and the cooling water at a lower cost than for the PP pipes. Therefore the PP pipes and the SS pipes were neglected from the cost analysis leaving only the GRP pipes and the carbon steel pipes.

The cost of pipes and pumps are presented in table 4.3. The values include all work related to the installation of the pipe, such as groundwork, welding (for the steel pipes) and transportation (Verkís, 2012).

Table 4.3 Cost for pipeline optimization model (Verkís, 2012)

Carbon steel pipe cost (metric price)	0.8	EUR/OD (mm)[1]
GRP pipe cost (metric price)	0.5	EUR/OD (mm)
Carbon steel pump installation cost	0.3	EUR/W
Stainless steel pump installation cost	0.35	EUR/W
Pump's maintenance cost	5	% of Installation cost per yr
Pipeline's maintenance cost	2	% of Installation cost per yr

The carbon steel pipes are pre-insulated but the GRP pipes are without insulation. Therefore the modeling of the required insulation thickness is different with regard to whether the pipe selected is carbon steel or GRP. If the pipe selected is carbon steel the model calculates whether the insulation thickness is enough to accommodate the allowed temperature drop. If the pipe material is however GRP the model calculates the temperature drop for multiple insulation thickness and selects the optimal thickness for the allowed temperature drop. The design must account when and if insulation is required and if so, optimize the thickness regarding to cost without exceeding the allowed temperature drop. A buried pipeline was considered in the analysis. Assumptions made for the thermal conductivity of materials and other design parameters are illustrated in table 4.4.

Table 4.4 Assumptions made for heat loss in pipelines (Bjornsson, 1980), (Set, 2012) , (Verkís, 2012)

Thermal Conductivity		
..for rock wool insulation	0.035	[W/m°C]
..for polyethylene plastic	0.4	[W/m°C]
..for sand	0.3	[W/m°C]
..for soil	1.5	[W/m°C]
..for GRP	0.29	[W/m°C]
..for PB plastic	0.22	[W/m°C]
..for carbon steel	50	[W/m°C]
Price of rock wool insulation	60	[EUR/m3]
Ambient design temperature	-25	[°C]
Average height of sand	150	[mm]
Average height of soil layer	550	[mm]
Plastic cover thickness	2	[mm]

It is assumed that the average well gives ~50 kg/s and the average well temperature is 250°C at the bottom of the well. It is further assumed that the reservoir is liquid dominated. These assumptions are based on current knowledge and speculations about the geothermal

[1] Price per mm of the pipe outside diameter

system based on interpretations of geophysical exploration data and discussions about the potential of the geothermal system at Olca (Monardes, et al., 2011). All assumptions regarding the power plant and its operation are illustrated in table 4.5.

Table 4.5 Assumptions for power plant modeling

Type	Single Flash	
Steam quality		
..bottom of well	0%	(liquid)
..entering turbine	100%	(steam)
..minimum exiting turbine	85%	
Efficiency		
..isentropic, turbine	80%	
..condensate pump	60%	
..fan	85%	
..fan, motor	95%	
Average productivity of high-temperature well	50	kg/s
Temperature		
..bottom of the well	250	°C
..cooling water entering condenser	~20	°C
..cooling water exiting condenser	~40	°C
..cooling water entering cooling tower	~40	°C
..cooling water exiting cooling tower	~20	°C
..ambient Design temp. for the cooling tower	15	°C
..condenser working temp.	~43	°C
Pressure		
Condenser working pressure	~0.09	bar
Ambient pressure	0.6	bar
Relative humidity	60%	

Table 4.6 Geothermal power plant cost estimation (Verkís, 2012)

Steam gathering system		
..Seperator station; inc. Seperators, silencers etc.	0.2	MEUR/MW
..steam piping, inc. Valves, foundations & installation etc.	0.1*Drilling Cost	MEUR
Mechanical- and electrical equipment (IC_{mech})		
..Turbine; including generator, pipes, pumps, valves, lube oil units, installation, electrical equipent etc.	0.6	MEUR/MW
..Condenser; inc. Condensate pumps and piping, gas removal system, ejectors, installation etc.	0.24	MEUR/MW
..Cooling Tower; wet evaporative, inc. Pumps, pipes, fan, overhead service gantry crane, installation etc.	0.24*TSF	MEUR/MW
Buildings		
Buildings; inc.site excavation, grating, fencing, turbine hall, electrical-,service-, personnel, workshop-, storage buildings etc	0.5*IC_{mech}	MEUR/MW

The power plant cost numbers presented in table 4.6 are in the accordance with values used by Verkís consulting engineers based on their experience participating in various power plant projects worldwide (Verkís, 2012). Some modifications were made since this is a new exploration area; The success rate in drilling is assumed 80% (Verkís, 2012) and additional exploration cost is added along with engineering cost that includes designing the power plant, supervision on site etc. The modifications are illustrated in table 4.7. Also, the TSF, which will be calculated in the study, is included in the cost analysis as the size of cooling tower will affect the cost directly. Based on these assumptions a 50 MW power plant with 10 production wells and tower size factor (TSF) equal to 1 will result in an investment cost of over a 150 million Euros (MEUR) for the whole investment.

Table 4.7 Modifications of the power plant cost estimation (Verkís, 2012)

Drilling success rate		80%
3 exploration wells		
Surface exploration inc. Seismic (MT and TEM surveys)		
Environmental Impact Assessment		
Total Exploration Cost	18	MEUR
Engineering design		
Site supervision		
Total Engineering cost	10% of total investment cost	MEUR

The temperature of the low-temperature fluid is assumed 70°C for this study (Monardes, et al., 2011). Samples need to be taken in order to measure the concentration of dissolved silica content in the low temperature fluid, DSC [mg/kg], but for this study it is assumed 100 mg/kg (Verkís, 2012). Drilling production wells, injection wells and a complete system-setup are needed for the low-temperature approach to be initiated for the purpose of heating the raffine fluid. The exploration cost is relatively low compared to the one for the exploration at the high-temperature area, but most of the exploring has already taken place in the low-temperature area. The cost for the low-temperature drilling and system-setup is presented in table 4.8.

Table 4.8 Low-temperature approach cost assumptions

Drilling low-temperature (LT) well	400,000	EUR/each
Drilling injection well, including gathering pipes and equipment on site	400,000	EUR/each
Well pump	75,000	EUR/each
Pipe gathering system on LT field; inc. collector pipes, storage tank, valves, electrical generators, installation etc.	2,000,000	EUR
Drilling success rate		80%
3 exploration wells		
Surface exploration inc. Seismic (MT and TEM surveys)		
Environmental Impact Assessment		
Total Exploration Cost	500000	MEUR
Engineering design		
Site supervision		
Total Engineering cost	10% of Total inv. Cost	MEUR

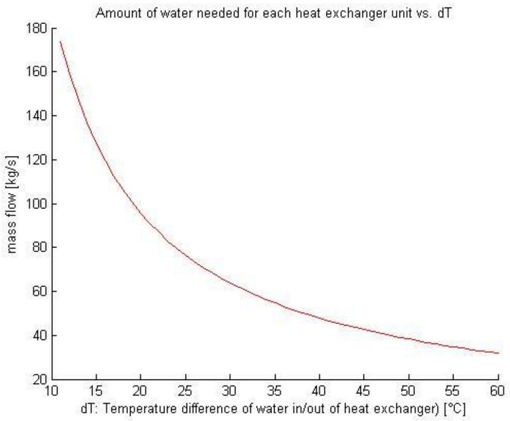
4.3 Preliminary Model Results

The preliminary model results illustrated the following:

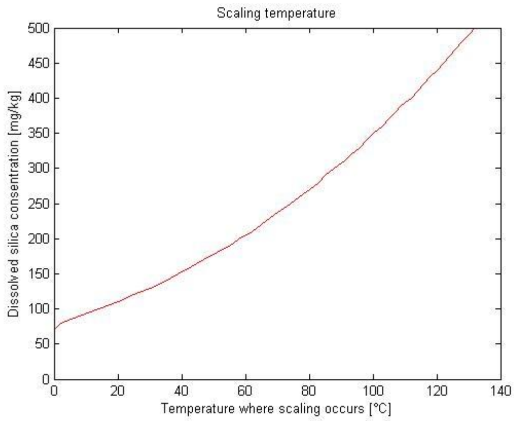
- T (Monardes, et al., 2011) he cooling range of the cooling water and the low-temperature fluid, and based on those results; the required amount of fluid needed for the raffine solution (RS) heating
- How the optimal pipe diameter is selected for same conditions and how declination in landscape and pipeline route lengths affects the cost of pipelines
- Minimum required size of power plant needed to provide enough mass flow of cooling water for the RS heating
- The affect the ambient temperature and relative humidity has on the size of cooling towers and the evaporation of cooling water in the towers

The results from the RS heating model (figure 4.2a) shows the amount of hot water (HW) needed to accommodate the thermal need of the RS for each heat exchanger (HX) unit, based on the cooling range of the HW. As the temperature difference (ΔT) of the HW increases less water is required for the same thermal load. The heap leaching (HL) process demands constant raffine solution of 800 l/s and 10°C to be heated up to 30°C and used in a closed loop system (Monardes, et al., 2011). The pinch point in the heat exchangers is assumed 10°C therefore the ΔT for the CW was selected on pair with the ΔT of the RS, namely from 40°C down to 20°C.

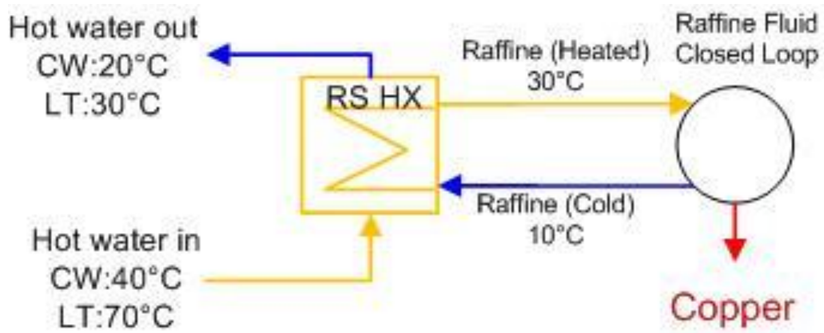
The cooling range for the low-temperature (LT) fluid is assumed to be higher, but the cooling range is necessitated by the temperature of the LT fluid entering the HX and a certain temperature where silica scaling occurs. The fluid temperature is assumed 70°C and the dissolved silica content 100 mg/kg. A special model was constructed in to compute the cooling range for the LT fluid. The temperature where scaling starts to form, with respect of the dissolved silica concentration, is illustrated in figure 4.2b. The model predicts that scaling starts to form in the fluid at 30°C where the dissolved silica concentration is 130 mg/kg. A fluid containing 100 mg/kg is considered without risk of scaling if utilized down to 30°C.



a. Raffine solution heating model preliminary results



b. Silica scaling model preliminary results



c. Selected heat exchanger for cooling water (CW) and low temperature (LT) approach

Figure 4.2 Raffine heating model preliminary results & heat exchanger selection

Both the types of heat exchangers serve 100 l/s of RS on the cold side as the first uses 40°C cooling water (CW) and utilize it down to 20°C while the other receives 70°C low-temperature (LT) fluid and utilizes it down to 30°C (figure 4.2c). The material selection for the HX is critical and it is assumed that the optimal design of the HX in terms of material selection will be made in cooperation with manufactures when the chemical analysis of the raffine solution is available. Several metals were used in the raffine solution heating in at the Minera Cerro Colorado (CMCC) copper mine, however the similarity of the solution at Collahuasi is unknown except the majority of the solution is sulphuric acid and pH is assumed to be around 1.5 - 2. The heat exchanger's specifications are presented in table 4.9.

Table 4.9 Heat exchanger specifications

Approach	Cost type[2]	Temperatures [°C][3]		ṁ / unit [kg/s][4]		# units	ṁ total [kg/s]	
		HW_{in}-HW_{out}	RS_{in}-RS_{out}	RS	HW		RS	HW
Cooling water	A	40-20	10-30	114	96	8	912	766
Low-temperature fluid	B	70-30	10-30	114	48	8	912	383

Based on the preliminary results from the raffine solution heating model, the pipeline optimization model (POM) and the insulation model (IM) are used. The POM is used for 1km pipeline on level ground, both for transporting 766 kg/s of 40°C cooling water (CW) and 383 kg/s of 70°C low-temperature (LT) water. The IM is used for 610 mm outer diameter carbon steel pipe (DN 600) transporting 200 kg/s of 70°C 5km distance. The optimal diameters of the pipes are evaluated by calculating the net present value (NPV) of the pipe and pump according to the assumptions presented in table 4.10.

Table 4.10 Net Present Value (NPV) assumptions

Interest rate	6	%
Depreciation time	25	years
Electricity cost	0.06	EUR/kWh

The NPV of the pipes are displayed on a graph both for Glass Reinforced Plastic (GRP) and carbon steel with respect to diameter. Figures 4.3a displays and compares the cost of transporting 383 kg/s of 70°C LT fluid 1 km using GRP or carbon steel pipes, the blue solid line indicates that GRP pipeline is more feasible option with respect to NPV (net present value) with 610mm optimal outer diameter. Figure 4.3b illustrates the cost of 1 km GRP pipeline transporting 766 kg/s of 40°C CW with respect to outside diameter (OD) of the pipe. DN 900 (914 mm OD) pipe is the optimal selection for these conditions. The GRP CW-pipeline is about 27% more expensive than the one used for LT for the same length and on level ground.

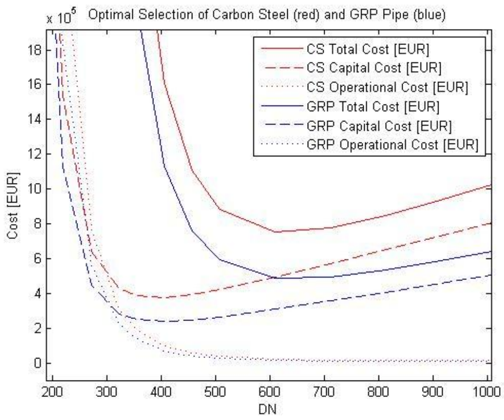
The CW approach involves the CW to be returned to the power plant and re-used in the condenser. This means that the elevations and locations of the production areas are critical in terms of feasibility. The effect the length on the pipelines has on the cost can be observed by for an example reducing the pipeline length of the pipe transporting LT water

[2] According to Chapter 4.2, table 4.2
[3] HW_{in} & HW_{out} ; temperatures of hot water entering and leaving the heat exchanger, respectively, RS_{in} & RS_{out}; temperatures of raffinate solution entering and leaving the heat exchanger, respectively.
[4] $\dot{m}$: mass flow of raffinate solution (RS) and hot water (HW)

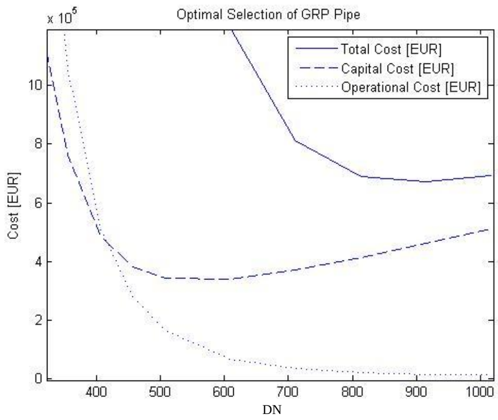
to 0.5km versus 1km CW pipeline, the CW pipeline is 64% more expensive, but setting the LT pipeline length value to 2 km results in the LT pipeline to be 45% more expensive.

To demonstrate the how elevation effects the total cost of the pipelines different declination of the CW route was presented, from zero declination (level ground) to 1% declination. The results are illustrated in figure 4.3c. The elevation difference has a major affect on the total cost. With only 1% declination on 1km, thus 10m elevation drop, pumping is no longer required and so the total cost depends entirely on the pipeline cost as installation cost for pumps is neglected. The reason that the slope illustrating the cost of the cooling water pipeline is not linear is because of the neglecting pump cost and the model selects the next standard pipe diameter.

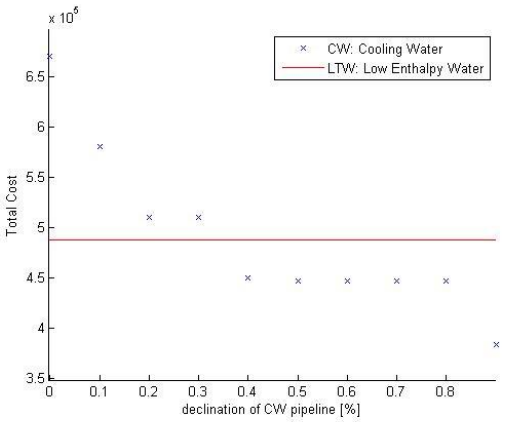
The insulation model was used with 200 kg/s of 70°C travelling 5km and a selected maximum temperature drop of 0.2°C. The minimum insulation thickness for these conditions is 30 mm. This is illustrated in figure 4.3d. This analysis verifies that for buried pipelines the soil and sand is a good insulation, and therefore the temperature drop is limited.



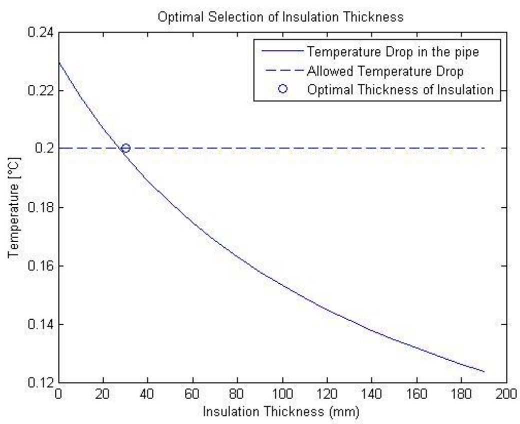
a. 383 kg/s, 70°C; GRP and carbon steel pipeline cost, 1km on level ground



b. 766 kg/s, 40°C; GRP pipeline cost, 1km on level ground



c. Declination affect; GRP pipeline cost

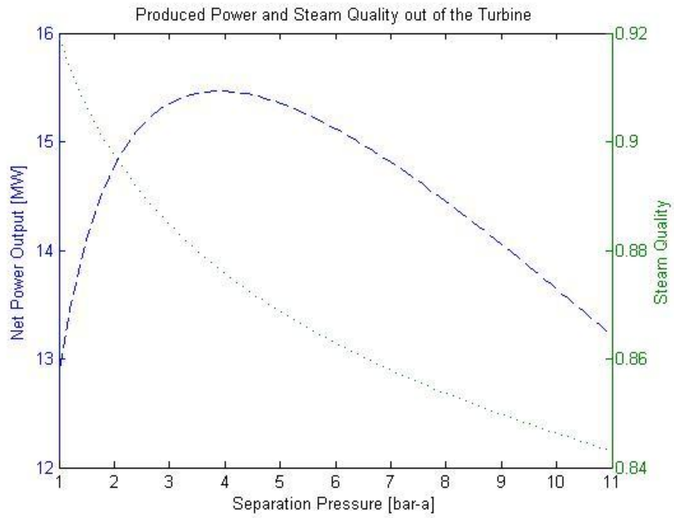


d. Insulation model preliminary results

Figure 4.3 Pipeline optimization model preliminary results

For the CW approach a model was constructed to simulate the power plant, planned on the Collahuasi site. The power plant model seeks to find the optimal electricity generation from given conditions. It also calculates the necessary values in order to determine if utilizing the cooling water (CW) for the raffine solution (RS) heating is possible.

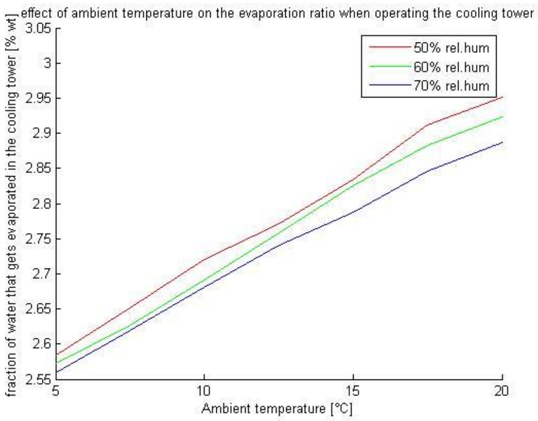
The minimum size of a power plant is 13.8 MW as it will require 766 kg/s for the condensing process, which is equal to the required amount of cooling water needed for the raffine heating. It is assumed that it will need at least 3 wells to achieve the required amount of fluid therefore the graph in figure 4.4 displays the power output from a power plant operating with 3 production wells. It will produce roughly 15 MW of electricity when operating the separator at 3.9 bar_a and provide sufficient mass flow of cooling water for the raffine heating. The graph in figure 4.4 shows the optimal net power output from the power plant when operating the separator at 3.9 bar, it also displays the quality of the steam coming out of the turbine which should be at least 85% when exiting the turbine.



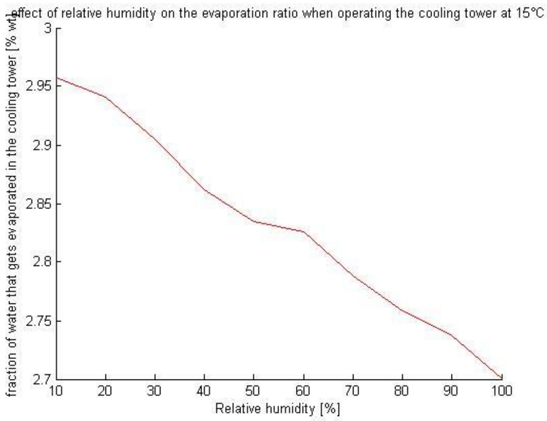
Optimal power output: steam from 3 production wells

Figure 4.4 Power plant model preliminary results

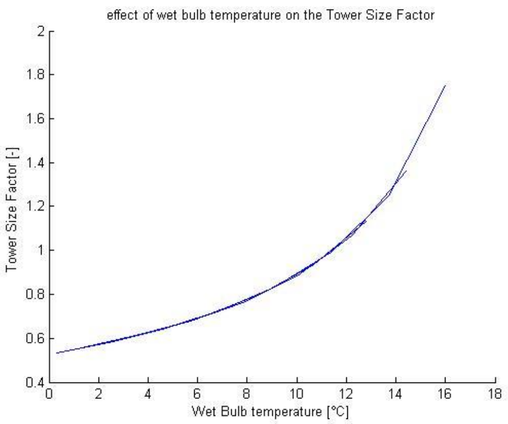
The cooling tower (CT) model was used to find the effects of ambient temperature and relative humidity (RH) on the evaporation ratio (ER) and the size of the cooling tower. The evaporation is the mass flow of water that escapes from the system when the dry air saturates in the CT and evaporates. The ER is therefore the evaporation mass flow divided by the mass flow of water that enters the CT. Mass flow of water by evaporation shall not exceed the mass flow from the turbine. If this happens, not enough water can be transported to the raffinery and additional makeup water is needed for the condensing process, but water is a very limited resource in the area. The size of the cooling tower is determined with the tower size factor (TSF) where TSF=1 is the average size of cooling tower for the given conditions.



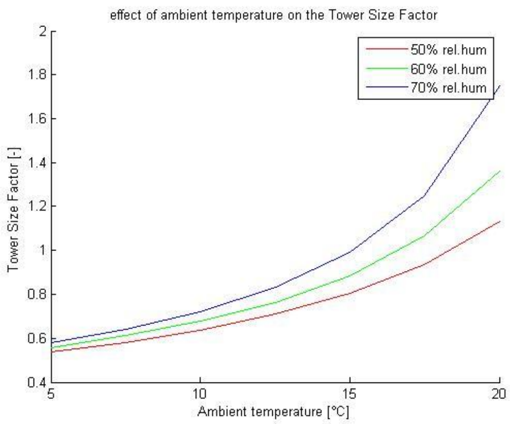
a. Evaporation ratio (ER) vs. ambient temperature, T_a



b. ER vs. relative humidity (rel)



c. Tower size factor (TSF) vs. T_{wb}



d. TSF vs. T_{amb}

Figure 4.5 Cooling tower model preliminary results

The affect the ambient temperature and relative humidity have on the evaporation ratio was investigated (figure 4.5a). The evaporation ratio decreases with increasing relative humidity of dry air (figure 4.5b), but it is based on assumption that the air leaving is saturated with water vapor and the mass flow of air is equal to the mass flow of water (Kröger D. G., 2004).

The cooling tower size is dependent on the wet bulb temperature (figure 4.5c) and the results from the model can be used directly in the cost estimation (Hensley, 1985). The affect the ambient temperature and relative humidity has on the tower size factor was investigated. Bigger towers are needed (or greater number of CT units) for higher relative humidity and the size increases with increasing ambient temperature (Figure 4.5d).

4.4 Pipeline Route Selection

The possible location of power plant and location of the low-temperature (LT) area has been indicated. For this study the power plant location has been selected at the lowest elevation point at the northwest boundary of the pre-determined area. Further the optimal route for the pipeline was selected using Geocontext-topographic profiler and information about the site. The starting point for the LT pipeline was selected in the centre of the low-temperature (LT) drilling area.

In the Appendix, drawings illustrate the optimal pipeline route chosen from location A to B. Location A is the starting point for the pipeline; power plant or centre of the LT drilling area but Location B is the raffinery where the raffine solution (RS) heating will take place. Accordingly the cooling water (CW) pipeline is 9.1 km with 50 m increase in elevation from point A to B. The LT pipeline is on the other hand 8.2 km with 40 m decrease in elevation from point A to B. The return of the CW is back to location A, but the return of the LT water is to a re-injection site 4 km from location A. The location of the injection site was selected as it is assumed that the fluid can re-enter the geothermal system from this location, yet close enough to utilize the negative elevation in the pipeline. The same route is used for return in both cases. Total length of the CW is thus 18.2 km but for the LT approach it is 12.2 km.

4.5 Cooling Water Approach

The cooling water (CW) approach is an alternative for heating the raffine solution (RS) using CW from the power plant. The implementation for this approach is based on model preliminary results and assumptions made for the power plant's operation (table 4.5).

The implementation involves a roughly 50 MW_e power plant with an open-condenser[5] operating at 43.1°C. The cooling water (and condensate, mixture) is pumped out of the condenser (C-Pump) then, instead of being transported entirely to the cooling tower it is partially transported to the raffinery by another pump, specially installed (R-Pump)

The implementation allows 3°C temperature drop before entering the CW pipeline and 0.2°C temperature drop in the pipeline. In the raffine heat exchanger the CW is utilized from 40°C down to 20°C. Thermal energy is therefore withdrawn from the CW in the heat exchange instead of a cooling tower (CT). The system setup is illustrated on a process flow diagram in figure 4.6.

[5] The open condenser is assumed for the analysis as it was considered more sutable for the circumstances at Olca, where cold groundwater is limited and the emphasis is on minimizing the loss of water. Leading the mixture of condensate and cooling water (CW) to the cooling tower assures that evaporation mass flow does not exceed the feed from the turbine, thus no additional makeup water is needed in the cooling tower. As this is assumed the mixing of CW and condensate is invitable the open condenser is selected as a cheaper option

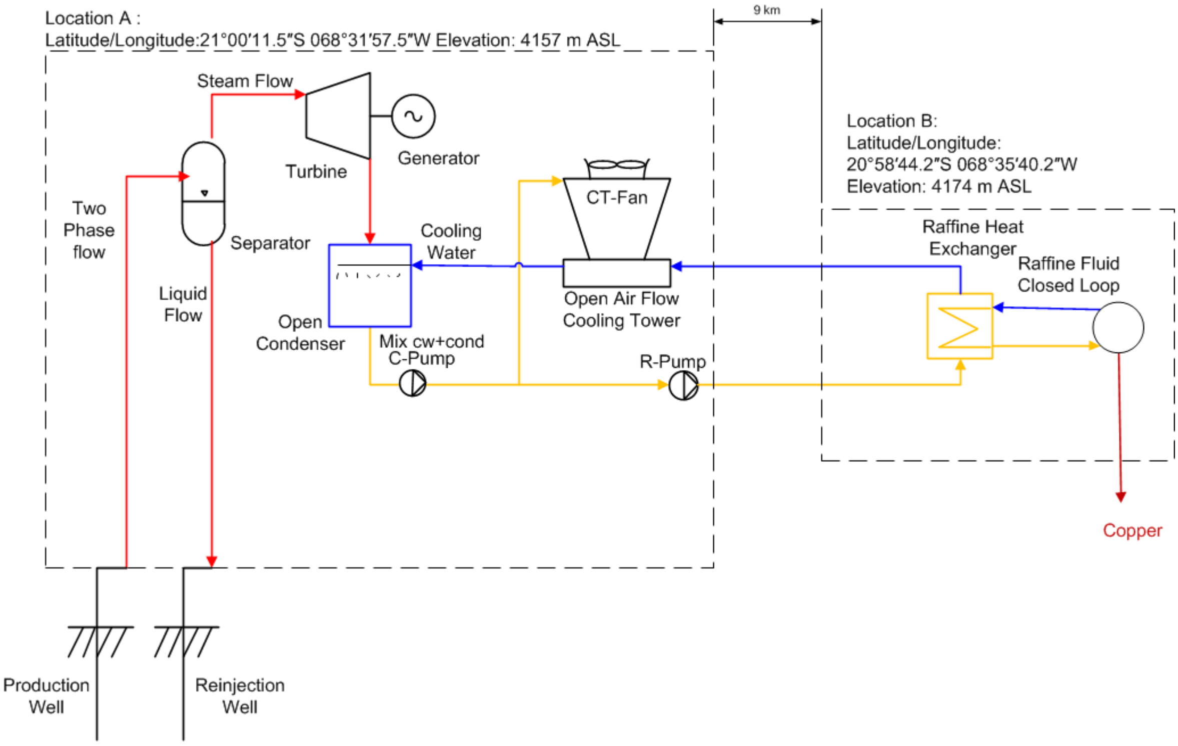


Figure 4.6 Process flow diagram of cooling water approach

The benefits of using this approach are that the cooling water is available, when the operation of the power plant initiates, and thermal energy needs to be withdrawn from the CW in a cooling tower or using heat exchange. This has positive effect on the operation of the power plant; firstly as less cooling is required consequently mitigating needs of cooling tower units to be in operation and therefore saving electricity otherwise used to drive fans in the CT and secondly, evaporation in the towers is less resulting in more fluid to be re-injected to the geothermal reservoir consequently lengthen the production life time of the resource. These two factors are assumed to have an impact on the selection of either of the two approaches for the RS heating.

The downside of this approach is the cost of transporting the water and constructing pipeline from the power plant to the raffinery and back. As results from the preliminary model results in chapter 4.3 shows, 766 kg/s of CW needs to be transported 18.2 km. This affects the installation and operational cost of pipe, pumps and heat exchangers more than if using fluid at higher temperatures.

4.6 Low-temperature Approach

Utilizing low-temperature (LT) water for the heating of the raffine solution (RS) is the second alternative, referred to as the LT-approach and can be carried out regardless of the operation of the power plant. The implementation for the LT-approach involves the fluid to be pumped out of the wells using downhole pumps and gathered in a tank located at the low-temperature drilling area. If pumping is required the LT water is pumped from the tank and on its way to the raffine heat exchangers where thermal energy from the LT water is utilized to heat up the RS. The temperature of the return water is assumed to be 30°C from the heat exchanger and it is re-injected to the geothermal system 4 km on the way back from the raffinery. The process flow diagram for this system setup can be observed in figure 4.7.

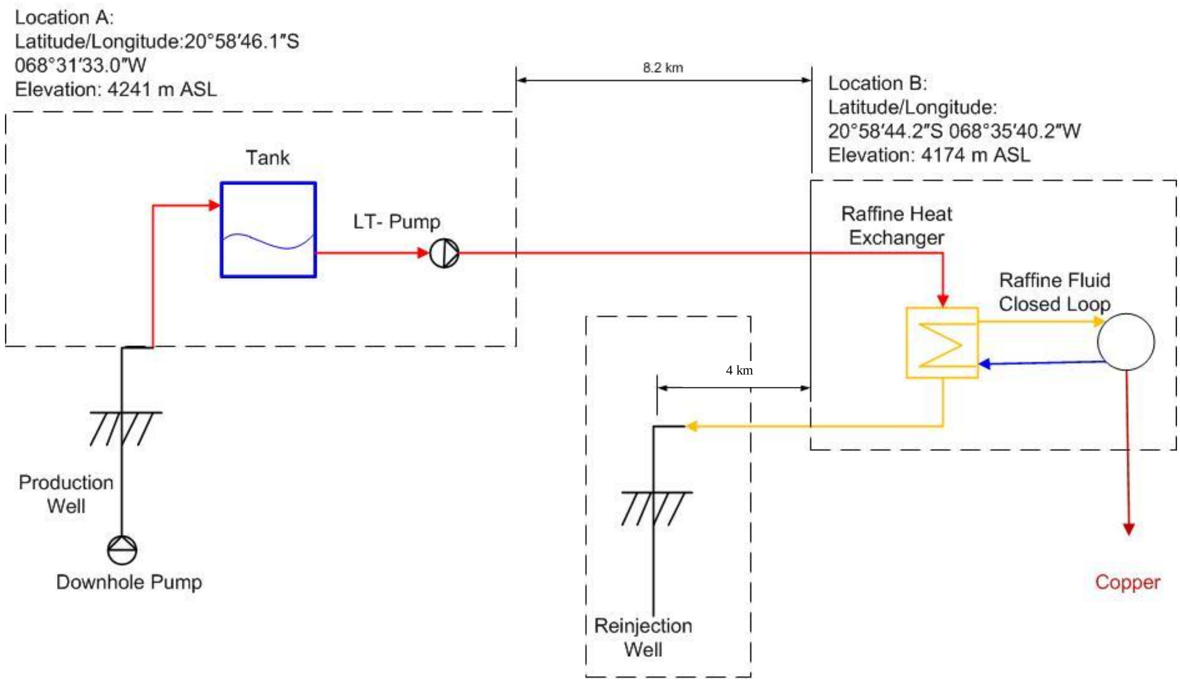


Figure 4.7 Process flow diagram of low-temperature approach

The benefits of this approach are that the temperature of the LT water is higher than the cooling water so the temperature difference is greater (cooling range of the LT fluid), resulting in less water needed for the same energy demanded for heating the RS, or 383 kg/s based on the results from the preliminary model results in chapter 4.3. This consequently leads to less installation and operational cost for the pipe, pumps and heat exchangers than for the CW-approach.

The downside of this approach is that it requires seeking fluid specifically for the RS heating process. It requires new production and re-injection wells to be drilled and a complete system set up only for the purpose of heating the raffine fluid. The uncertainty of this approach is greater than for the CW approach. Firstly, risk involved in the drilling, getting productive wells, enough heat and mass flow and correct chemical composition to be able to carry out a successful utilization. Secondly, if the return fluid re-injected to the reservoir through injection wells will return to the geothermal system - and if so without causing cooling in the system jeopardizing the long-term productivity of the field.

5 Case Study Results

5.1 Cooling Water Approach Results

The main results from the cooling water model can be reviewed in table 5.1 and will be discussed in this chapter. The complete model results with input and output variables, tables and figures are shown in the Appendix.

The total cost of heat exchangers is 1.2 MEUR. The cost of buildings including the raffine heat exchange thermal station (where heating of the raffine solution takes place, called Raffinery), is 1.5 MEUR.

The optimal choice of pipe to transport the fluid from the power plant to the raffinery is 914mm (DN900) Glass Reinforced Plastic (GRP) pipe. The initial cost of this pipeline from the power plant to the raffinery (including pumps) is 4.4 MEUR with annual 0.5 MEUR operational costs. The total pipeline cost is therefore 7.6 MEUR.

The optimal pipe selected for transporting the CW from the raffinery is the GRP pipe. Optimal pipe outside diameter is 711mm (DN700), and the capital cost is 3.2 MEUR, there are no pumps installed and so the operational cost is about 0.1 MEUR annually.

If GRP is to be chosen as a material for the buried pipeline from the power plant to the raffinery the temperature drop will be within an allowable range without any insulation. The total engineering cost is 0.9 MEUR.

The total investment cost of the CW approach is therefore 11.3 MEUR. The cost of installation and operation of the power plant is excluded in this analysis but will be covered in Chapter 5.5. Breakdown of total capital cost for this approach is presented in figure 5.1.

Table 5.1 Cooling water approach, main results

Thermal energy need for heating of the raffine solution	64	WMth
Mass flow of cooling water required to the raffinery	766	kg/s
Optimal pipeline selection: power plant to the raffinery		
- Pipe material	GRP	
- Optimal diameter	914	mm
- Insulation thickness	0	
- Temperature drop in pipeline	<0.2	°C
Optimal pipeline selection: raffinery to the power plant		
- Pipe material	GRP	
- Optimal outside diameter	711	mm
- Insulation thickness	0	
- Temperature drop in pipeline	<0.2	°C
Total investment cost	11.2	MEUR
Total operational cost	825,000	EUR

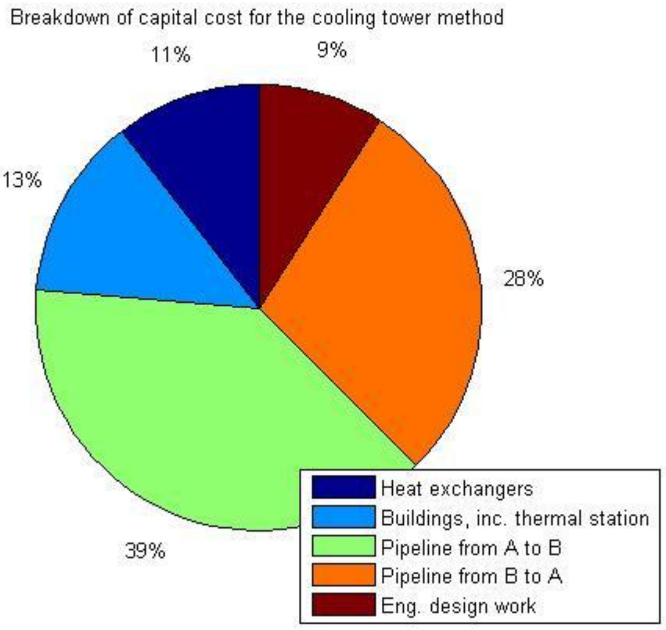


Figure 5.1 Cooling water approach's capital cost breakdown

5.2 Low-Temperature Approach Results

The complete model results can be reviewed in detail in Appendix. The main results are illustrated in table 5.2 and covered in this chapter.

The total cost of the eight heat exchanger units is 400,000 EUR. Construction of buildings including thermal station is 1.5 MEUR.

Eight production wells are predicted to be sufficient for providing 383 kg/s of LT water at 70°C. This corresponds to an average flow rate of 48 kg/s from each well. It is also assumed that success rate during drilling is 80% resulting in a total number of 10 wells to be drilled inside the suggested low-temperature drilling area. Drilling cost for production wells is therefore 4.0 MEUR and 2.0 MEUR for injection wells. Installation of down-hole pumps cost is 600,000 EUR. The cost of gathering pipes and equipment on the low-temperature drilling area, including tank, is 2 MEUR.

First pipe installed is the one transporting the LT fluid (at 70°C) from the LT-geothermal area to the raffine heat exchange station (Raffinery). The installation cost of this pipeline is 2.1 MEUR with annual 75,000 EUR operational costs. Second installed pipe is the one transporting the LT fluid (at 30°C) from the raffinery to re-injection site located 4 km away. The cost of this pipeline is 3 MEUR, with operational cost about 20,000 EUR annually. The optimal choice of both these pipelines is 508mm (DN500) Glass Reinforced Plastic (GRP). There is no need for installing pumps, as the declination in the landscape allows self-discharge of the fluid in the pipelines. The total pipeline cost is therefore 2.2 MEUR.

The total capital cost of the LT approach is assumed 15.5 MEUR, included in that figure is the cost of exploration on the geothermal area which is assumed 0.5 MEUR. The reason for the low exploration cost is that most of the exploring has already taken place in the area. The low-temperature area boundaries have been indicated with MT and TEM and four investigation wells have been drilled verifying the existence of it. If however first few wells turn out to be unsuccessful and re-exploration has to be undertaken somewhere else this cost can increase rapidly. For buried GRP pipes the temperature drop is negligible so no extra insulation is needed. The total engineering cost is assumed 1.4 MEUR. Breakdown of total capital cost is presented in figure 5.2.

Table 5.2 Low-temperature approach main results

Thermal energy need for heating of the raffine solution	64	WMth
Mass flow of cooling water required to the raffinery	383	kg/s
Optimal pipeline selection: Low-temperature drilling area to the raffinery		
- Pipe material	GRP	
- Optimal outside diameter	508	mm
- Insulation thickness	0	
- Temperature drop in pipeline	<0.2	°C
Optimal pipeline selection: raffinery to re-injection area		
- Pipe material	GRP	
- Optimal diameter	508	mm
- Insulation thickness	0	
- Temperature drop in pipeline	<0.2	°C
Total capital cost	15.5	MEUR
Total operational cost	285,000	EUR

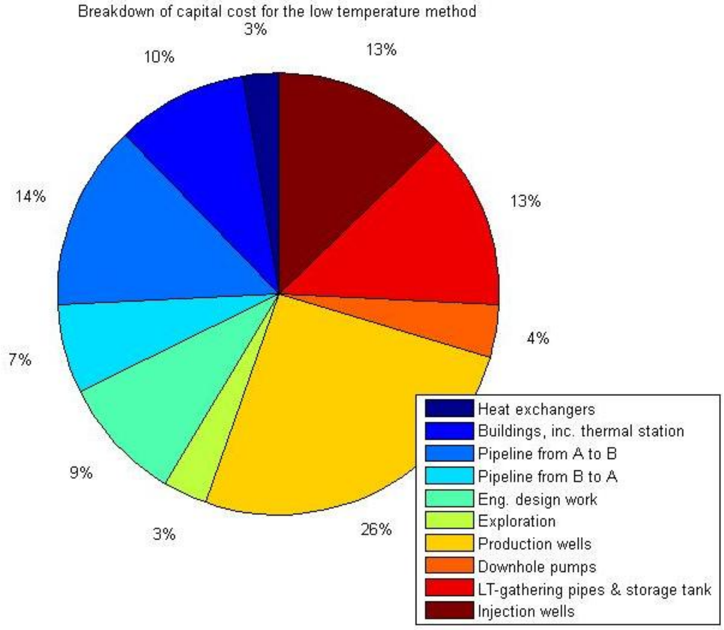


Figure 5.2 Low-temperature capital cost breakdown

5.3 Feasibility study

The economic benefits of the raffine solution (RS) heating project depend on the increase of copper production extraction ratio, price of copper, selling rate and increased production cost. The copper extraction ratio is 40% but 64% with RS heating. Based on last-year's numbers, the amount of Cu cathodes produced in 2010 from heap leaching (HL) was 38,836 ton – the increase in production with the RS heating is therefore roughly 23 thousand ton. The selling rate is assumed 80% and the price of copper is 6 EUR/kg. Assumption is made that increased production cost is 1 EUR per kg of copper produced (23.3 MEUR annually), mostly for transportation and electro winning process. This results in increased net income of 88.5 MEUR per year for copper trading, with the increased production cost extracted from the income. This is presented in table 5.3.

Table 5.3 Copper production and increase in net income for copper trading

Amount of Cu cathodes produced in 2010 by heap leaching (HL)	38,836	ton/yr
Current extraction ratio	40	%
Price of Cu[6]	6	EUR/kg
Increase in copper (Cu) extraction per °C	1.2	%/°C
Optimum Extraction ratio	64	%
Increase in Cu cathode production with optimum RS temperature	23,302	ton/yr
Increased annual net income for copper trading	88.5	MEUR

With the increased net income from copper trading calculated along with the cost analysis of the two approaches suggested in this study, it is possible to evaluate the net present value (NPV) of both approaches (table 5.4).

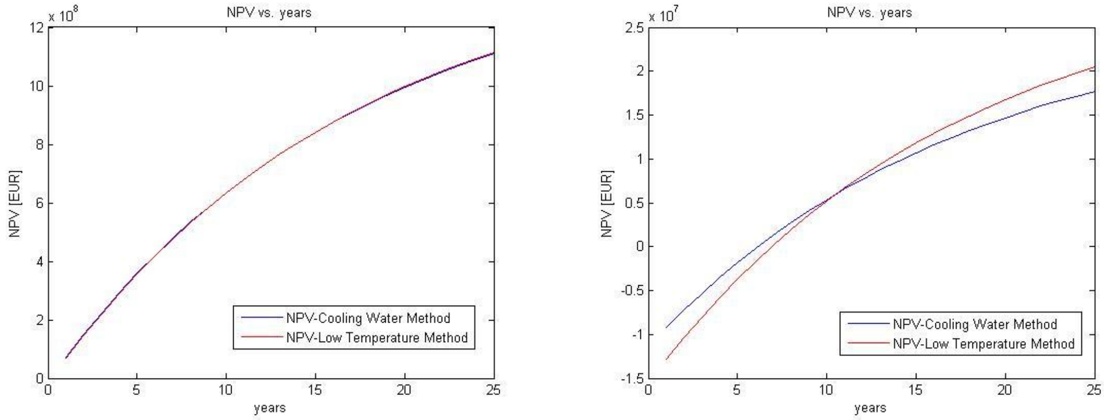
Table 5.4 Net present value (NPV) of both approaches

Raffine heating approach	*NPV*	
Cooling Water (CW)	*1,110*	*MEUR*
Low-Temperature (LT)	*1,113*	*MEUR*

Comparison of the two NPV values presented in table 5.4 show that the approaches are almost equally profitable, with low-temperature (LT) approach slightly more profitable. The capital cost of the LT-approach is 38% higher than for the CW approach, yet the operational cost is higher for the CW approach, while the income is the same (88.5 MEUR per year), explaining this result. The NPV of the two approaches show that over a billion dollars is returned in the next 25 years, when assuming 6% interest rate. Figure 5.3a illustrates the NPV over these 25 years for both approaches. Consequently, based on these results, the payback time is less than a year for both approaches.

[6] The price of copper in the beginning of 2012 (Mundi, 2012)

To emphasize the economic gain of the project the rate of increase in copper production was reduced significantly and the NPV calculated based on the new rates. The increase in copper production was set to 816 ton per year which is only about 3.5% of the 23,302 ton assumed to be produced annually, as an addition to current production, with the raffine heating. The selling rate of the two is the same, or 80% (figure 5.3b).



a. Cu production increase: 23,302 ton/yr

b. Cu production increase: 816 ton/yr

Figure 5.3 Net present value (NPV) results

The 816 ton increased production gives the NPV of 18 and 20 MEUR for the CW and LT approach, respectively – with 7 year payback time. This simple example shows that even with only 816 ton increase in production (3.5% of what is expected) the project starts making growth after 7 years, regardless of which of the two approaches will be chosen.

5.4 Sensitivity analysis

The sensitivity analysis is carried out to display changes in NPV for both approaches when changing the following assumptions.

- Electricity price
- Pipeline cost
- Cost of drilling
- Copper extraction ratio
- Copper price
- Average flow from low-temperature wells

The following table (table 5.5) shows the changes made on each value. These parameters were selected as changing them could lead to different results in terms of which project is considered more economically feasible. Because of the high uncertainty of the values in the analysis +/- 50% range was selected.

Table 5.5 Range of value changes in the sensitivity analysis

	Unit	-50%	Opt. Value	+50%
Electricity price	EUR/kWh	0.03	0.06	0.09
GRP pipeline cost	EUR/mm	0.25	0.5	0.75
Carbon steel pipeline cost	EUR/mm	0.4	0.8	1.2
Cost of drilling	EUR/well	200000	400000	600000
Copper extraction ratio	%/°C	0.6	1.2	1.8
Copper price	EUR/kg	3	6	9
Average flow from low-temperature wells	kg/s	25	50	75

Changing the values for these parameters affects the model as it re-calculates the whole process based on the new assumptions. This can result in different pipe material and/or diameter selection, number of wells needed to be drilled and most importantly the new NPV calculated for the approaches can favor one approach over the other.

The sensitivity analysis was divided into two main categories: the factors affecting cost (price of electricity, flow from wells, drilling- pipeline cost) and the factors affecting income (the copper price and extraction ratio). The results from the first group are presented in figure 5.4 while the second are illustrated in figure 5.5. The results for individual parameter changes can be seen in Appendix.

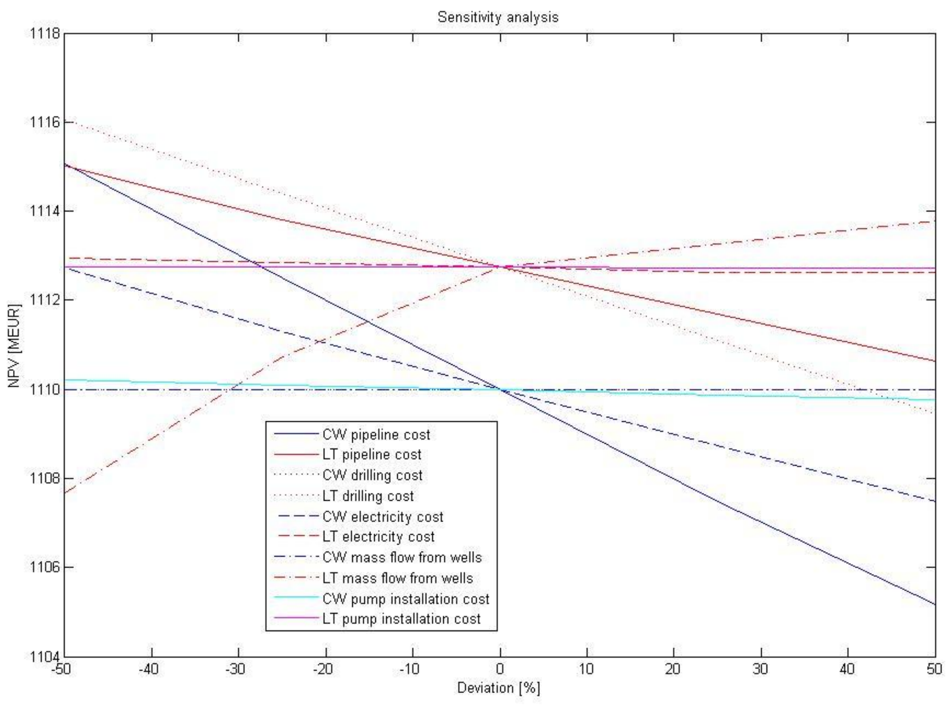


Figure 5.4 Sensitivity analysis for the cost factors

The sensitivity analysis shows how the most critical cost factors affect the NPV. The cost of drilling, mass flow from the wells and pipeline cost are the most critical factors affecting the low-temperature (LT) approach. For the cooling water approach two cost parameters affect the NPV: cost of electricity and pipeline cost. Note that the LT-drilling cost and the mass flow from the LT-wells do not affect the outcome of the cooling water approach. The pipeline and electricity cost affect both projects directly.

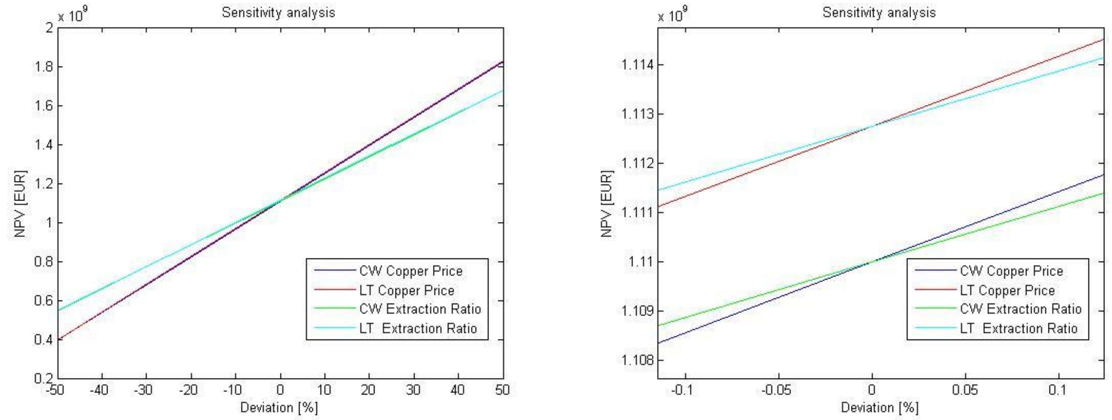
The graph further illustrates that electricity has relatively small effect on the LT- approach as pumping is not required from the LT-geothermal area to the raffinery but highly affects the cooling water approach as the pumping is a big portion of the operational cost. The pump installation cost only affects the projects slightly as the pumping cost is just a small part of the overall installation cost of the pipeline, it has though relatively much more affect on the CW approach as it involves more pumping installation.

The sensitivity graph in figure 5.4 illustrates when one project is more feasible than the other. In cases where one of the red lines is found above one of the blue lines it indicates that the NPV of the LT-approach is higher for that particular assumption change. This works both ways of course.

The sensitivity analysis shows that if the mass flow from the LT-wells becomes much less on average than expected the CW approach becomes more profitable. If the combined mass flow from the LT-wells do not accommodate the flow rate requirements for the RS heating new wells must be drilled. Changing the average flow rate from the wells consequently causes the curve in the NPV as new wells must be drilled.

Pipeline cost is a much greater portion of total cost of the CW approach than of the LT one; namely 67%. Therefore, if the pipeline cost decreases by 50% the CW approach becomes more profitable than the LT approach. The CW approach is highly dependent on the fluctuations in electricity cost. This can, and probably will, affect the whole profit of the project during its operation as electricity cost is often changing. The sensitivity analysis shows that the approaches can be close to equally profitable if the cost of electricity goes down to 0.03 EUR/kWh, but that is a relatively low electricity price and not considered very likely to happen.

The second part of the sensitivity analysis was carried out to analyze how changing the copper price and increase in extraction ratio per degree Celsius would affect the NPV of the project. It turns out that the profit of the project is highly dependent on these parameters, for both approaches. Changing the copper price from 6 to 9 increases the overall NPV of the project by 64%, similarly it reduces the NPV of the project by the same percentage when reducing the copper price down one half. If the extraction ratio of the copper in the heap leaching process reaches only half of what is assumed for the project (or 0.4% per degree Celsius) it will, regardless of which approach is carried out, be 51% less profitable, but on the other hand 51% more profitable if the extraction ratio reaches 1.8%/°C. The results from the second part of the sensitivity analysis is presented in figure 5.5, where figure 5.5a illustrates the complete change in NPV versus the deviation of parameters from -50% to 50%, while the other, figure 5.5b, is a closer look on the changes near zero percentage deviation. By having this closer look it is possible to review the difference between the approaches, which is relatively small.



a. NPV vs. Deviation   b. NPV vs. Deviation, closer view

Figure 5.5 Sensitivity analysis for the income factors.

5.5 Additional Considerations and Benefits of the Cooling Water Approach

The power plant was developed by assuming 10 production wells providing 50 kg/s of high-temperature fluid each. Saturated liquid conditions are assumed at the bottom of the well and 250°C, so the enthalpy content of the fluid is 1086 kJ/kg. This results in roughly 51.5 MW_e produced by the turbine when operating the separator at ~3.9 bar_a and condenser at 0.09 bar (43.1°C). This is presented in figure 5.6 and covered in more detail in the Appendix.

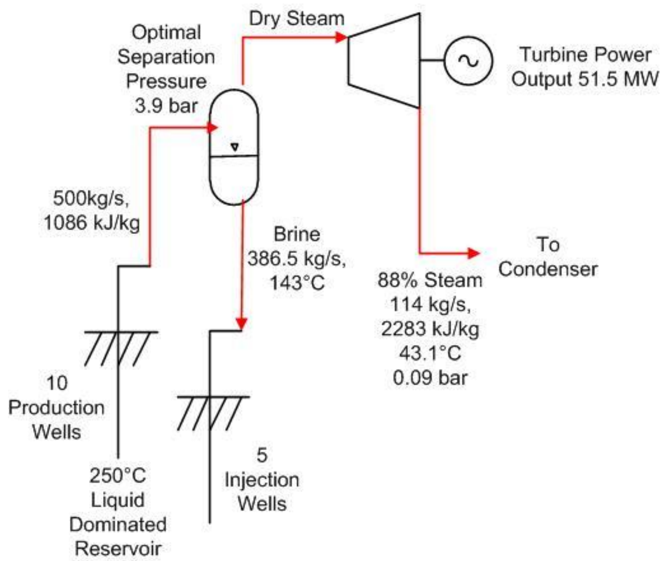


Figure 5.6 Power plant results

To condense the steam from the turbine at this temperature, ~2854 kg/s of ~20°C CW are utilized in an open condenser. The power plant is designed to work independently, with cooling towers big enough to serve the cooling needs of the plant (figure 5.7). If the CW approach initiates, the cooling process of the CW changes, as part of the water (766 kg/s) is lead to the raffinery, but the rest (2202 kg/s) to the towers (figure 5.8).

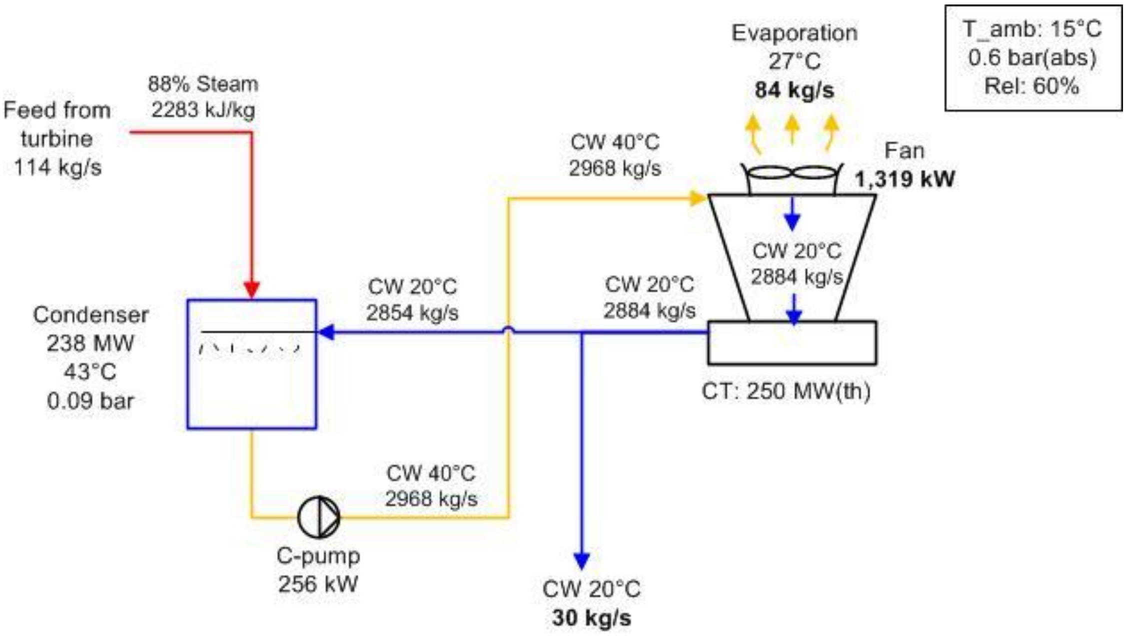


Figure 5.7 Operation of cooling towers and condensers, without CW-approach

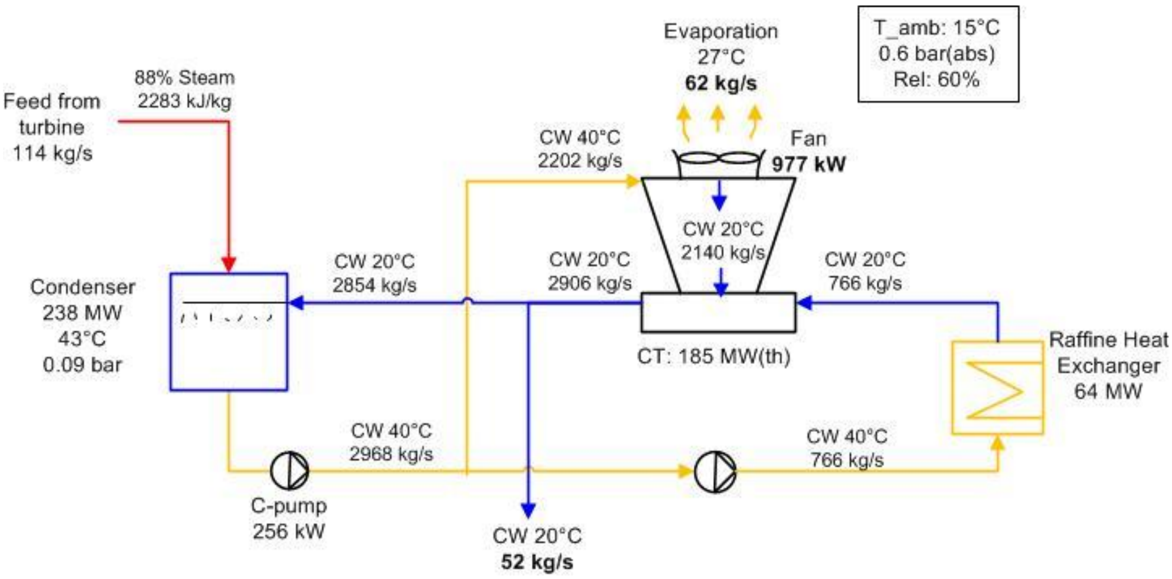


Figure 5.8 Operation of cooling towers and condensers, with CW-approach

The first benefit factor of the CW-approach is the energy savings for the power plant: When part of the CW is lead to the Raffinery less air is needed for cooling. Then, the calculated energy needs for the motors driving the CT-fans is 979kW reduced from 1,319 kW. This enhances the net power output of the power plant[7,8]. To illustrate the affect this

[7] Note that the pumping CW to the Raffinery is included in the feasibility study for the CW approach

has on the operation of the power plant, the NPV of the power plant is calculated, both with and without the RS heating (table 5.6). This is done by using the same assumptions as before regarding electricity price (0.06 EUR/kWh), period (25 years) and interest rate (6%). Based on assumptions in chapter 4.2, the capital investment cost for this power plant is calculated 189 MEUR, with 18 MEUR net income but different operational cost. This is presented in table 5.6. The results from the NPV analysis show that after the first 25 years of operating the power plant the NPV is 5% higher associated with the RS heating than operating without it. The NPV calculations further illustrate the prospective payback time for power plants build in the area. The net cash flow for the 50MW power plant becomes positive in the 17th year of operation.

Table 5.6 CW approach benefits

Size of power plant	*50MW*	
With/without RS heating	without	with
Energy demands of motors driving CT-fans [kW]	**1,319**	**979**
Power plant capital cost [MEUR]	189	189
Power plant operating cost [MEUR]	8.82	8.64
Annual income (selling electricity) [MEUR]	27	27
NPV of power plant project [MEUR]	**44.6**	**46.9**
Payback time [years]	17	17
Mass flow of cooling water evaporating in cooling towers [kg/s]	84	62
Mass flow of cooling water to inject to the geothermal reservoir [kg/s]	**30**	**52**

The 35% energy savings with RS heating is equivalent to resting one CT-unit out of four installed for the plant, as usually numbers of CT units are installed instead of one big unit. Consequently, if the net power output from the power plant is equal of what is originally planned, then with using RS heating the net power output from the plant is above what is originally planned as less energy is needed for its operation. In addition, while one unit is in rest, maintenance can be carried out without any depletion in the power plant performance.

The second benefit regards certain weather conditions that might occur: During high relative humidity and/or high ambient temperature the wet bulb temperature increases and thus the efficiency of the cooling reduces. It is however, possible to operate RS heating using CW and, by taking the same example as in previous paragraph, having all four units operating. This will enhance the cooling of the CW reducing the temperature in the condenser closer to its original value. Each year, the site experiences so-called “Bolivian Winter” where high relative humidity strikes for a two-month period, and during these times, operation of RS heating and all four units help sustaining full power plant performance.

[8] Net power output is equal to the power output from the turbine minus the pumping required for removing the condensate and cooling water mixture out of the condenser (256 kW) minus the energy demands of motors driving CT-fans [kW]

The third benefit regards evaporation in the towers. The 50MW power plant has cooling towers designed for $250MW_{th}$ or to cool 2968 kg/s of ~40°C hot CW down to ~20°C. This results in 84 kg/s evaporating in the CT –therefore, as 2584 kg/s is circulated back to the condenser, 30 kg/s of this water can be re-injected to the geothermal reservoir. However, when operating the RS heating the cooling capacity of the towers, under normal conditions, is $185MW_{th}$ which results in evaporation mass flow of 62.2 kg/s –therefore, with the mass flow of cooling water returning to the condenser unchanged, 52 kg/s can be re-injected to the geothermal reservoir. This is 22 kg additional fluid that is re-injected every second when the power plant is in operation or equivalent to more than 690,000 cubic meters a year. This considerable amount of fluid would most definitely help in terms of sustainable utilization. This is presented in table 5.6 and illustrated on figure 5.7 and 5.8.

All these three factors are considered very important for Collahuasi to evaluate in terms of selecting either of the two approaches for the heating of the RS. These calculations further illustrate that the evaporation mass flow is less than the mass flow of steam entering the condenser; therefore no makeup water is needed for the condensing process.

6 Conclusions and Recommendations

The purpose of this project is to study the feasibility of using geothermal energy for heating a raffine solution used in copper production. A model has been developed to simulate the main processes involved, and it has been applied as a case study of the Collahuasi copper mine in Chile. Two alternatives for a heat source have been compared, namely cooling water from a prospective power plant and low-temperature geothermal fluid from a nearby reservoir.

The model implementation of the cooling water approach was initiated with the modeling of a single-flash power plant, associated with an open condenser that uses cooling water to condense steam flowing from the turbine, by mixing cooling water and condensate. The cooling water and condensate mixture are circulated partly to the cooling towers and, if the cooling water approach is chosen, partly to the raffine heat exchangers. The cooling water then needs to travel 9.1 kilometers to the raffinery and back the same route to the power plant. This is based on a preliminary assumption regarding the prospective power plant's location.

The implementation of the low-temperature fluid approach proposes pumping the fluid from the wells to a gathering tank and transporting it 8.2 kilometers to the raffine heat exchangers. This is based on the tank being located at the center of the low-temperature drilling area. After the thermal exchange between the fluids has taken place, the low-temperature fluid would be re-injected into the geothermal reservoir through injection wells located four kilometers from the raffinery.

The study shows that the capital cost of the cooling water approach accounts for less than two thirds of the low-temperature approach capital cost. This is despite that much large diameter pipe is required for the transportation of the cooling water and the pipeline is much longer. The reason for this is that the other approach requires drilling and a complete system set-up, entirely dedicated to the heating of the raffine solution, which accounts for more than half of the investment capital cost. However, the operational cost of the cooling water approach is almost threefold when compared to the low-temperature approach. The low-temperature fluid travels with self-discharge the entire way while the cooling water approach requires pumping, explaining this difference. The low-temperature approach is therefore slightly more profitable than the cooling water approach. This was verified using a sensitivity analysis, and for all occasions, except in the case of either very high drilling costs or decreasing average well productivity, the low-temperature approach is more profitable. The positioning of the power plant, low-temperature area and injection area with respect to the raffine heating facilities plays a big role in terms of cost.

The feasibility study showed an indisputable increase in net income of cash flow from copper trading by the geothermically-enhanced copper production, for both approaches. In fact, the payback period was evaluated at less than a calendar year, after which point, this investment would begin generating profit. To verify the conclusion that the installation of the Collahuasi raffine heating system is indeed profitable, the increase in copper production levels was reduced to 3.5% from the original assumption. The project remained profitable even under the more conservative production criteria.

The question is therefore not whether to initiate the raffine heating project, but in what manner. Despite that the low-temperature approach is considered more profitable, the inherent risk of not locating an adequate productive reservoir must be considered. The study showed the prospective benefits for the operation of a geothermal power plant associated with the cooling water approach and these factors can therefore affect the decision regarding initiating one approach over the other.

However, because of the prospective high profits associated with the utilization of geothermal energy for the raffine heating, and because the fact that the power plant has not been initiated on the Collahuasi site yet (and it will take some years to commence), it is recommended that drilling in the low-temperature area should commence as soon as possible, in order to initiate the low-temperature approach for the raffine heating. If however, the low-temperature reservoir cannot withstand the proposed utilization, it would be possible to re-evaluate the feasibility of the cooling water approach, after the commencement of the prospective power plant.

The Collahuasi mine has the opportunity to be the first copper mine with a geothermically-enhanced heap leaching system. For this to happen, the optimal temperature of the raffine fluid and the increased copper extraction ratio must be re-evaluated and confirmed through testing on a large scale.

The simulation model developed specially for this study can be updated when more information about the geothermal systems are available, such as location of the power plant, prices and other parameters that have at this point been based on assumptions. It could therefore not only be used as a decision-making tool at later stages of the Collahuasi raffine heating project but it could also be expanded to predict the feasibility of similar projects. It is further recommended to identify copper mines close to prospective geothermal energy sites as this study showed an indisputable net profit by using geothermal energy for the heating of raffine solution for copper production.

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Appendix

Specific Humidity Calculations

The specific humidity is a function of dry bulb temperature T_a and wet bulb temperature T_{wb} in Kelvin [K] and the ambient pressure p_a and p_v in Pascal [Pa] which is the vapor pressure of the humid air.

$$\omega\{T_a, T_{wb}, p_a, p_v\}$$

where p_v is a function of the wet bulb temperature

$$p_v\{T_{wb}\}$$

First, the specific humidity of saturated air at T_a is determined, thus $T_a = T_{wb}$, so p_v is determined using (Kröger D. G., 2004):

$$p_v = 10^{z1+z2+z3}$$

where z_1, z_2 and z_3 are parameters calculated as follows (Kröger D. G., 2004):

$$z_1 = 10.79586 * \left(1 - \frac{273.16}{T_{wb}}\right) + 5.02808 * \log_{10}\left(\frac{273.16}{T_{wb}}\right)$$

$$z_2 = 1.50474e^{-4} * \left(1 - 10^{-8.29692*\left(\left(\frac{T_{wb}}{273.16}\right)-1\right)}\right)$$

$$z_3 = 4.2873e^{-4} * \left(10^{4.76955*\left(1-\frac{273.16}{T_{wb}}\right)} - 1\right) + 2.786118312$$

which is then used to determine the specific humidity (ω_{sat}) of saturated air, given in kg of water per kg of dry air [kg/kg]. Following equations are used to determine specific humidity (ω) (Kröger D. G., 2004):

$$\omega = y_1 * y_2 - y_3$$

where y_1, y_2 and y_3 are parameters calculated as follows (Kröger D. G., 2004):

$$y_1 = \frac{(2501.6 - 2.3262 * (T_{wb} - 273.15))}{(2501.6 + 1.8577 * (T_a - 273.15) - 4.184 * (T_{wb} - 273.15))}$$

$$y_2 = \frac{0.62509 * p_v}{p_a - 1.005 * p_v}$$

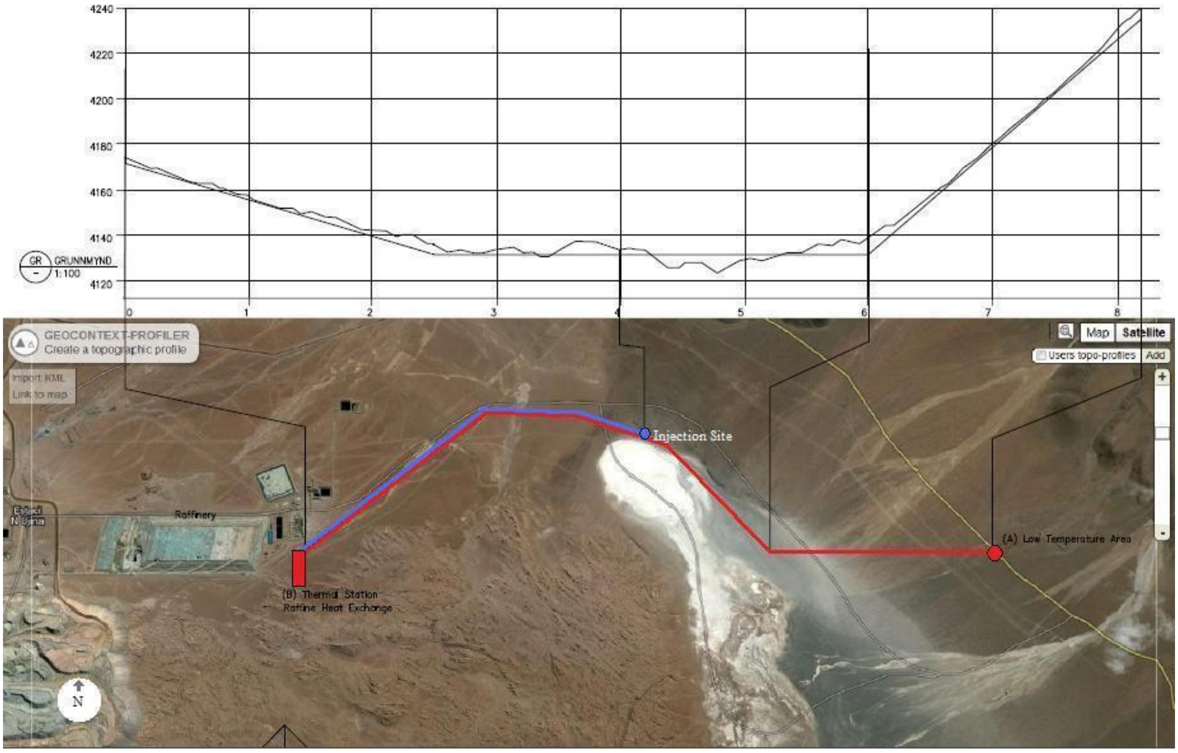
$$y_3 = \frac{\left(1.00416 * (T_a - T_{wb})\right)}{\left(2501.6 + 1.8577 * (T_a - 273.15) - 4.184 * (T_{wb} - 273.15)\right)}$$

When ω_{sat} has been determined for $T_a = T_{wb}$ it is possible to compute the specific humidity of unsaturated dry air entering the cooling tower (ω_{in}). This is possible because the unsaturated specific humidity of air is a function of the saturated specific humidity ω_{sat} and its relative humidity, as (Kröger D. G., 2004):

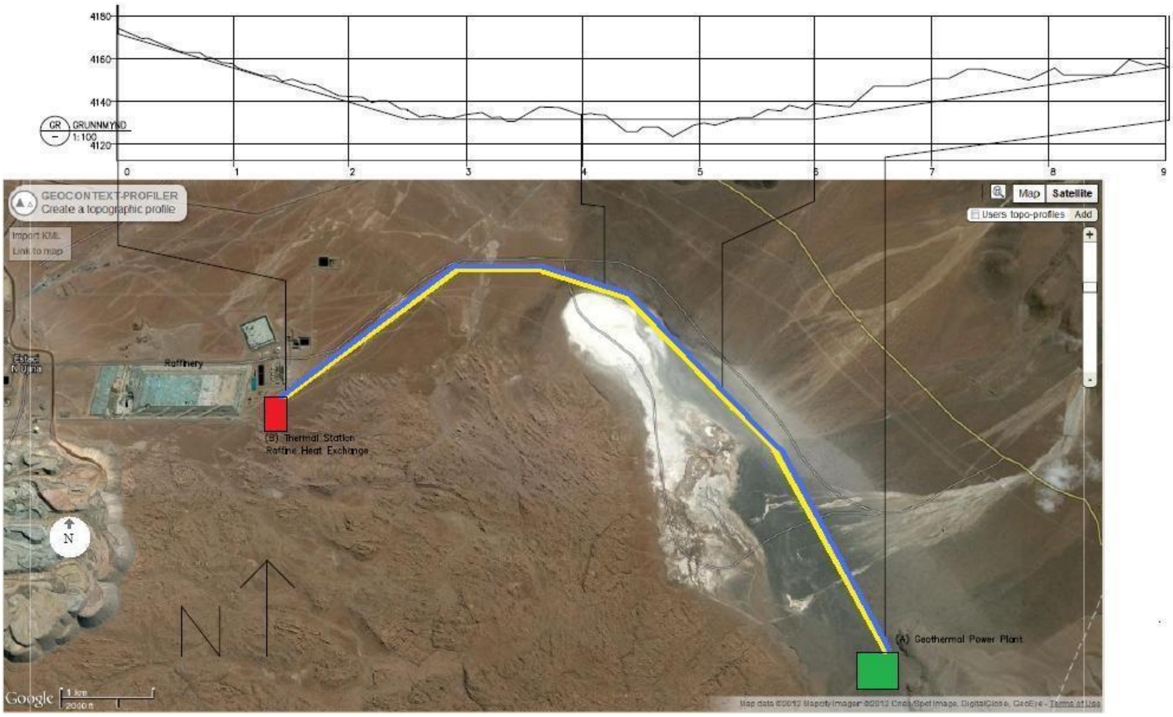
$$\omega_{in} = rel * \omega_{sat}$$

where ω_{in} is the specific humidity of the unsaturated air entering the CT and *rel* [%] relative humidity of the air.

Pipeline Routes



LT Pipeline route



CW Pipeline route

Cooling Water Model & Power Plant Model Inputs & Outputs

Table below: Inputs and Outputs - Raffine heating model for cooling water approach

Vaiable type			
Input	Temperature of the Cooling Water entering the heat exchanger	40	°C
	Temperature of the raffine fluid entering heat exchangers	10	°C
	Temperature of the raffine fluid exiting heat exchangers	30	°C
Output	Mass flow of cooling water to the raffinery	766	kg/s
	Temperature of the Cooling Water exiting the heat exchanger	20	°C
	Thermal energy extracted in the raffine process	64	WM_{th}

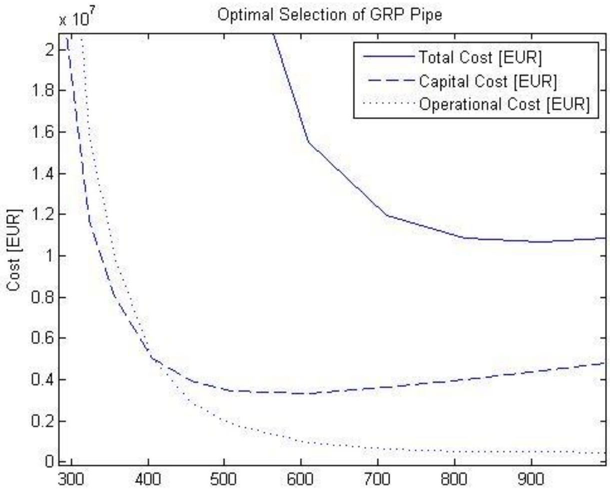


Figure above: Optimal diameter of GRP pipe transporting cooling water from power plant to the raffinery.

Table below: Cost comparison of different materials for pipelines transporting cooling water from power plant to the raffinery

	Optimal Diameter [mm]	Total cost [MEUR]	Capital cost [MEUR]	Operational cost [MEUR]
Stainless Steel pipe	711	27.6	15.9	0.9
Carbon steel pipe*	914	14.1	6.9	0.6
Glass Reinforced Plastic	914	10.7	4.4	0.5

*Not applicable for cooling water

Table below: Inputs and Outputs – Pipeline optimization model for transportation of cooling water from power plant to raffinery

Vaiable type			
Input	Average temperature of the water in the pipe	~40	°C
	Average mass flow	766	kg/s
	Design pressure	15	bar
	Design temperature	50	°C
	Elevation (increase)	+50	m
	Pipe length	9100	m
	Interest rate	6	[%]
	Depreciation time	25	[yr]
	Electricity cost	0.06	[EUR/kWh]
Output	Pipe material	GRP	
	Optimal diameter	914	mm

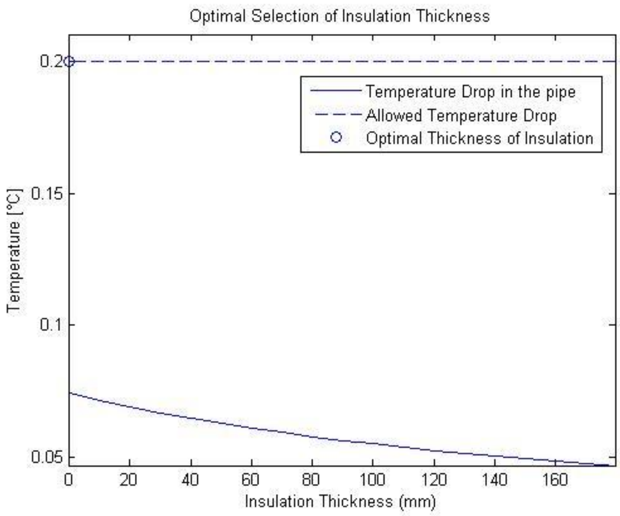


Figure above: Temperature drop in a buried 9100m GRP pipeline from power plant to raffinery with different insulation thicknesses

Table below: Inputs and outputs - Insulation thickness model for a buried 9100m GRP pipeline from power plant to raffinery

Vaiable type			
Input	Temperature of the water exiting the pipe	40	°C
	Inside diameter of pipe	898	mm
	Outside diameter of pipe	914	mm
	Allowed temperature drop in pipeline	0.2	°C
Output	Minimum insulation thickness	0	mm
	Cooling number of the combined thermal resistance (k)	0.4024	-
	Temperature drop in the pipeline	0.0744	°C
	Temperature of the water entering the pipe	40.0744	°C

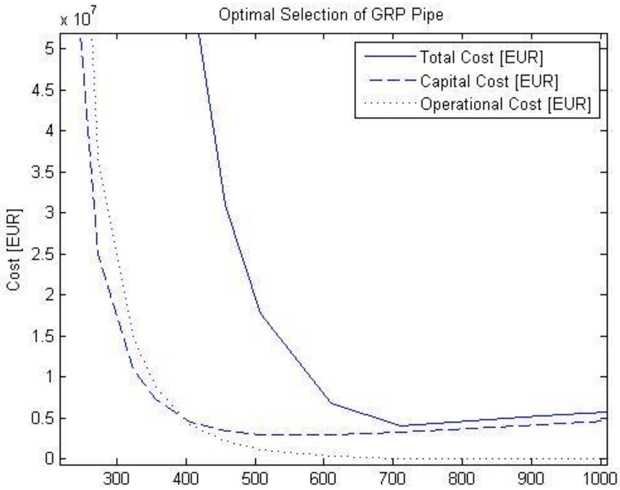


Figure above: Optimal diameter of GRP pipe transporting cooling water from raffinery to the power plant

Table below: Cost comparison of different materials for pipelines transporting cooling water from raffinery to the power plant

	Optimal Diameter [mm]	Total cost [MEUR]	Capital cost [MEUR]	Operational cost [MEUR]
Stainless Steel pipe	711	19.5	15.5	0.3
Carbon steel pipe*	711	7.2	5.2	0.2
Glass Reinforced Plastic	711	4.1	3.2	0.1

*Not applicable for cooling water

Table below: Inputs and Outputs – Pipeline optimization model for transportation of cooling water from raffinery to power plant

Vaiable type			
Input	Average temperature of the water in the pipe	~20	°C
	Average mass flow	766	kg/s
	Design pressure	10	bar
	Design temperature	30	°C
	Elevation (decrease)	-40	m
	Pipe length	9100	m
	Interest rate	6	[%]
	Depreciation time	25	[yr]
	Electricity cost	0.06	[EUR/kWh]
Output	Pipe material	GRP	
	Optimal diameter	711	mm

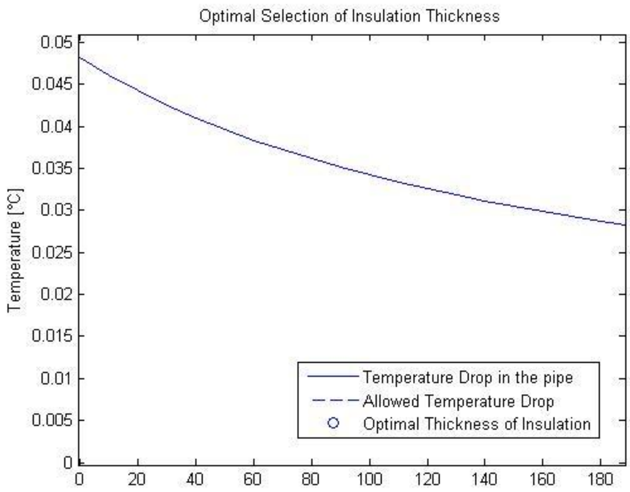


Figure above: Temperature drop in a buried 9100m GRP pipeline from raffinery to power plant with different insulation thicknesses

Table below: Inputs and outputs - Insulation thickness model for a buried 9100m GRP pipeline from raffinery returning to power plant

Vaiable type			
Input	Temperature of the water entering the pipe	20	°C
	Inside diameter of pipe	701	mm
	Outside diameter of pipe	711	mm
	Allowed temperature drop in pipeline	10	°C
Output	Minimum insulation thickness	0	mm
	Cooling number of the combined thermal resistance (k)	0.3777	-
	Temperature drop in the pipeline	0.0482	°C
	Temperature of the water exiting the pipe	19.9518	°C

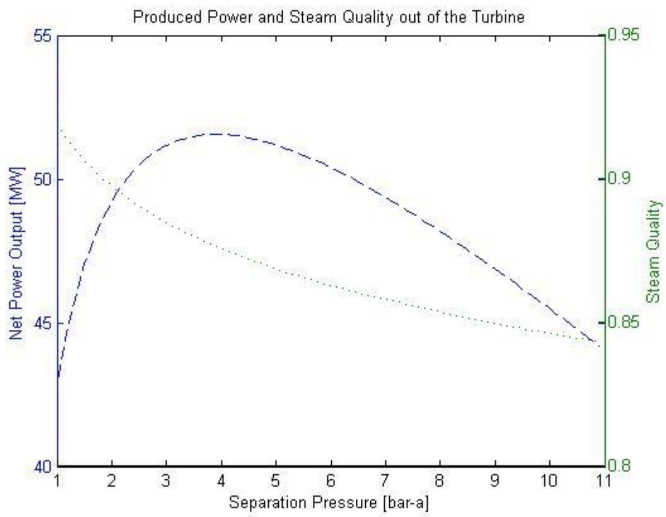


Figure above: Optimal power output from the turbine in the conceptual power plant

Table below: Inputs and outputs – Power plant model

Vaiable type			
Input	Number of production wells	10	
	Average production (mass flow) per well	50	kg/s
	Characteristics of reservoir	liquid dominated	
	Enthalpy of the production fluid	1086	kJ/kg
	Temperature in condeser	~43,1	°C
	Pressure in condenser	~0.09	bar
Output	Optimal seperation pressure	3.9	bar
	Power output from the turbine	51.5	MWe
	Net power output from the power plant, excl. cooling tower	51.3	MWe
	Quality of steam exiting the turbine	88%	
	Condenser thermal energy demand	238.7	MWth
	Enthalpy of steam exiting the turbine	2283	kJ/kg
	Mass flow of steam exiting the turbine	113.5	kg/s
	Steam quality of fluid after seperator	23%	
	Temperature of brine exiting seperator	142.7	°C
	Energy demand of condenser pump	256	kW
	Mass flow of brine exiting the separator	386.5	kg/s

Table below: Inputs and outputs – Cooling tower model, with raffine heating associated

Vaiable type			
Input	Temperature of fluid entering cooling tower	40.07	°C
	Temperature of fluid exiting cooling tower	19.95	°C
	Mass flow of cooling water entering cooling tower	2201.9	kg/s
	Design ambient temperature	15	°C
	Design ambient pressure	0.6	bar
	Design relative humidity	60%	
Output	Wet bulb temperature	10	°C
	Evaporation mass flow	62.2	kg/s
	Blowdown mass flow	51.3	kg/s
	Evaporation ratio	2.8%	
	Mass flow of cooling water returning from cooling tower	2139.6	kg/s
	Makeup water needs	0	kg/s
	Approach Temperature	18	°F
	Temperature of the air exiting the cooling tower	26.9	°C
	Energy needs for motor driving fan	978.6	kW
	Tower size factor	0.89	
	Energy extract in the cooling tower	185	MWth

Table below: Inputs and outputs – Cooling tower model, without raffine heating associated

Vaiable type			
Input	Temperature of fluid entering cooling tower	40.07	°C
	Temperature of fluid exiting cooling tower	19.95	°C
	Mass flow of cooling water entering cooling tower	**2967.9**	kg/s
	Design ambient temperature	15	°C
	Design ambient pressure	0.6	bar
	Design relative humidity	60%	
Output	Wet bulb temperature	10	°C
	Evaporation mass flow	**83.9**	kg/s
	Blowdown mass flow	**29.7**	kg/s
	Evaporation ratio	**2.9%**	
	Mass flow of cooling water returning from cooling tower	**2884**	kg/s
	Makeup water needs	0	kg/s
	Approach Temperature	18	°F
	Temperature of the air exiting the cooling tower	26.9	°C
	Energy needs for motor driving fan	**1319.3**	kW
	Tower size factor	0.89	
	Energy extract in the cooling tower	250	MWth

Low-temperature Model Inputs & Outputs

Table below: Inputs and Outputs - Raffine heating model for low-temperature approach

Vaiable type			
Input	Temperature of the low-temperature fluid entering the heat exchanger	70	°C
	Temperature of the raffine fluid entering heat exchangers	10	°C
	Temperature of the raffine fluid exiting heat exchangers	30	°C
Output	Mass flow of low-temperature fluid to the raffinery	383	kg/s
	Temperature of the low-temperature fluid exiting the heat exchanger	30	°C

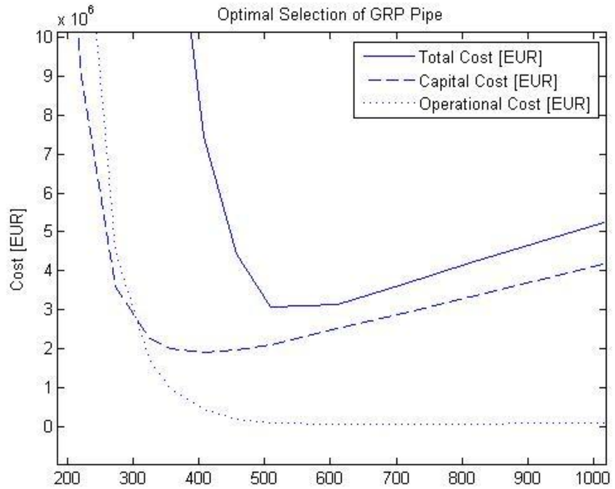


Figure above: Optimal diameter of GRP pipe transporting low-temperature water from production site to the raffinery

Table below: Cost comparison of different materials for pipelines transporting low-temperature water from production site to the raffinery

	Optimal Diameter [mm]	Total cost [MEUR]	Capital cost [MEUR]	Operational cost [EUR]
Stainless Steel pipe	508	13.1	10	240,000
Carbon steel pipe	610	5	4	80,000
Glass Reinforced Plastic	508	3.1	2.1	75,000

Table below: Inputs and Outputs – Pipeline optimization model for transporting low-temperature water from production site to the raffinery

Vaiable type			
Input	Average temperature of the water in the pipe	~70	°C
	Average mass flow	383	kg/s
	Design pressure	15	bar
	Design temperature	80	°C
	Elevation (decrease)	-40	m
	Pipe length	8200	m
	Interest rate	6	[%]
	Depreciation time	25	[yr]
	Electricity cost	0.06	[EUR/kWh]
Output	Pipe material	GRP	
	Optimal diameter	508	mm

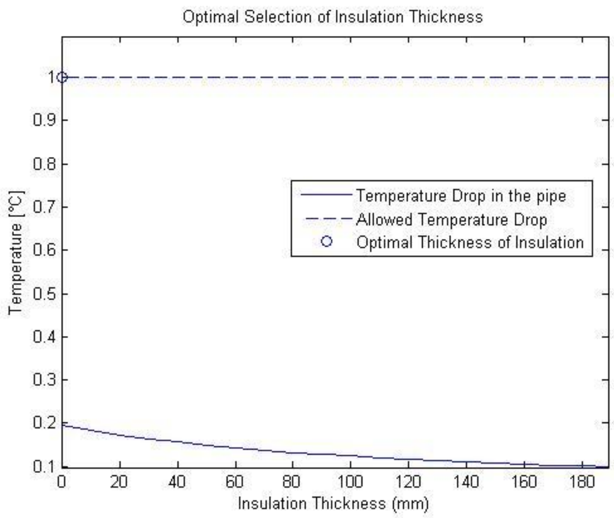


Figure above: Temperature drop in a buried 8200m GRP pipeline from low-temperature production site to the raffinery

Table below: Inputs and outputs - Insulation thickness model for a buried 8200m GRP pipeline from low-temperature production site to the raffinery

Vaiable type			
Input	Temperature of the water exiting the pipe	70	°C
	Inside diameter of pipe	496.8	mm
	Outside diameter of pipe	508	mm
	Allowed temperature drop in pipeline	1	°C
Output	Minimum insulation thickness	0	mm
	Cooling number of the combined thermal resistance (k)	0.4049	-
	Temperature drop in the pipeline	0.1967	°C
	Temperature of the water entering the pipe	70.1967	°C

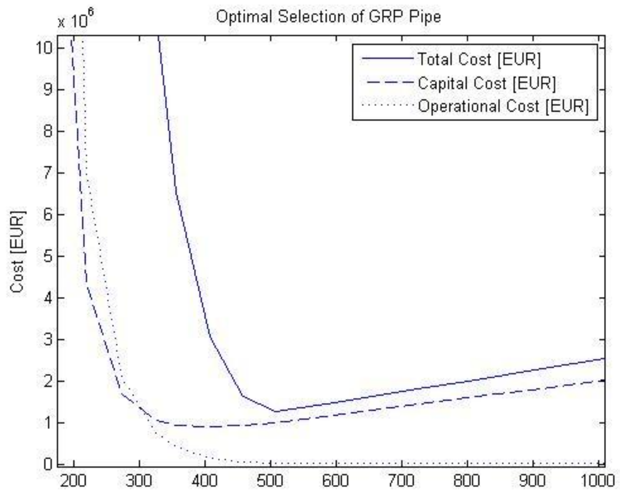


Figure above: Optimal diameter of GRP pipe transporting low-temperature water from raffinery to injection site.

Table below: Cost comparison of different materials for pipelines transporting low-temperature water from raffinery to injection site.

	Optimal Diameter [mm]	Total cost [MEUR]	Capital cost [MEUR]	Operational cost [EUR]
Stainless Steel pipe	457	6.1	4.4	130,000
Carbon steel pipe	508	2.1	1.6	40,000
Glass Reinforced Plastic	508	1.3	1	20,000

Table below: Inputs and Outputs – Pipeline optimization model for transportation of low-temperature water from raffinery to injection site.

Input	Average temperature of the water in the pipe	~30	°C
	Average mass flow	383	kg/s
	Design pressure	15	bar
	Design temperature	40	°C
	Elevation (decrease)	-30	m
	Pipe length	4000	m
	Interest rate	6	[%]
	Depreciation time	25	[yr]
	Electricity cost	0.06	[EUR/kWh]
Output	Pipe material	GRP	
	Optimal diameter	508	mm

Table below: Inputs and Outputs – Silica scaling model

Variable type			
Input	*DSC* (Dissolved silica concentration)	100	mg/L
	Temperature of geothermal fluid	70	°C
	SSI (Silica Saturation Index)	1	bar
Output	Saturation silica concentration at 70°C	128	mg/L
	SSI at 70°C	0.78	
	Scaling temperature with 100 mg/L *DSC*	14	°C

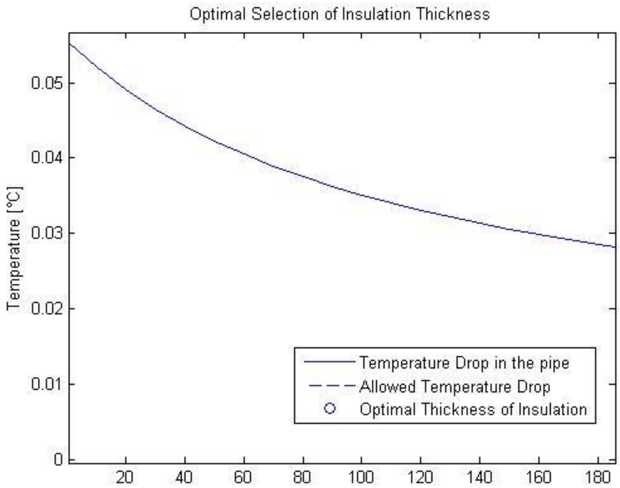
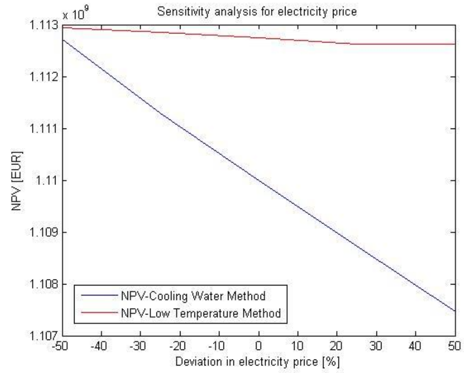


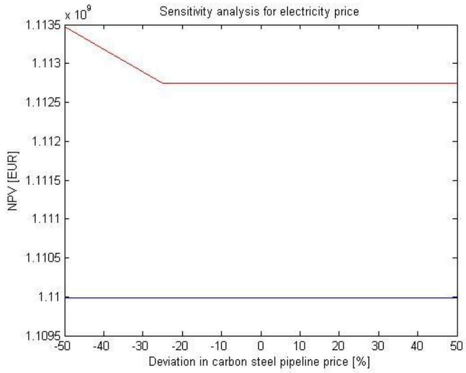
Figure above: Temperature drop in a buried 4,000 m GRP pipeline from raffinery to injection site with different insulation thicknesses

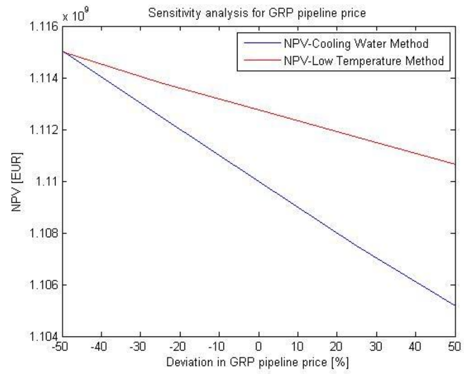
Table below: Inputs and outputs - Insulation thickness model for a buried 4,000 m GRP pipeline from raffinery to injection site

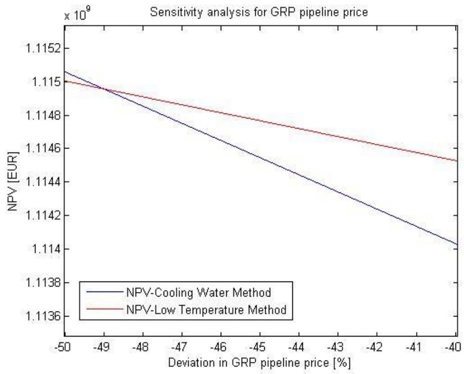
Vaiable type			
Input	Temperature of the water entering the pipe	30	°C
	Inside diameter of pipe	496.8	mm
	Outside diameter of pipe	508	mm
	Allowed temperature drop in pipeline	16	°C
Output	Minimum insulation thickness	0	mm
	Cooling number of the combined thermal resistance (k)	0.4049	-
	Temperature drop in the pipeline	0.0556	°C
	Temperature of the water exiting the pipe	29.9444	°C

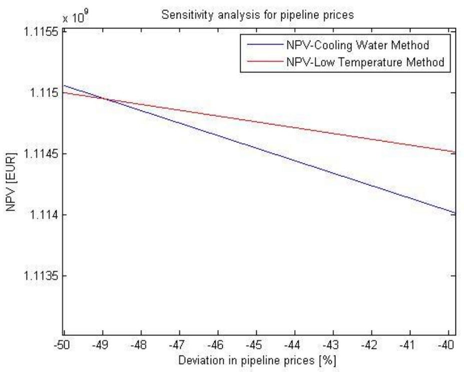
Sensitivity Analysis

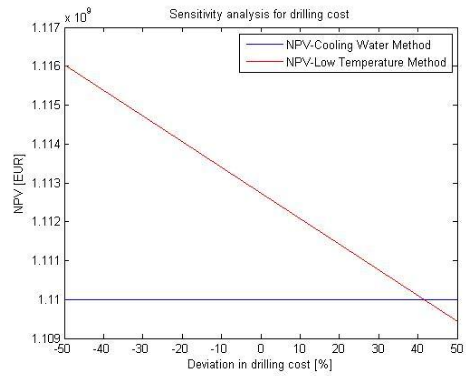


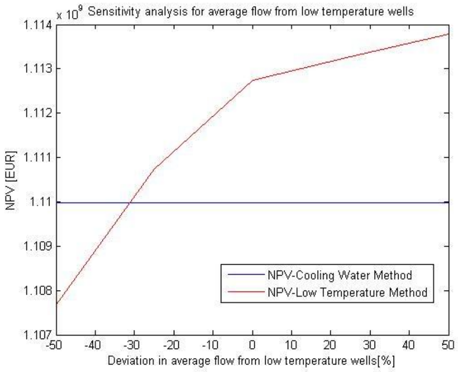




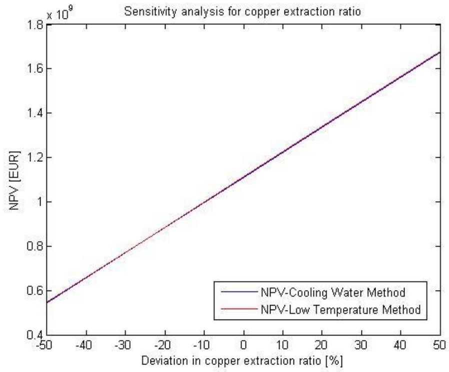




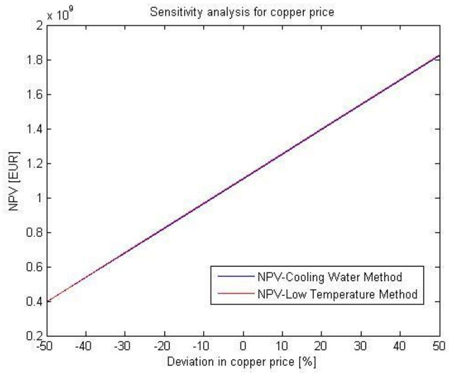




x 10^9 Sensitivity analysis for average flow from low temperature wells
NPV [EUR]
Deviation in average flow from low temperature wells[%]
NPV-Cooling Water Method
NPV-Low Temperature Method



x 10^9
Sensitivity analysis for copper extraction ratio
NPV [EUR]
Deviation in copper extraction ratio [%]
NPV-Cooling Water Method
NPV-Low Temperature Method



x 10^9
Sensitivity analysis for copper price
NPV [EUR]
Deviation in copper price [%]
NPV-Cooling Water Method
NPV-Low Temperature Method