

# Comparative Study of Thermal Waste Recovery Systems Deployed in Three Different Chemical Units

**Qamar, Rizwan Ahmed**

*Chemical Engineering Department, NED University of Engineering & Technology, Karachi, Sindh, PAKISTAN*

**Mushtaq, Asim\*<sup>+</sup>**

*Polymer and Petrochemical Engineering Department, NED University of Engineering & Technology, Karachi, Sindh, PAKISTAN*

**Ullah, Ahmed; Ali, Zaeem Uddin; Tarique, Syed Muhammad; Muzaffar, Ayesha; Jahangir, Fatima; Firdous, Tahira; Fatima, Aeliya**

*Chemical Engineering Department, NED University of Engineering & Technology, Karachi, Sindh, PAKISTAN*

**ABSTRACT:** Waste thermal energy is enough energy that is rejected to the atmosphere in the form of flue gases, streams of air, and liquid rejected from industries. It arises from the equipment, less efficient processes, and limitations due to thermodynamics' laws on operations. It is obvious that it is not possible to regenerate all waste energy, but most of the time, some waste heat can be used to achieve useful purposes. Waste heat recovery is the most important key to carrying out most of the research areas. The major areas of research and it is necessary to make the process more energy-efficient in chemical industries. To save energy, Heat Exchanger Network's (HEN) synthesis is essential. They are designed to reach energy targets. HEN design is the thermal integration between cold and hot utilities by pinch analysis at minimum temperature difference. HENs are important for utility saving because it helps in recovering heat from hot streams to others which reduces utility consumption and requirements. The heat exchangers are designed with simplified models for different industries using pinch technology. Most thermal recovery is obtained, and then some HEN network is required for a particular targeted area. In this research, improvements in energy recovery systems and HENs, and synthesis helps in capital savings, and pollutant emission can also be reduced.

**KEYWORDS:** Heat Exchanger Network's synthesis; Thermal recovery; Pollutant emission; Energy-efficient, Rankine cycle.

## INTRODUCTION

Cleaner and more feasible energy discovery has been a rapidly increasing approach because of global warming and an increase in price related to the conventional

resources of fuels. The energy utilities use by the earth are not infinite and have to end. The natural energy resources are petroleum, uranium, coal, and natural gas is ending the day

---

\* To whom correspondence should be addressed.

+ E-mail: engrasimmushtaq@yahoo.com

1021-9986/2022/4/1458-1500 43/\$/9.03

by day and will end within 50 years. The energy sources possess a negative impact on the environment because CO<sub>2</sub> is emitted and affects the atmosphere. An increase in the number of industries and fossil fuel use raises the emission of CO<sub>2</sub> by 30%. The excessive emission of CO<sub>2</sub> is directly affecting global warming. Global warming's negative effects involve the increase of sea level, higher interphase temperature, the yield of dislocation in agriculture, and a decrease in the snow of the northern hemisphere. The consequence of an increase in hazardous gas emissions is 22% per 10 years, with an increase in carbon dioxide. The adverse effect of using conventional energy sources is solved using renewable resources like solar, wind, hydro, wave, biogas, geothermal, and waste heat recovery [1-3]. In many industries, large quantities of thermal energy are produced as a byproduct. Huge quantities of thermal energy produce automobiles, heaters, nuclear power, the cement industry, thermal power plants, and the glass industry. Furthermore, thermal sources that are below 212°F are accessible from nature, for example, solar energy and geothermal. Using this thermal energy for the production of electricity can decrease a large quantity of money, less fuel prices, and will be useful to nature. WHR is important for making chemical process industries energy efficient. There are currently many waste heat recovery techniques at different levels of applications in oil and gas industries, petrochemical, and other industries, as shown in Table 1 [4, 5].

Today, almost 20 to 50% of the energy used in the industry is emitted into the environment as waste energy. A major part of this rejected heat is at less temperatures, which goes to the environment from cooling water and flue gases. Mostly, overhead streams of distillation at temperature ranges of 212-313°F removed heat from coolers, and the utilities that are at a temperature lower than 212°F removed the thermal energy from the cooling system [5, 6].

The main aim of the research is the provision of a complete analysis of thermal recovery systems and the best technology used for maximum recovery of heat and evaluation of the quantities and quality of the thermal sources. To identify recent barriers to heat recovery, to describe current waste heat recovery practices and the economic feasibility of the process. The energy is related to air, liquid, or flue gases which are typically industrial wastes, and the heat associated with them is the energy that is produced during the procedure that is not used in the

**Table 1: Typical sources of waste heat recovery.**

Sources of Waste Heat	Uses for Waste Heat
Combustion Exhaust	
Glass melting furnaces	Combustion air preheating
Boiler	Load preheating
Cement kiln	
Cooling Water From	
Air compressors	Power generation
Furnaces	Steam generation
Process Off-Gases	
Steel EAF	Boiler feedwater preheating
Aluminum furnace	Water preheating

application and is rejected and eventually wasted in the atmosphere. In the definition of rejected energy, it can say that the waste utilities are combined with air and groundwater, and the energy they contain within themselves becomes useless. The absorption of rejected heat energy by the atmosphere is often called thermal pollution.

#### **Factors affecting waste heat recovery**

The three major factors utilized in the quantification of rejected thermal energy are quality, quantity, and temporal availability. The thermal waste temperature is the most important matter for determining the feasibility of thermal recovery. Temperatures of these can change greatly. The quality of the waste heat can be roughly estimated as the temperature content of the waste utilities. The greater the temperature more will be the waste heat as a substitution for the energy and the easier optimization of the waste thermal recovery process. The use of a heat pump can increase the temp of thermal waste economically over a finite limit. For heat transfer and heat regeneration, this is important that the thermal source is at a higher temperature than the sink temperature.

Furthermore, the scale of change in temperatures among the heat source and heat sink is essential for determining streams of waste heat. The source and sink delta T affect the change of thermal energy transmitted per unit area of the exchanger and the max theoretical efficiency of converting heat from the energy of the thermal source to a different kind of energy. The amount of available thermal waste is denoted in the enthalpy of waste heat utilities [7-9].

$$H = m \cdot h \quad (1)$$

Where,  $H$  = Total enthalpy of the waste stream (unit: Btu/h),  $m$  = waste stream's mass flowrate (kg/h),  $h$  = Specific enthalpy of the waste stream (Btu/kg).

Enthalpy cannot be measured directly; it should be calculated with a reference point (for example, enthalpy of a matter at 25°C and atmospheric pressure at 1atm). It is how much thermal heat is available for thermal heat at the time of need. Comparing it with the load is very important for making thermal recovery effective. That's why much heat is useful and doesn't rely on the amount of heat but depends on the quality available that meets the requirement at the time of need or not [8, 10, 11].

### **Thermal recovery systems**

This method includes capturing waste heat and transferring it to the liquid or gas process, and sending recycled to increase the energy. The extra energy source can be utilized to generate electrical and mechanical power and additional heat. Related to two types of thermal waste and its sources, and to justify which type of heat recovery management must be utilized, it's important to know the quantity of recoverable thermal energy in industries. Many methods are there for the recovery of thermal waste, and that mainly comprised heat exchangers for thermal energy regeneration. All the heat recovery units work on the same principle capture, recovery, and utilization of the thermal waste with the amount of potential energy of the operation [8, 9, 12].

Heat exchangers are widely used for the transformation of heat from flue gases of combustion to the combust air which is going to the furnace. Thus, preheated air goes inside lower energy that is provided by fuel. Several technologies' for preheating air are furnace regenerators, rotary regenerators, recuperators, air preheaters, heat wheels, coil, recuperative burners, regenerative burners, plate heat exchangers, heat pipe heat exchangers, and direct electrical conversion devices. These are utilized to recover thermal waste from exhausting gas to implement normal to a higher temperature. Recuperators are made up of convection, radiations, or combination. Recuperators are made from metals or ceramics.

Boilers for the two-pass boilers that are used from normal to high temp flue gas to produce water vapors are thermal waste boilers. In cases where thermal waste isn't enough to generate levels of water vapors that are desired,

an afterburner or an auxiliary burner can add to achieve a high output of steam. Steam is mostly utilized for heating or power production. Superheated steam generation requires an extra heater for the system.

### **Optimization of Energy Systems**

The efficiency of the power of industries will be increased through good recovery methods & to decrease the emission of heat at a lower price. These techniques are the cause of reliable answers in many material industries, and they are expected to decrease energy consumption. Many concepts are put forward for the optimization of the process by a central resource system. Pinch analysis is a widely used and applied approach for process integration. A lot of techniques are there to optimize and design heat exchanger networking [13-15].

### **Problem statement and need for this study**

For developing a problem statement, a series of operations convert the raw input into useful output. The process requires utilities for heating and cooling. A unit that is not dependent and contains few or a lot of operations provided by utilities. An administrated region of most industrial activities. Targeted techniques were readily provided for the determination of thermal recovery at certain levels. The sole purpose of this report is the identification of efforts that are required to expand waste heat recovery practices. Several sources indicate a significant percentage (20 to 50%) of industrial energy inputs lost, approximating from 6 to 14 quadrillion Btu/yr. But still, there is a deficiency of details on the sources of greatest losses of waste heat in various industries, regions, multiple processes, and types of waste heat sources ( for example, the quality of thermal waste). The Energetics study, which was held in 2004, describes losses at different production levels of multiple products. It doesn't describe the quantity of waste heat but informs that these losses can comprise 20 to 50% of the total energy delivered to various plants. The study also uses roughly approximated efficiency improvement techniques to approximate that about 1.6 quadrillions Btu/yr can be saved through different heat recovery methods. PNNL (Pacific Northwest National Laboratory) also conducted a study in 2016 that included the estimation of chemical energy in wasted heat streams (the energy of unburnt or uncombusted gases like CO, CH<sub>4</sub>).

### Cost-efficiency of waste heat

Cost-effective waste thermal recovery involves identifying the source of thermal waste of optimal quantity, quality, temporal availability, and heat loads that can reuse the waste heat. Multiple industrial processes are available in the low to medium range of temperatures that can be reused, most of which are found in food industries, forest products, textiles, chemical, and petrochemical industries. The heat integration process will have an overall process implementation for the entire facility. There are proper strategies, tools, and methods available for identifying the most appropriate method of waste heat recovery while simultaneously ensuring optimal energy efficiency and process integration for the entire industrial facility.

### Benefits of waste heat recovery

Thermal recovery provides multiple usefulness to industry; all regenerated waste heat takes the place of the purchasing of energy which decreases energy costs. Re-usage of thermal waste permits less power conversion of components, causing less fixed expenditures. Waste Heat recovery decreases capital costs which decreases the operating costs of the equipment. As waste heat recovery took place in the purchasing of extra energy, it ultimately minimizes the impact on the environment on water and air. It also reduces GHG emissions that are directly related to plant operations.

### Economics of waste heat recovery and advancements in heat recovery technology

The economics of the thermal waste recovery system varies with the capital recovery that relies on the annual fuel saved. The most accurate kind of waste heat recovery equipment relies on feasibility, annual, and capital cost savings. It might be dangerous to use a simple payback period for economic analysis. In that place, appropriate discounted cash flow analysis must be analyzed for accuracy in comparison to options. Changes in the construction of heat pumps and heat exchangers have made new implementations and advancements in paybacks for numerous processes. Recent heat recovery units are made under special materials to bear higher temperatures and corrosion. As an example, the most widely used metallic radiation recuperators can't withstand inside temperatures that are increasing from 2000°F, but ceramic ones can bear temperatures of exiting gas up to 2800°F.

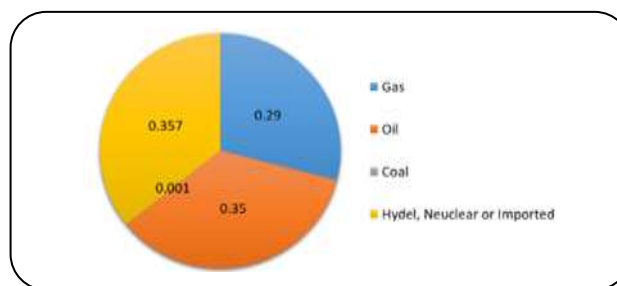


Fig. 1: Use of non-renewable energies

Current changes in heat pumps for capital cost reduction include improvements in compressors and heat exchangers [13, 14].

### Scope of study and waste heat recovery in Pakistan

Pakistan is facing horrible conditions related to energy in the past years to achieve electricity in the country and energy shortage; renewable energy resources are now being made. Maximum energy is used as an alternative to non-renewables because of their non-harming nature. There is a rapid growth in demand for electricity in Pakistan at 12–14% annually. In 2006, the total energy requirement was 58 million Tons of Oil Equivalent in Pakistan, and it can cross 180 million TOEs by 2021. The shortfall of electricity in Pakistan was approximately 8500MW in total in 2013. Pakistan imported 1399 million\$ in 1996 which negatively affected the economy of the country. 19% of the total expenditures of foreign exchange were used on importing non-renewable fuels.

In contrast, about 15 billion \$ was spent on non-renewable energy resources, equal to 39% of the entire import done in the country. The country's energy target is greater than the generation's, and it is estimated to increase three times by 2040. Pakistan is purely dependent on non-renewable energy, for example, thermal, water, and gas, to meet its energy target, as displayed in Fig. 1.

In the past ten years, Pakistan has suffered badly from energy shortfalls. As a result, to overcome this and improve economics, it's necessary to find and use renewable sources. Pakistan has green energy sources naturally, which are wind, solar, hydro, tidal, biogas, wave, geothermal, and WHR. Electricity is mostly generated from solar & wind that can't meet the power target, so it is important to use other commercially, like tidal waves, geothermal, biogas, biodiesel, and WHR. Pakistan's cement industry has about 12% consumption of the total energy of industries per year (720 MW). Depending on the

**Table 2: Sources of waste heat with temperatures**

Source of Waste Heat	Range of Temperature (°F)	Range of Temperature (°C)	Cleanliness
Heating system or furnace flue gases	590 - 1900	316 - 1,100	Variable
Flue gases from gas turbines	800 - 1,250	480 - 600	Uncontaminated
Jacket cooling water	185 - 225	90 - 100	Uncontaminated
Exhaust gases (that are used for gas fuels)	895 - 1,210	480 - 600	Most area is clean
Hot surfaces	150 - 600	65 - 316	Uncontaminated
Compressor heat in cooling water	100 - 175	40 - 90	Uncontaminated
Hot products	190 - 2,650	95 - 1,400	Most area is clean
Steam leaks or vents	300 - 625	115 - 322	Most area is clean
Condensate	145 - 535	70 - 275	Uncontaminated
Emission control devices that are thermal oxidizers	150 - 1,500	65 - 816	Mostly clean

technology and age of the cement plant, the mean use of electricity is from 80-130 kWh per year [15-17].

Energy for use in the industrial plant is spread over multiple subsystems categorized into three main types. Onsite power generation and other functions—such as steam, water, and air—are distributed and installed in the industry. Usage of process energy is spread through various plant systems such as refrigeration, process heating, and cooling, machine drive, electrochemical and other usages. The non-processed energy is used in facility heating, air conditioning and ventilation, support services, lighting, and on-site conveying.

Analysis of energy usage shows, that up to 80% of the total energy use of plants comes from on-site power production and applications of process energy. On-site power production and process energy use are also fields of significant loss of energy. For example, the loss of energy from process heating accounts for 25 to 55% of the total energy consumption. Because many forms of heat (such as convection or radiation, latent heat, and sensible) are emitted from the system and dissipated into the environment, more or less all the losses of energy from on-site power production and applications of processed energy will be lost. Recusing this wasted energy and reusing it in different industrial systems is the most effective way to lower the energy intensity of manufacturing plants. Table 2 shows the sources of waste heat with temperatures [3, 11, 13, 14].

#### **Previous studies and findings from previous reports**

Many reports from different departments and organizations such as the Department of Energy (DOE)

look at the heat sources that are wasted, mainly from different industry heating systems. Such reports range from approximating the losses of various industrial heating systems annually Brutus to examining thermal wasted energy in many industrial sectors and recognizing universal research and development options. Table 3 is the outline of some reports of waste heat [14].

BCS and PNNL published "Engineering Scope Study of Thermoelectric Generator (TEG) Systems for Industrial Waste Heat Recovery," a report in 2006 November. The report evaluates the TEG system focusing on industrial processes examined to quantify and identify industrial thermal sources of waste that can utilize TEGs. The surrounding of operation describes what a Thermo Electric Generator could come across in chosen industry processes and measures the performance of the Thermo Electric Generator system. Recognize design, economics, and execution demands required for Thermo Electric Generators to operate in particular industrial operations. Recognize development, and investigation required to get control over the restrictions that obstacle the development and the use of Thermo-Electric Generators for industrial heat waste recovery. The major three processes of industries were discussed for the examination of the applications of TEGs for waste heat recovery; aluminum Hall-Heroult cells (around 960°C), glass furnaces (485°C to 1,400°C), and reverberator furnaces (around 760°C). Examination based on opportunities, it is concluded that the application of TEG in glass furnaces would generate more than \$26 million in annual sales, assuming the higher efficiency of TEG [14, 18].

**Table 3: Energy systems and losses.**

Energy Systems	Energy Percent Wasted
Steam arrangements	Steam tubing and traps - 22% Boilers - 21% Steam transport/heat exchangers - 16%
Production of power	Conventional power - 46% (6200 Btu/kWh) Combined heat and power - 22% (4500 Btu/kWh)
Distribution of energy	Electricity and fuel distribution pipes and lines (not steam) - 3%
Conversion of energy	Heaters - 20% Onsite transport system - 45% Cooling systems - 15% Electrolytic cells - 10% Extra - 10%
Mechanical systems	Pumps - 50% Compressed air - 85% Fans - 30% Resources processing - 92% Resources handling - 3% Motor windings - 6% Refrigeration - 4%

Research on pasts experiments done in different studies accompanied by positive contact with industries and providers of equipment shave represented that a huge quantity of thermal energy was wasted and not utilized in a couple of ranges of temperature: 120°C and 870°C. There is a lack of heat recovery on a wide scale in both effective ranges which proves to be primarily because of problems accompanied by materials, technology, and economics.

#### **Thermal Waste in Major Industrial Sectors**

The research examines multiple industries with the availability and sources of wasted thermal energy. These implementations are possible for the recovery of the thermal energy waste under and over the plant, studies in different thermal waste recovery studies, and the hindrance to the valuable recovery of this great asset. Aluminum in melting, reusing, and primary production, glass in fiberglass manufacturing, paper in the paper mill, coatings in vinyl coating mill, steel in the steel mill, small mill or EAF mill, cement in dry process and shaft furnaces, food in food (snack) manufacturing, chemicals, and petroleum refining in major operations. The processes and operations of plants, different sources of rejected thermal energy, and the current application or disposal of wasted thermal streams. The energy is applied for utilities or in any extra area when the unused thermal energy or its transformed form of energy is then used for power generation or any other thing. Studies in unused thermal energy equipment designing (recycling). Economic

criteria or the respective requirements required for the justification and application of unused thermal power management thesis [18, 19].

#### **Typical waste heat streams in plant operations**

The waste heat flows via process industries are exhaust gases or flue gas, cold hydrogen oxide from heat exchangers, hot components, or unused side-products (solids, liquids). The interface of high-temperature heat-providing appliances involves primarily steam, hot gases, and the hot water supply sector. The exhaust gases or vapors can be classified as the hot flue gases from major industries and can be found useful in recovering heat as their steam is clean and does not provide any harm to the equipment by which try to reuse heat. Impurities or unwanted materials present in the stream sometimes cause huge problems, as they adversely affect the machines' surface and tubes. Waste gases containing more oxygen, mostly greater than 14%, cause no significant harm to the environment. The moisture content of such gases is also very less. Gases with the contamination that is combustible cannot be freely used and exposed to the environment. Its contamination can cause combustion even with fewer efforts which is very harmful to the forest environment mostly. Such gases are treated before their exit into the environment.

Uncontaminated hot water gets out from an alternate system of reducing temperatures such as steam condensers, product, or process cooling. This flow does not carry any sort of gaseous, solid, or liquid contamination. Material streams

carrying particular impurities such as small-sized granules or wasted cement particles should be filtered before removing the heat from such streams. The contamination causes a problem due to the increment of the heat capacity of water. Wash water from process industries usually contains solved impurities. Thus, require the pretreatment of these streams to make them pure and make the heat exchanger process effective.

Liquids and vapor products of the chemical industries are high in temperature. These can be used to generate electricity or to store or recover a great amount of heat. Solid products of very hot industries such as hot slabs and ash are cooled in an unmanaged manner, as they contain a variable range of heat every time. So, the environmental air and process water are used to cool them up. Some byproducts of industries from which heat is not recovered before their disposal to the environment. High-temperature surfaces can be categorized in the types of a large amount of heat lost due to the huge temperature range of some surfaces, for example, furnace walls and boiler walls [13, 19].

#### **Barriers to waste heat recovery**

Combustion products of high-temperature or hot flue gases that are comparatively clear. The lowered thermodynamic ability for efficient thermal recovery is because of restrictions on materials that need gases to be diluted. Heat transfer is limited on the side of flue gas in steam generation or any other power generation systems applications of heat exchangers. Flue gases at high temperatures or combustion products containing contaminants such as condensable vapors or particulates. Cost of materials and availability that are specially designed to control the contaminant's corrosive effects. Lack of design advancement that will allow self and pre-cleaning of thermal recovery equipment to reduce maintenance. Limitations of heat transfer on the gas side of heat exchanging equipment. Flue gases or heated air with high (>14%) O<sub>2</sub> without containing a large quantity of moisture content and particulates. Heat exchanger size limitations prevent retrofit usage, which can be because of design issues and heat transfer limitations. Lack of availability of combustion systems for small (<1 MM Btu/h) to use O<sub>2</sub> exhaust gases that are low as the air of combustion for fired systems [14, 16].

By-product gases or process gases and vapors are containing flammable in vapor or gaseous form. The absence of existing, economically sound vapor

concentrators for thermal waste heat recovery and then reusing materials that are not inorganic and combustible components that would remove the requirement to heat a huge quantity of dilution air and a huge machine size as a result. The deficiency of compact waste thermal recovery methods can minimize the heat exchanger like larger generators.

Make-up or process air is mixed with combustion products, and a large amount of moisture is combined with little quantity of particulates but not with organic vapors that can be liquefied. The prompt drop in performance of normal heat exchanger. Lack of availability of designs that allows cleaning of the surfaces of transfer of heat on equipment. The absence of design innovation allows the usage of condensing heat exchangers to have no corrosive effects.

Extra streams of gases are barriers specific to the implementation. Clean and hot water is released by a cold system which is indirect like the product or process cooling or steam condensers—the non-existence of usage of low-grade thermal energy inside the boundary of Equipment. The absence of economically sound systems can convert the heat of lower grades into a usable and transportable kind of energy. Heated water contains a huge quantity of impurities like particles coming from the process. No enormous practical hindrance to the cleaning of water. Scarcity of utilization of lower-grade heat inside the plant. Economically correct energy translation systems. Table 4 represents the current status of waste heat recovery technologies [9, 11, 14, 16].

#### **Process selection**

Heat recovery opportunities are there in different industries, especially in energy-intensive industries. The industries are categorized according to energy consumption into three types non-energy-intensive manufacturing industries, nonmanufacturing factories, and intensive manufacturing factories, as shown in Table 5 [14].

Industries use energy for different purposes like steam and cogeneration, heating and cooling of the process or process fluid, lighting, and heating. Energy consumption in industrial sectors varies by region and country, according to the differences in industries' gross output, the intensity of energy (consumption of energy per unit of gross output). Different industries' different compositions-improvement of processes of industries and increasing

**Table 4: Commonly used waste heat recovery system.**

Ultra-High Temperature (>870°C)	High Temperature (650 to 870°C)	Medium Temperature (315 to 650°C)	Low Temperature (120 to 315°C)	Ultra-Low Temperature (120°C)
Refractory (ceramic) regenerators	Convection recuperator (metallic) - mostly tubular	Convection recuperator (metallic) of many different designs	Convection recuperator (metallic) of many different designs	Shell and tube type heat exchangers
Heat recovery boilers	Radiation recuperator	The finned tube heat exchanger (economizers)	The finned tube heat exchanger (economizers)	Plate type heat exchangers
Regenerative burners	Regenerative burners	Shell and tube heat exchangers for water and liquid heating	Shell and tube heat exchangers for water and liquid heating	Air heaters for waste heat from liquids
Radiation recuperator	Heat recovery boilers	Self-recuperative burners	Heat pumps	Heat pumps
Waste heat boilers including steam turbine-generator based power generation	Waste heat boilers including steam turbine-generator based power generation		Metallic heat wheel	HVAC applications (recirculation water heating or glycol- water recirculation)
	Metallic heat wheels (regenerative system)			
	Load or charge preheating			

**Table 5: Energy-intensive industries.**

Industry type	Representative industries
Food	Manufacturing of beverages and food.
Refinery	Petroleum refineries and coal and gas feedstock
Iron and steel	Steel and iron manufacturing
Non-ferrous metals	Industries of aluminum, copper, tin zinc industries
Non-metallic minerals	Cement, lime, and glass industries
Basic chemicals	Resins, organic and inorganic chemicals, and agricultural chemical industries

process efficiencies by recovering waste heat and heat loss, increasing the use of cogeneration, and recycling materials and fuel inputs to reduce costs and improve efficiency. Amine gas contactor that operates at high pressure to ensure acid gas absorption in amine solvent typically operates at about 83 bar or 1200 psi. The amine regenerator operates at comparatively low pressure to remove the acid gas from the amine solvent and reproduce the lean solvent and reuse it with some makeup solvent in the gas contactor. Choose an amine-treating unit and recover heat from its inefficient manner to reduce wasted heat and make the process environment-friendly and economical. This process is a type of gas sweetening unit commonly used in refineries and gas plants to improve safety, meet environmental regulations, and prevent corrosion and deactivation of catalysts used in proceeding units.

#### **Process description of amine sweetening unit**

The amine gas sweetening process is extensively used to remove acid gases that are CO<sub>2</sub> and H<sub>2</sub>S from natural

gas. However, this process is intensive of energy because of the high requirements for heating and cooling. Reducing the energy requirements by optimization of operating parameters is limited. Therefore, modifications to the process can be made considerably, and a reduction in energy consumption results in reducing the operating cost of the process. The selection of an amine for gas sweetening is complex several process considerations should be encountered and analyzed before solvent selection. These factors are analyzed based on experimental data and a computer simulation program for gas sweetening called TSWEET. The possible solvent can vary between primary to tertiary amines, but a mixture of these amines can also be used.

Primary solvents are used to remove acid gas even at low pressure, but it degrades during the process, so it cannot be reused. An example of a primary amine is di glycol amine. Secondary amines remove acid gases at high pressure; one of the advantages of using secondary amine is the easiness in its regeneration as only by flashing



**Table 6: Composition of diethanolamine.**

Components	Mole fractions
DEA	0.27
Water	0.7282
CO <sub>2</sub>	0.002
Total	1.000

**Table 7: Composition of sour gas.**

Components	Mole fractions	Components	Mole fractions
Methane	0.8692	n-Hexane	0.0018
Ethane	0.0393	n-Heptane	0.0072
Propane	0.0093	Water	0.005
i-Butane	0.0026	Nitrogen	0.0016
n-Butane	0.0029	Carbon dioxide	0.0413
i-Pentane	0.0014	Hydrogen disulfide	0.0172
n-Pentane	0.0012	Di ethanolamine	0
Total	1.000		

that is reducing the pressure so the gases can be separated from the solvent. An example of secondary amine is diethanolamine (DEA). Tertiary amines have a selectivity toward hydrogen disulfide than carbon dioxide, which means that it absorbs more H<sub>2</sub>S than CO<sub>2</sub>, so it is used in the unit where process fluid has more hydrogen disulfide to carbon dioxide ratio. Solvents can be made by mixing different amines as one to enhance the absorbent characteristics as all the individual properties of each amine combines to give that single absorbent. For example, if tertiary amine is mixed with primary and secondary amines, it can improve the absorption rate of carbon dioxide in tertiary amines. The process uses a DEA potential of 28% in H<sub>2</sub>O as a solvent because the contactor is operating at high pressure. There is no need for selectivity of H<sub>2</sub>S over CO<sub>2</sub>, and Table 6 shows the composition of diethanolamine. The DEA temperature conditions are 95°F, the pressure is 995 psi, the molar flow rate is 4203 lb mol/h, and the phase is liquid [20, 21].

Sour gas at 100°F and 1000psia enters to the inlet gas scrubber V-100 removes entrained liquid before entering DEA contactor V-120. It was assumed the pressure drop in the separator was 0 psi for designing. The outlet gas is free from liquid entrainment and is routed to the DEA contactor. Table 7 shows the composition of sour gas. The inlet gas temperature condition is 86°F, the pressure is

1000 psi, the molar flow rate is 2745 lb mol/h, and the phase is a vapor [21].

The gas then enters the DEA Contractor V-120 from the bottom. Absorbent (DEA) is pumped into the column from the top. There are a total of 20 trays in the contractor. Raw gas encounters this solution while moving at the top of the column through the trays' sieves. The absorption of acid gases and heavy hydrocarbons takes place on these trays. Sweet gas exits the column from the top and flows towards dehydration units for the removal of moisture. The contactor's design operating conditions are an operating pressure 1000 psi, a top tray temperature is 95°F, and a bottom tray temperature is 140 °F. The strength of the DEA solution used is about 27%. Under this condition, absorption of acid, and gases are carried out concentrated DEA solution is removed from the bottom of the column and fed to the flash tank. The differential pressure between the inlet and outlet gases of DEA Contractor is to be maintained below ten psi at full bank load. DEA consumption is a maximum of 3.0 lb. per MMSCFD of purified gas.

In the flash tank, V-101 pressure is maintained at 90 psi. By flashing liquid from high to low pressure, any dissolved hydrocarbons are released and routed to the hydrocarbon flare line. Firstly, the liquid is passed through a valve at 1000psi, and its pressure is dropped to 90 psi. Then it enters vessels, and flashing occurs, which causes

**Table 8: Composition of rich and lean amine.**

Composition of rich amine		Composition of lean amine	
Components	Mole fractions	Components	Mole fractions
DEA	0.0573	DEA	0.0604
Water	0.9056	Water	0.9389
CO <sub>2</sub>	0.0263	CO <sub>2</sub>	0.0007
Methane	0.0001	Methane	0
H <sub>2</sub> S	0.0107	H <sub>2</sub> S	0
Total	1.000	Total	1.000

the removal of any hydrocarbons contained in the rich amine. The stripping of H<sub>2</sub>S and CO<sub>2</sub> occurs at approximately 240-270 °F temperature and 15-17 psi pressure. Desorption (stripping) of acid gas is carried out at a high temperature and low pressure. H<sub>2</sub>S/ CO<sub>2</sub>, along with water vapors, are stripped out from the top of the regenerator. Lean DEA is then removed from the downward of the regenerator. The composition of rich and lean amine is shown in Table 8 [5, 22].

## THEORETICAL SECTION

### Process description of the sulfur recovery unit

The feed of this plant includes flue gas from another plant, and the process is applied to remove the maximum possible sulfur content from the flue gas. The feed contains several components which are C<sub>7</sub> alkanes, H<sub>2</sub>O, H<sub>2</sub>S, CO<sub>2</sub>, and N<sub>2</sub>. The feed is entered into the separator. The flue gas is initially fed into the separator to separate the vapor and liquid phase. The two-stream getting out from the separator are water (liquid stream, mainly water with some quantity of CH<sub>4</sub> and CO<sub>2</sub>) and process gas (vapor stream). The process gas is now fed into the absorber, where it is absorbed in recycled lean amine (lean amine PO). After the absorption in lean amine, the two outlets steam rich amine and sweet gas. The absorber's function is to remove the sulfur content from the process gas, and after the removal of H<sub>2</sub>S, it becomes sweet gas, rich in amine.

Valve increases the pressure drop of rich amine. The outlet stream of the valve is rich out. The pressure becomes 704.8psia, and the temperature is slightly increased during this pressure variation. Rich out is led to the separator for the separation of liquid and vapor phases. The outlet streams of the separator are flash VAP and flash LIQ. The tube side stream is flash liquid, and the lean

amine is on the shell side. Flash LIQ after gaining temperature up to 208.4°F becomes 'Rich to REG.' The outlet of lean amine is Lean amine (HO) has a temperature of 197.3°F. The pressure drop of the heat exchanger is zero on both sides.

Rich to REG is entered as a feed into the regenerator. The top product of the regenerator is Acid gas, and the bottom product is a lean amine. The number of plates in the regenerator is 15; the feed plates are 9. H<sub>2</sub>S is mainly separated from lean amine, and the lean amine is sent back to the absorber after going through heat exchangers and pumps. Acid gas and air are fed into the conversion reactor. H<sub>2</sub>S and O<sub>2</sub> are reacted here for H<sub>2</sub>O and SO<sub>2</sub>. The outlet's streams are furnace gas and liquid product. Furnace gas is sent to the cooler, in which it is cooled from 2526°F to 599°F. The pressure drop is zero, and the duty of the cooler is 1.73\*10<sup>7</sup>. The cooled stream is named furnace gas out.

This stream is then sent to the separator. The separator's vapor outlet stream is catalytic converter feed1, and the liquid outlet stream is sulfur liquid 1. The properties of catalytic converter feed1 are a temperature is 599°F, a pressure is 14.69psia, a molar flow rate is 841.2lbmol/h, and a phase is a vapor. The sulfur dioxide and H<sub>2</sub>S are reacted to form H<sub>2</sub>O and sulfur. It has two outlet streams; one is liquid product1, and the other is catalytic converter1 gas. This stream is sent to the condenser, where 1.3% is converted into liquid. During this condensation, the process temperature is dropped to 302°F. This stream is now named cooled converter gas1. This stream is then sent to the separator. The separator's vapor outlet stream is catalytic converter2 feed, and the liquid outlet stream is sulfur liquid 2. This stream is sent to the heater, heated from 302°F to 599°F with zero

pressure drop. The heater duty is 713.8hp. The heated stream is named CC2 feed. The sulfur dioxide and H<sub>2</sub>S are reacted to form H<sub>2</sub>O and sulfur. It has two outlet streams; one is liquid product<sub>2</sub>, and the other is catalytic converter 2 gas. This stream is sent to the cooler, where its temperature is dropped to 302°F; the cooler duty is 826.1hp. This stream is now named cooled 3. This stream is then sent to the separator. The separator's vapor outlet stream is catalytic converter<sub>3</sub> feed, and the liquid outlet stream is sulfur liquid 3. This stream is sent to the heater, heated from 302°F to 500°F with zero pressure drop. The heater duty is 470.1hp. The heated stream is named CC3 feed [23, 24].

The sulfur dioxide and H<sub>2</sub>S are reacted to form H<sub>2</sub>O and sulfur. It has two outlet streams; one is liquid product 3, and the other is catalytic converter<sub>3</sub> gas. This stream is sent to the cooler named cooled CC gas, where its temperature is dropped to 302°F, and the cooler duty is 492.6hp. Then sent to the separator and the separator's vapor outlet stream is combustor feed, and the liquid outlet stream is sulfur liquid 4. Combustor feed is burned in air and fuel gas in the radiant zone of the furnace. The outlet stream is flue gas, applied many alternatives, and the main goal was to optimize the process and reduce or reuse the heat of flue gas. As the temperature of flue gas was very high, up to 778.7°F, it had great potential to provide us ground to achieve the targets. After applying many alternatives, it came to the result that not every approach is fruitful to optimize the process thermally. Methods represented a comparison study below that tried to recover the thermal waste of this plant [24].

#### **Rankine and Rankine with water**

In this first attempt, a goal was to use the heat of flue gas to increase the thermal energy of fluid heated in the boiler. This research used three different fluids that are water, n-butane, and i-pentane. All the fluids showed different behavior and different heat acceptance. The following data shows brief data to identify the best fluid, which reduced the maximum amount of heat and generated a significant amount of energy simultaneously. Flue gas from the plant was used to heat the liquid water into super-heated steam. This exchange of heat was done in the boiler of the Rankine cycle. As can observe that little change in the heat of the flue gas has put a significant amount of effect on the heat of water, increasing its temperature

to a great extent. This great difference in heat transfer is due to the huge change in the flow rates of both streams. The power generated by this Rankine cycle is 31.48 hp. Here flue gas from the plant was used to heat the liquid n-butane into super-heated steam. This exchange of heat was done in the boiler of the Rankine cycle.

Observation gathered from here told us that little change in the heat of the flue gas had put a significant amount of effect on the heat of n-butane, increasing its temperature to a great extent. This great difference in heat transfer is due to the huge change in the flow rates of both streams. The power generated by this Rankine cycle is 14.62 hp. The power generation and heat transfer from this system of the Rankine cycle here have used n-butane as an organic fluid have much lower efficiency as compared to the Rankine cycle in which water was used. Lastly, flue gas from the plant was used to heat the liquid i-pentane into super-heated steam. This exchange of heat was done in the boiler of the Rankine cycle. The last observation gathered here told us that little change in the heat of the flue gas had put a significant amount of effect on the heat of i-pentane, increasing its temperature to a great extent. This great difference in heat transfer is due to the huge change in the flow rates of both streams. The power generated by this Rankine cycle is 11.75 hp. The power generation and heat transfer from this system of the Rankine cycle here have used i-pentane as an organic fluid and have much lower efficiency as compared to the Rankine cycle which used water and n-butane. The Rankine cycle is an efficient method to optimize the plant and reuse the thermal waste, observing the behavior of flue gas when it was passed through the boiler of the Rankine cycle. Three different fluids were used for this experiment and concluded that water is the most efficient fluid among those three, as mentioned above. The power generated by water and the temperature gradient of flue gas were highest in the case of water, as shown in Table 9.

As already noticed that it was unable to extract a significant amount of heat from the flue gas by the application of the Rankine cycle even with three different fluids. Therefore, we have applied a different approach to decrease the remaining heat from the flue gas including the exchange between the flue gas and two different heat exchangers. This approach is also useful as it is diminishing the need for heaters that are employed between the separators and catalytic converters.

**Table 9: Temperature change in flue gas**

Fluid used	Change in temperature of flue gas (°F)	Power generated (hp)
Water	38.3	31.48
n-Butane	17.7	14.62
i-Pentane	16.7	11.75

The assembly is such that the outlet of separator V102 is heated by flue gas, and then it becomes the feed of catalytic converter 2. Similarly, the outlet of separator V103 is heated by the same flue gas that was used in the previous heat exchange as it still has much capacity to donate heat, and the heated outlet of separator V103 becomes the feed of catalytic converter 3. Flue gas in both cases is in the shell, and the fluid that is to be heated is in tubes.

As compared to the Rankine cycle here, the difference in the temperature of flue gas is much higher than that of all three Rankine cycle cases. Another benefit of this approach is the need for heaters is also cut, which is being a big economical saving. This heat exchanger network has collected thermal waste of 417°F. Therefore this approach is much better than the organic ranking cycle.

#### **Heat exchanger after Rankine cycle**

After observing the behavior of the approaches mentioned above, both extract the maximum amount of thermal waste. To do that, applied a Rankine cycle just after the flue gas outlet where water was used as the fluid to produce energy. The outlet of this Rankine cycle that is flues gas at 741.3°F was then used as the heating fluid for the two outlets of the separator in a similar manner described in the previous alternative. It was a collective target by using both methods to optimize the plant and save the environment simultaneously. Now, this flue gas goes to the heat exchanger. The third alternative is proved to be the most efficient one as here the temperature of flue gas is decreased to the maximum extent. Side by side, the need for heaters is also cut. Therefore, a Rankine cycle accompanied by the heat exchanger network should be applied in this plant to optimize it and make it economically sound.

#### **Process description of the benzene production unit**

Earlier techniques for the production of benzene involve the distillation of light oils produced during coke production. Benzene is manufactured by catalytic reforming/stream cracking of liquid petroleum feed, hydro dealkylation

of toluene (HDA), and toluene disproportionation (TDP). In some countries, this kind of aromatic compound is produced by the pyrolysis of gasoline coproduced in the steam cracking of naphtha, gas oil, or condensate to make unsaturated double-chain compounds. And in some, benzene is mostly produced from benzene. Naphtha and hydrogen are blended, and this is the feed for the reactor. The reactor contains a catalyst and is operated at 425°F to 530°F and 7 bars to 35 bars. Blends full of ring-shaped benzene derivatives are extracted from the reformat. All the above-discussed processes have one thing in common. Benzene has to be separated from the aromatic compounds that are achieved due to the two processes. Various separation processes are enervated to recover the benzene.

The process of selective disproportionation of toluene is making its place in most industries. Benzene in this process is produced along with the production of a paraxylene-rich xylene stream. There is also a process in which hydro-dealkylation of toluene is done either using catalytic or thermal processes. The cyclizer is the process that converts butane and propane into ring-structured compounds. There is only one plant that is in Saudi Arabia that has been running on this technology. There are two feeds in this process toluene and hydrogen gas. Initially, only toluene is fed into the tank. Hydrogen is added in a few further steps. Now the toluene is fed into the tank. Toluene entered into a tank where a portion vaporized and some portion is left as the liquid. Both streams (liquid and vapor) are separated from the tank. The liquid stream of toluene is then led to a pimp. The pump increases its pressure and the pump leads the S2 stream to the mixer. The vapor stream S1 is not used in the further process. After the action of the pump, the stream is now named S3. It can observe that the pressure of liquid toluene has increased from 27.56 psi to 374.2 psi.

Three streams are mixed in this mixture. The three streams are Liquid Toluene (S3), Hydrogen Gas, and a recycle stream of the separator. The hydrogen and recycle streams are in the vapor phase, and the toluene is in the liquid phase. The mixed stream or the outlet of the mixer is called S4. S4 is then preheated by using a heater which increases the stream temperature from 79.82°F to 437°F. The pressure of the stream remains constant in this heating process. The duty required by the heater is 8771kBtu/h. The heated stream is now called S5, and it is now a vapor phase. The steam is then fed into the furnace for further process. The furnace is used to greatly increase

the S5 stream temperature to a very great extent. S5, Fuel gas, and air are the inputs in the furnace. The fuel used here is comprised of 50% Hydrogen and 50% Methane. All the entering materials are in the vapor phase. Two streams are getting out from the furnace, S6, the required stream, and the flue gas. The minimum air-fuel ratio for the furnace is one, and the maximum is 40. Steady-state efficiency is 60%, and oxygen mixing efficiency is 100%.

Stream S6 and recycle streams are entered into a conversion reactor. Hydrogenation of toluene is done in this reactor. Stream S7 is an outlet from the top of the reactor and S8 is an outlet from the bottom of reactor S8. Now, S7 is fed into a cooler, which decreases its temperature to 100.4°F. Cooler duty required here is 18460 kBtu/h with a pressure drop of 1.45 psi. The stream is now S9. S9 is entered into the separator, where vapor and liquid phases are separated. The vapor stream is called S10 (main hydrogen with traces of benzene), and the liquid stream is called S11 (main toluene with a significant amount of benzene). S10 is divided into two parts with the help of Tee based on their flow rates. It is divided into streams S20 and S21 with flow rates 243.8 lb mol/h and 659.1 lb mol/h. S20 is entered into the compressor, where its pressure is increased to 369.8 psi. The required duty of the compressor is 5816 Btu/h. Fluid temperature is slightly increased during this compression. The outlet stream is S22. This S22 is again fed into a tee divided into two streams, S23 and S24, with flowrates 231.6 lb mol/h and 12.19 lb mol/h, respectively. S23 is recycled into the mixture where it is previously combined to form Stream S4. S24 is fed into the conversion reactor along with S6. Stream S21 is fed into the valve, where its pressure is dropped to 37.71 psi. During this pressure variation, a slight temperature elevation is observed. The outlet stream of the valve is called S25.

The separator's liquid outlet stream is entered into the valve, where its pressure is decreased to 42.06 psi. Temperature slightly is also increased, and some part of the stream is vaporized. The outlet stream is S12. S12 is then entered into the separator, where liquid and vapor are separated. S13 is a vapor outlet, and S14 is a liquid outlet. S14 is entered into the heater, where its temperature is increased to 194°F with slight vaporization. Heater duty is 842.2 kBtu/h. It is now S15. The stream S13 flow through the valve, where its pressure is decreased to 41.92 psi. The outlet of the valve is called S26. The feed of the component

**Table 9: Change in temperature of flue gas.**

Fluid used	Change in temperature of flue gas (°F)	Power generated (hp)
Water	888.7	83.14
i-pentane	881	82.33
i-butane	888.7	131.2

The splitter is S15. In this column, the component is split into two streams which are S16 (hydrogen and methane) and S17 (main toluene with a little amount of benzene). S16, S25, and S26 are mixed using this mixer, and they leave as flue gas. The feed of this column is S17. The distillate is S18 which is benzene (with traces of toluene), and S19 which is the bottom product and mainly consists of toluene with traces of benzene, the reboiler duty 1841 kBtu/h, and condenser duty 706.2 kBtu/h. The benzene stream is then cooled to 100.4 °F, the cooler duty is 4714 Btu/h, and the cooler's pressure drop is 2.901 psi, resulting in the final product.

This work applied many alternatives, and the main goal was to optimize the process and reduce or reuse the heat of flue gas. As the temperature of flue gas was very high, up to 1039°F, it had great potential to provide us with the ground to achieve the targets. After applying many alternatives, it came to a result that not every approach is fruitful to optimize the process thermally. In this first attempt, a goal was to use the heat of flue gas to increase the thermal energy of fluid heated in the boiler. This research used three different fluids that are water, i-butane, and i-pentane. All the fluids showed different behavior and different heat acceptance. Table 9 shows brief information that can identify the best fluid, which reduced the maximum amount of heat and generated a significant amount of energy simultaneously.

Flue gas from the plant was used to heat the liquid water into super-heated steam. This exchange of heat was done in the boiler of the Rankine cycle. As observed that little change in the heat of the water has put a significant effect on the temperature of flue gas. This great difference in heat transfer is due to the huge change in the flow rates of both streams and their enthalpies. The power generated by this Rankine cycle is 83.14 hp. Lastly, flue gas from the plant was used to heat the liquid i-pentane into super-heated steam. This exchange of heat was done in the boiler of the Rankine cycle.

The last observation gathered here told us that little change in the heat of i-pentane had a significant effect on the heat of flue gas, decreasing its temperature to a great extent. This great difference in heat transfer is due to the huge change in both streams' flow rates. The power generated by this Rankine cycle is 82.33 hp. Here flue gas from the plant was used to heat the liquid i-butane into super-heated steam. This exchange of heat was done in the boiler of the Rankine cycle. Observation gathered from here told us that i-butane temperature is changed much smaller than flue gas temperature, decreasing to a great extent. This great difference in heat transfer is due to the change in the flow rates of both streams. The power generated by this Rankine cycle is 131.2 hp. The power generation and heat transfer from this system of the Rankine cycle when using i-butane as an organic fluid have much higher efficiency as compared to the Rankine cycles in which water and i-pentane were used.

The Rankine cycle is an efficient method to optimize the plant and reuse thermal waste. Observed the behavior of flue gas when it was passed through the boiler of the Rankine cycle. Three different fluids were used for this experiment and concluded that i-butane is the most efficient fluid among those three, as mentioned above. The power generated by i-butane and the temperature gradient of flue gas was the highest among the three cases. We have applied another approach to decrease the heat from flue gas which includes the exchange between the flue gas and two different heat exchangers. This approach is also useful as it reduces the duty of heaters that are employed mixer and fired heaters another one is employed between the separators and component splitter. This alternative sent the flue gas to heat the stream coming out of the mixer and going into the heater. Heat this stream with flue gas to reduce the duty of the upcoming heater. Due to this alternative, the heater duty is reduced from 8771 kBtu/h to 4792 kBtu/h. This cooled flue gas is sent to another heat exchanger to reduce the heater's duty that is installed between the separator and component splitter.

Along with the single heat exchanger, it reduces another duty from 842 kBtu/h to 814 kBtu/h. This approach's benefit is that heaters' duty is also reduced, which is a big economical saving. This heat exchanger network has collected the thermal waste of 888 °F from flue gas.

### **Process model**

In Chemical Engineering, process modeling is defined as a technique to model or design processes. Aspen HYSYS software helps us to determine the behavior of a system in a steady state as well as in a dynamic state of interrelated components. It requires the characteristics and physical properties of the materials attached to a system to get better results. The processing time for convergence came out to be 10 seconds.

### **Components of amine sweetening plant**

The first step to starting a simulation in Aspen HYSYS is to identify and add all the components to a component list. After selecting all the list components, the HYSYS window is shown in Fig. 2 (a, b, and c) [6, 25, 26].

### **Classification of the property package**

Property packages are a group of methods to determine the properties of materials by calculating and estimating the value of parameters in any type of simulation. It is another name for thermodynamic models. After establishing a component list, the second step is to combine it with the appropriate property package; this merge of the component list and property package is then called a fluid package in Aspen HYSYS. These models are used to represent the behavior in different phases along with their energy levels of pure and a mixture of components. ASPEN HYSYS has more than thirty thermodynamic models classified in various ways, but four important classifications are, Equations of State, Semi-Empirical Models, Vapor Pressure Models, and Specialty Models. It includes Peng Robinson, PRSV, SRK, PR-Twu, SRK-Twu, (TST), Generalized Cubic Equation of State (GCEOS), MBWR, Lee-Kesler-Plöcker, BWRS, Lee-Kesler-Plöcker, Kabadi Danner, Sour PR/Sour SRK. The Vapor Pressure Model includes Modified Antoine Model, Braun K10 Model, and Esso K Model. The semi-Empirical Method includes the Chao-Seader model and the Grayson-Streed model. Specialty Model includes Glycol Package, Clean Fuels, OLI, Amines Models, and Steam Table Models [1, 26].

This study has used amine models for amine treatment and sulfur removal plants because both contain an amine component. The models used for amine plants are of three types acid gas physical solvents, liquid treating, and chemical solvents. In this model, lighter hydrocarbons are added according to the process's need, but diethanolamine

Component	Type	Group
Nitrogen	Pure Component	
CO2	Pure Component	
H2S	Pure Component	
Methane	Pure Component	
Ethane	Pure Component	
Propane	Pure Component	
i-Butane	Pure Component	
n-Butane	Pure Component	
i-Pentane	Pure Component	
n-Pentane	Pure Component	
n-Hexane	Pure Component	
n-Heptane	Pure Component	
H2O	Pure Component	
DEAmine	Pure Component	

(a)

Component	Type	Group
SO2	Pure Component	
SO3	Pure Component	
CO	Pure Component	
S_Liq_280	Pure Component	
H2O	Pure Component	
Methane	Pure Component	
Ethane	Pure Component	
Propane	Pure Component	
i-Butane	Pure Component	
n-Butane	Pure Component	
i-Pentane	Pure Component	
n-Pentane	Pure Component	
n-Hexane	Pure Component	
n-Heptane	Pure Component	
H2S	Pure Component	
CO2	Pure Component	
Nitrogen	Pure Component	
Oxygen	Pure Component	
DEAmine	Pure Component	

(b)

Component	Type	Group
Toluene	Pure Component	
Hydrogen	Pure Component	
Benzene	Pure Component	
Methane	Pure Component	
Oxygen	Pure Component	
Nitrogen	Pure Component	
H2O	Pure Component	
CO	Pure Component	
CO2	Pure Component	

(c)

Fig. 2: HYSYS window components (a) amine treatment unit (b) sulfur recovery unit (c) benzene-producing plant

cannot be added. It only supports DEPG. That's why this property package is not used in those processes. This property package accepts more amines than acid gas-physical solvents. It supports MEA, DEA, MDEA, DGA, MDEA + PZ, and use this property package, but there is another package that can accept a vast range of amines in it. This property package is best for a wide range of amine that's why to use it in the process. It supports (MDEA), Piperazine (PZ), (DEA), (DGA), (MEA), (DIPA), Triethanolamine (TEA), Sulflane + MDEA, Sulfolane + DIPA, Sulfolane+ PZ + MDEA, MDEA + DEA +MEA as shown in Fig. 3.

#### The benzene production plant

In the benzene-producing plant, used the SRK system

of equations, using the property package because it best suits the conditions as shown in Fig. 4. The assumptions are that it is a steady-state process, and heat losses are neglected by every piece of equipment.

## RESULTS AND DISCUSSION

### Amine treatment unit

#### Mass balance on a single heat exchanger

A simple mass balance provides data to be used in energy balance for calculation. Individual component mass transfer heat energy within the system giving rise to vapor phase formation. The mass balance of a single heat exchanger is shown in Fig. 5 (a).

Fig. 6 (a, b) shows the composition and mass flow of a single heat exchanger.

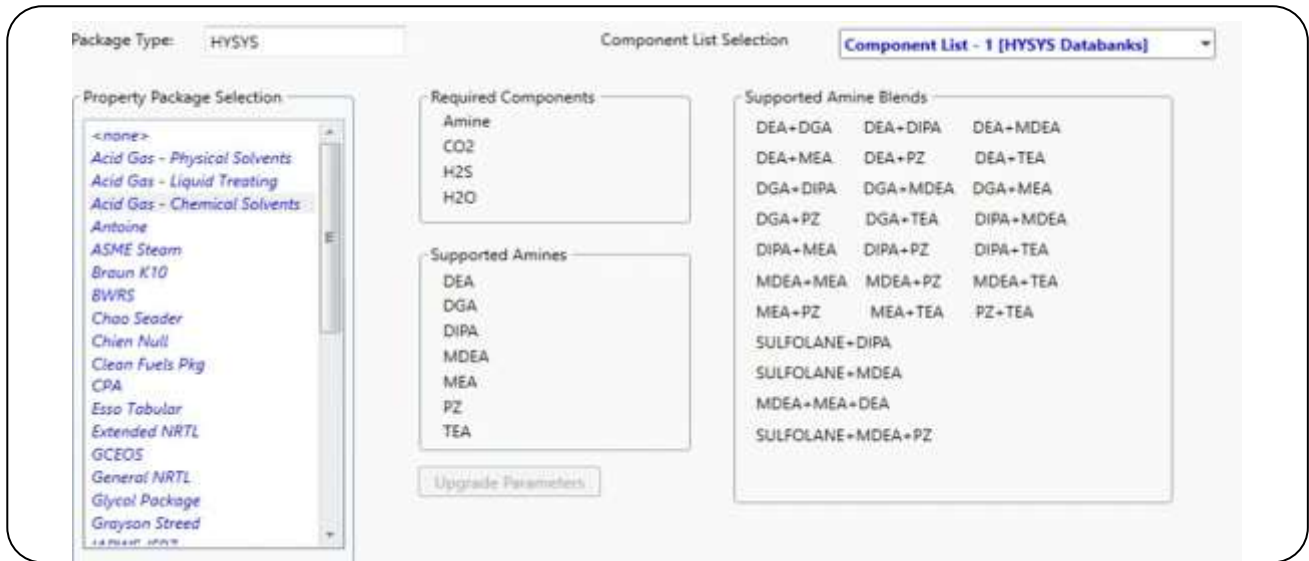


Fig. 3: Property packages for amine and sulphur plant.

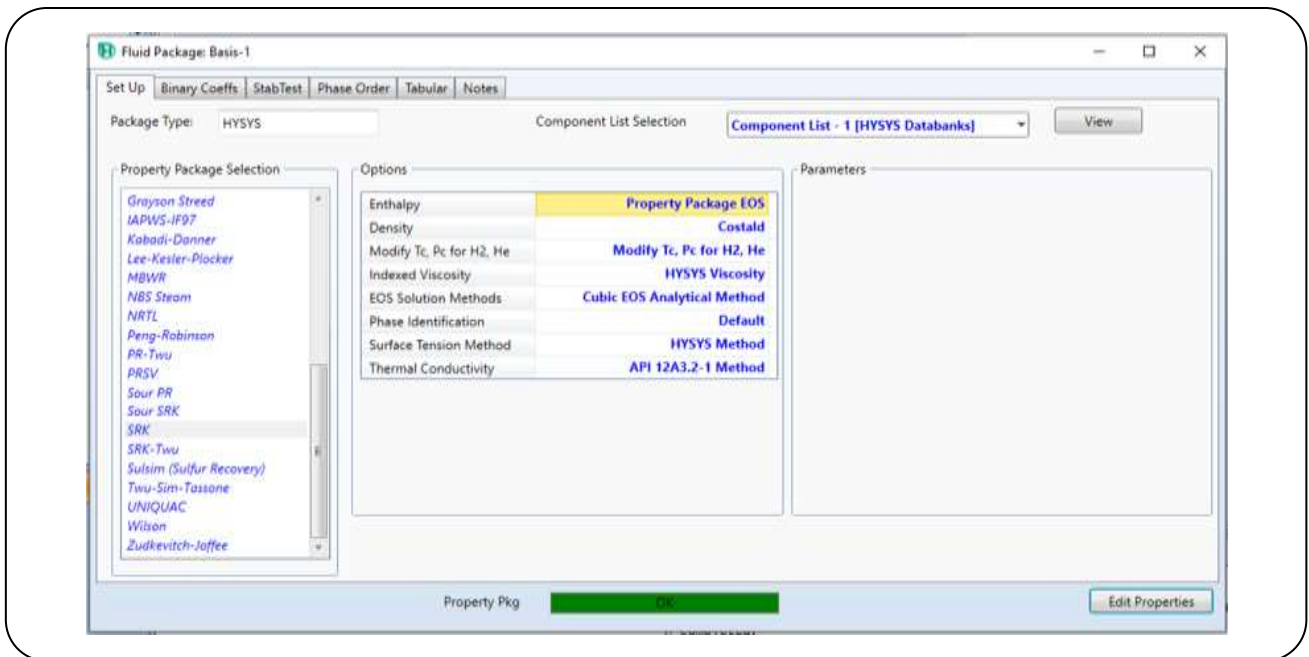


Fig. 4: Property package for benzene plant.

$$m_{in(Acid\ Gas)} + m_{in(Rich\ Amine)} = \tag{2}$$

$$m_{out(Acid\ Gas)} + m_{out(Rich\ Amine)}$$

$$3457lb/h + 47340lb/h = 3457lb/h + 47340lb/h.$$

$$99830lb/h = 99830lb/h.$$

The mass balance on the heat exchanger between the rich amine (outlet of exchanger 1) and the lean amine are shown in Fig. 5 (b). Fig. 7 (a, b) shows the compositions and mass flow between the rich and lean amine.

$$m_{in(Lean\ Amine)} + m_{in(Rich\ Amine)} = \tag{3}$$

$$m_{out(Lean\ Amine)} + m_{out(Rich\ Amine)}$$

$$95710lb/h + 91590lb/h = 95710lb/h + 91590lb/h.$$

$$187300lb/h = 187300lb/h.$$

The mass balance on the modified Rankine cycle is shown in Fig. 5 (c). Fig. 8 (a, b) compositions and mass flow of modified Rankine cycle.



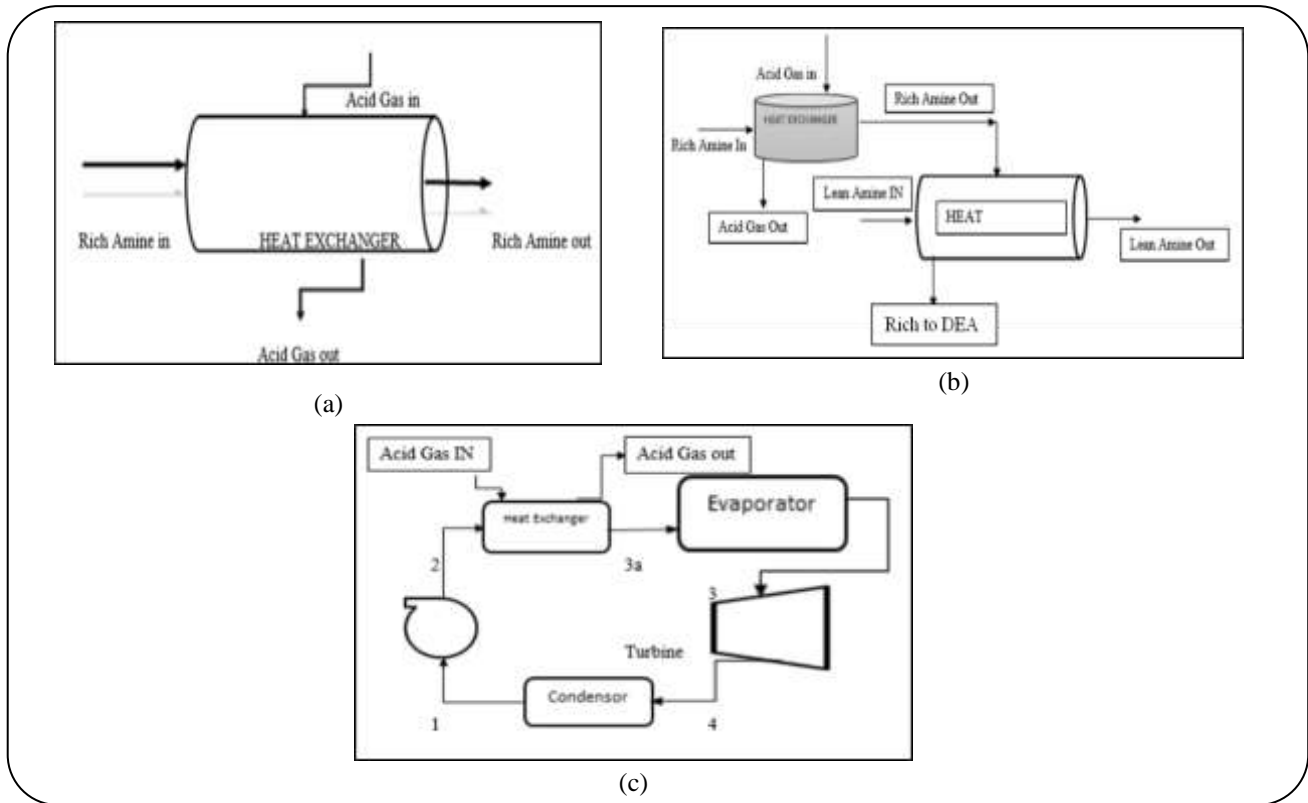


Fig. 5: Mass balance (a) single heat exchanger (b) Heat exchanger between acid gas and lean amine (c) Modified Rankine cycle.

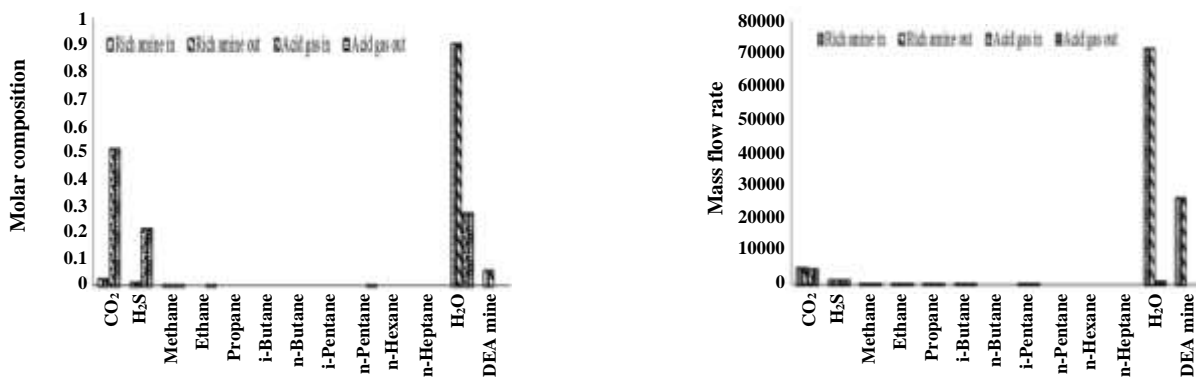


Fig. 6: (a) Compositions of the single heat exchanger (b) Mass flows of the single heat exchanger.

Defining the system mass balances, and composition from these calculations within the stream gives the accuracy of the specific heat capacity.

Energy balance

Heat exchanger between acid gas and rich amine as shown in Table 10 and Fig. 9 (a) [12, 27].

Sensible heat gain or loss within the stream is calculated as;

$$Q \text{ (BTU / hr)} = m C_p \Delta T \tag{4}$$

where; Q is the symbol for heat transfer, m is the mass of the substance, ΔT is the temperature change, and Cp stands for specific heat. Assuming no heat loss in the environment;

$$Q \text{ (Acid Gas)} = Q \text{ (Rich Amine)}$$

Moreover the calculation of heat transfer area for; [U= 1.779 BTU/h. ft<sup>2</sup> °F, ΔTm=29°F from simulation] gives;

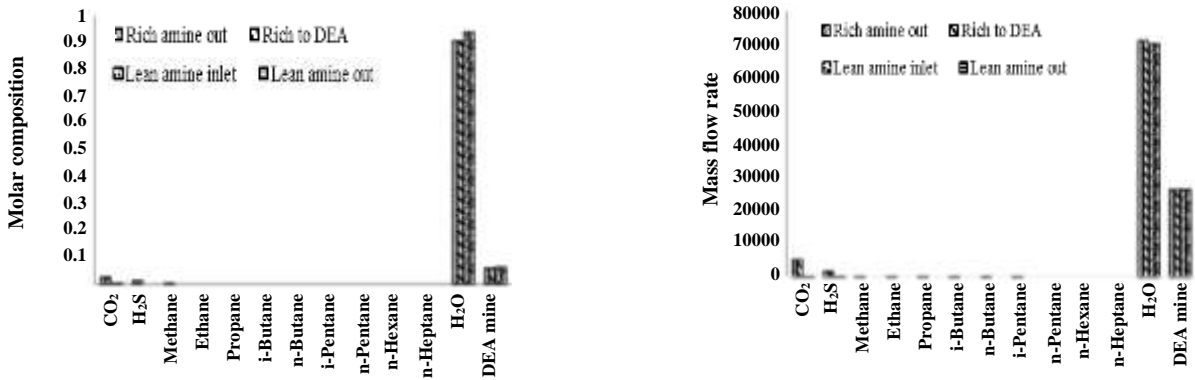


Fig. 7: (a) Compositions between the rich amine and lean amine (b) Mass flows between the rich amine and lean amine.

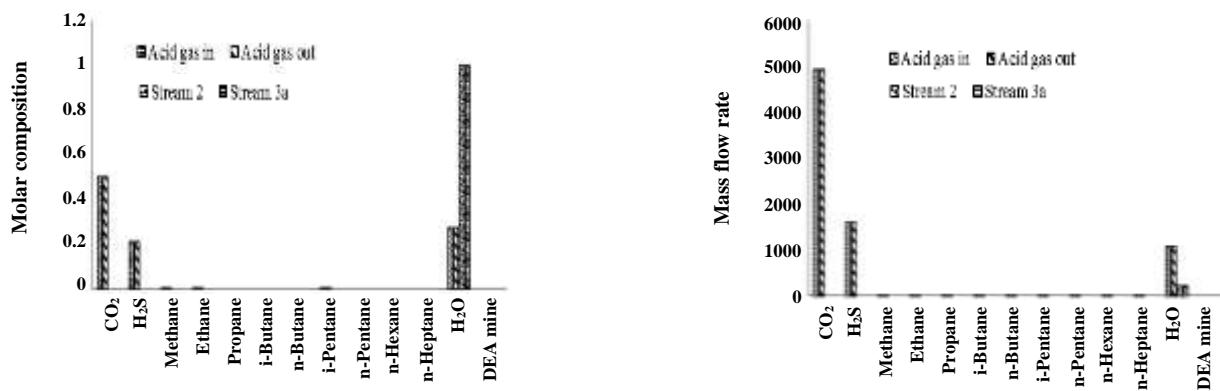


Fig. 8: a) Compositions of modified Rankine cycle (b) Mass flows of modified Rankine cycle.

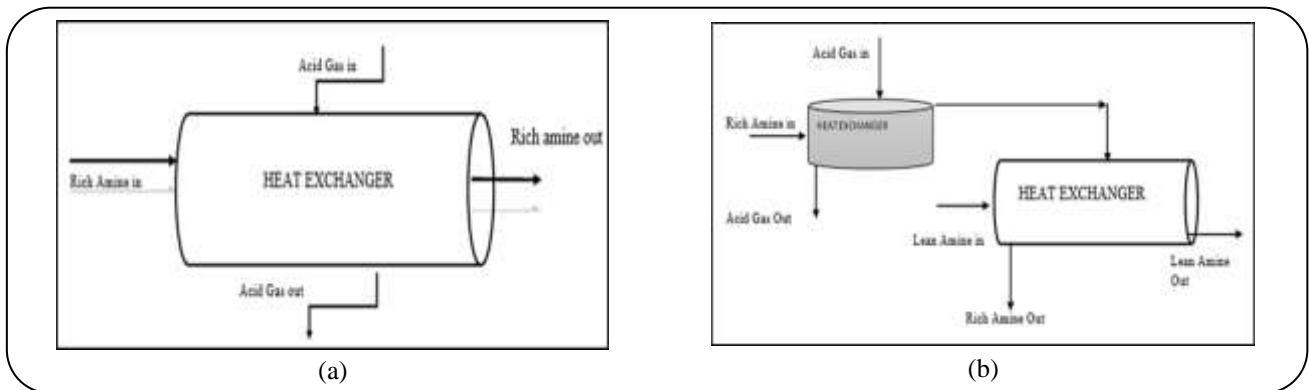


Fig. 9: The heat exchanger (a) between acid gas and rich amine (b) between the lean amine and rich amine

$$Q = UA\Delta T_m$$

Where;  $Q$  is the symbol for heat transfer,  $U$  is the overall heat transfer coefficient,  $A$  is the total area of heat transfer, and  $\Delta T_m$  is the effective mean temperature difference.

$$Q = 3.35E+04 \text{ BTU/h}, U = 1.779 \text{ BTU/h.ft}^2 \text{ }^\circ\text{F}, \Delta T_m = 29 \text{ }^\circ\text{F}$$

$$A = Q / U * \Delta T_m \tag{5}$$

$$A = 6.50E+02 \text{ ft}^2$$

Using Eq. (4) the heat flow is calculated to be

$$Q_{(RA)} = Q_{(LA)} = 5.27E+06$$

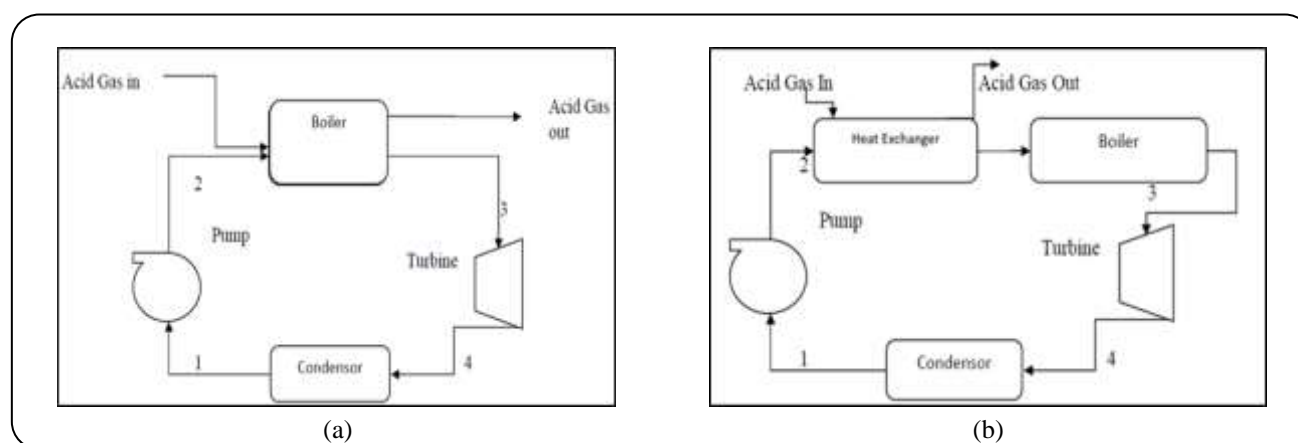
Heat exchanger area between rich amine (outlet of Heat exchanger 1) and the lean amine is also calculated from the simulation data as shown in Table 10 and Fig. 9 (b).

**Table 10: Heat balance of in heat exchanger.**

Stream	m (lb/h)	C <sub>p</sub> (BTU/lb °F)	T in (°F)	T out (°F)	ΔT (°F)
	Heat balance of acid gas and rich amine in the heat exchanger				
Acid gas	7.62E+03	0.229	180.7	161.5	19.2
Rich amine	1.04E+05	0.765	141	141.42	0.42
Heat balance between rich amine (outlet of exchanger 1) and lean amine					
Rich amine	1.04E+05	0.86	141.2	200	58.8
Lean amine	9.67E+04	0.972	256.7	200.6	56.1

**Table 11: Heat balance of Rankine cycle and modified Rankine cycle.**

Stream	Mass Flow (lb/h)	Mass Enthalpy (BTU/lb)	Enthalpy (BTU/h)
Heat balance of Rankine cycle			
Stream 1	220.5	6809	1501384.5
Stream 2	220.5	6804	1500282
Stream 3	220.5	6020	1327410
Stream 4	220.5	6238	1375479
Heat balance of modified Rankine cycle			
Stream 1	220.5	6809	1501384.5
Stream 2	220.5	6804	1500282
Stream 3	220.5	6020	1327410
Stream 4	220.5	6238	1375479
Stream 3a	220.5	6720	1481760

**Fig. 10: (a) Rankine cycle (b) Modified Rankine cycle.**

The Rankine cycle (at the outlet of the regenerator) is shown in Fig. 10 (a), and heat balance represents in Table 11.

Evaporator duty;  $Q_{(in)} = H_2 - H_3$   
 $Q_{(in)} = 1.7E+05$  BTU/h.

Table 12: Energy balance of additional exchanger.

	m (lb/h)	Cp (BTU/lb.°F)	T in (°F)	T out (°F)	ΔT (°F)
Acid gas	7622	0.241	180.6	170.1	10.5
Water	220.5	0.99	87.81	176	88.19

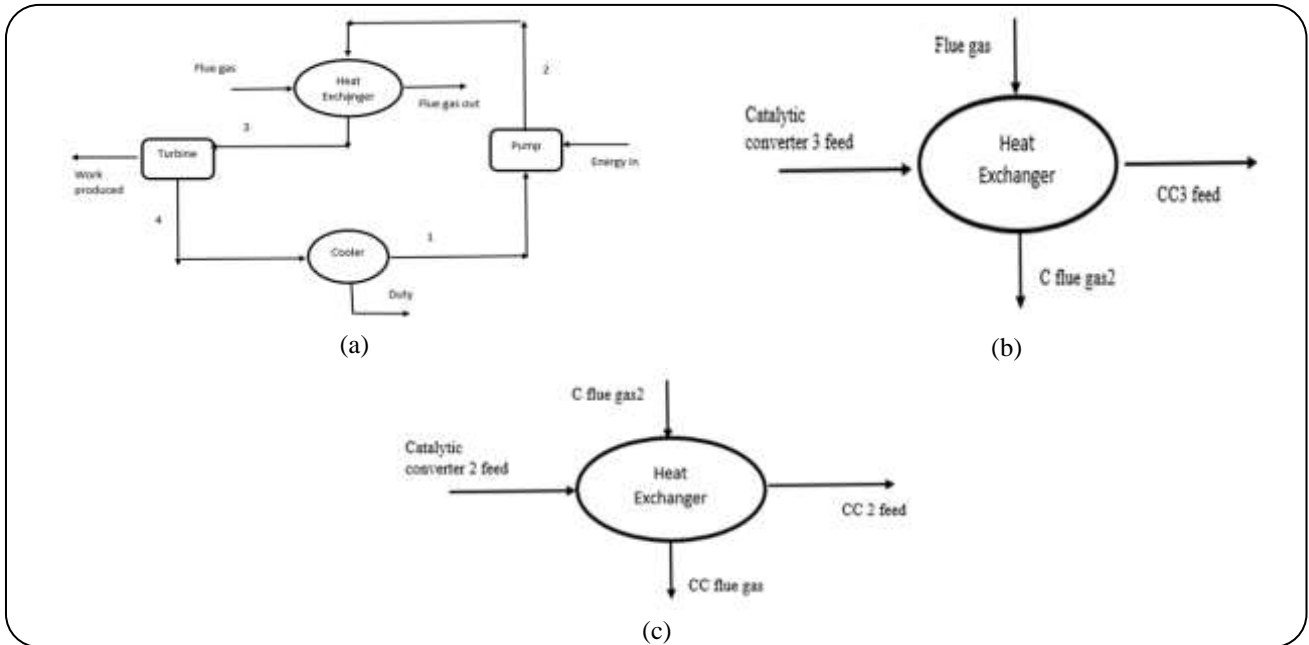


Fig. 11: (a) Rankine cycle (b) Heat exchanger between the flue gas and CC feed (c) Heat exchanger between flue gas two and CC feed

Condenser duty;

$$Q_{(out)} = H_1 - H_4$$

$$Q_{(out)} = 1.3E+05 \text{ BTU/h.}$$

$$\text{Efficiency; } \eta = 1 - [Q_{(out)} / Q_{(in)}] \quad (6)$$

$$\eta = 2.7E-01 = 27.16\%$$

Rankine cycle (with some modification) as shown in Fig. 10 (b) and heat balance is represented in Table 11.

$$\text{Boiler duty; } Q_{(in)} = H_{3a} - H_3 \quad (7)$$

$$Q_{(in)} = 1.5E+05 \text{ Btu/h.}$$

Comparison between both Rankine cycles;

$$\text{Energy saved} = Q_{(in)(3a)} - Q_{(in)(3b)} \quad (8)$$

$$\text{Energy saved} = 1.9E+04 \text{ BTU/h.}$$

The energy balance using the additional exchanger is shown in Table 12.

$$Q_{(acid \text{ gas})} = mC_p\Delta T \quad \text{BTU/h.}$$

$$Q_{(acid \text{ gas})} = 1.9E+04 \text{ BTU/h.}$$

$$Q_{(water)} = mC_p\Delta T \quad \text{BTU/h.}$$

$$Q_{(water)} = 1.9E+04 \text{ BTU/h.}$$

$$Q_{(Acid \text{ gas})} = Q_{(water)} = Q_1$$

$$\text{Now; } Q_{(in)(3a)} = Q_2 = 1.7E+05 \text{ BTU/h.}$$

$$Q_{(in)(3b)} = Q_3 = 1.5E+05 \text{ BTU/h.}$$

$$\text{Therefore, } Q_1 + Q_3 = Q_2$$

$$1.7E+05 = 1.7E+05 \text{ BTU/h.}$$

After applying a dual heat exchanger, the external power requirement is 1.542E+05 BTU/h instead of 1.7E+05 BTU/h. Amount of power saved = 1.9E+04 BTU/h.

### Material Balance of Sulfur Recovery Unit

We designed five alternatives to optimize the plant. Three alternatives are based on Rankine cycles using different fluids, as shown in Fig. 11 (a).

In this alternative, have exchanged flue gas heat for i-pentane in the heat exchanger. All the other equipment except the heat exchanger contains i-pentane, so component-wise mole fractions and mass flow rates are not mentioned as they all equal zero. That is why the only mass balance on the heat exchanger is as Fig. 12 (a, b).

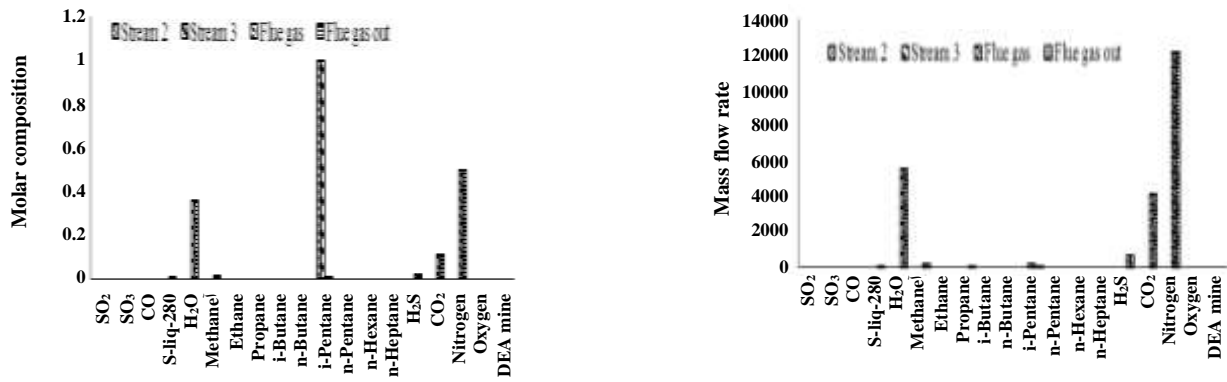


Fig. 12: Rankine with i-pentane (a) Molar compositions (b) Mass flow rate.

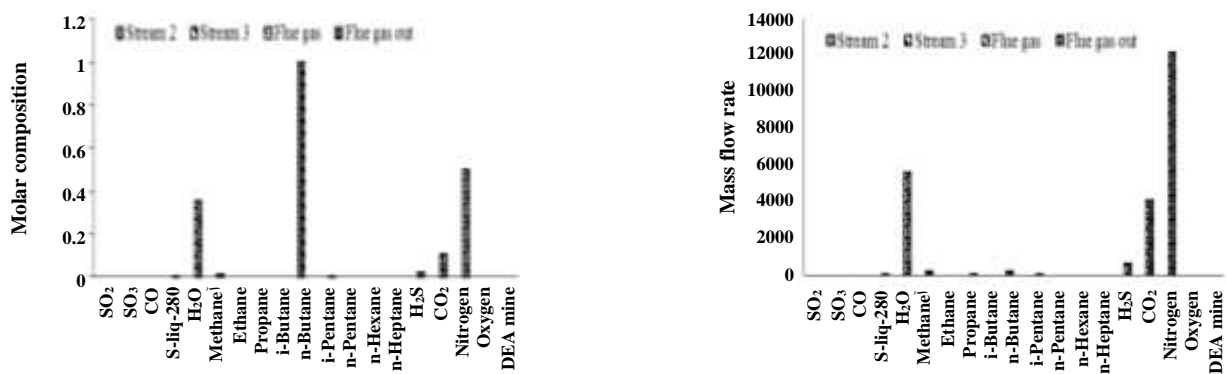


Fig. 13: Rankine with n-butane (a) Molar compositions (b) Mass flow rate.

The mass flow rate in □ Mass flow rate out

$$m_{(\text{flue gas})} + m_{(2)} = m_{(\text{flue gas out})} + m_{(3)} \tag{9}$$

$$22728.61951 \text{ lb/h.} + 220.46 \text{ lb/h.} \square 22728.61951 \text{ lb/h.} + 220.46 \text{ lb/h.}$$

In this alternative, have exchanged flue gas heat for n-butane in the heat exchanger. All the other equipment except the heat exchanger contains n-butane, so component-wise mole fractions and mass flow rates are not mentioned as they are all equal to zero. That's why the only mass balance on the heat exchanger is as Fig. 13 (a, b).

Mass flow rate in □ Mass flow rate out

$$m_{(\text{flue gas})} + m_{(2)} \square m_{(\text{flue gas out})} + m_{(3)}$$

$$22728.61951 \text{ lb/h} + 220.46 \text{ lb/h} \square 22728.61951 \text{ lb/h} + 220.46 \text{ lb/h}$$

In this alternative, have exchanged flue gas heat for water in the heat exchanger. All the other equipment except the heat exchanger contains water, so component-wise mole fractions and mass flow rates are not mentioned

as they are all equal to zero. That is why the only mass balance on the heat exchanger is as Fig. 14 (a, b).

Mass flow rate in □ Mass flow rate out

$$m_{(\text{flue gas})} + m_{(2)} \square m_{(\text{flue gas out})} + m_{(3)}$$

$$22728.61951 \text{ lb/h} + 220.46 \text{ lb/h} \square 22728.61951 \text{ lb/h} + 220.46 \text{ lb/h}$$

This alternative replaced two heaters with two heat exchangers, which utilized hot flue gas to heat process gases inside the plant, as shown in Fig. 11 (b). The molar compositions and mass flow rate of the catalytic converter with feed as shown in Fig. 15 (a, b).

The mass flow rate in □ Mass flow rate out

$$m_{(\text{flue gas})} + m_{(\text{catalytic converter 3 feed})} = \tag{10}$$

$$m_{(\text{c flue gas 2})} + m_{(\text{cc3 feed})}$$

$$22728.32951 \text{ lb/h} + 19534.1046 \text{ lb/h} \square 22728.32951 \text{ lb/h} + 19534.1046 \text{ lb/h}$$

Fig. 11 (c) is shown the heat exchanger between the flue

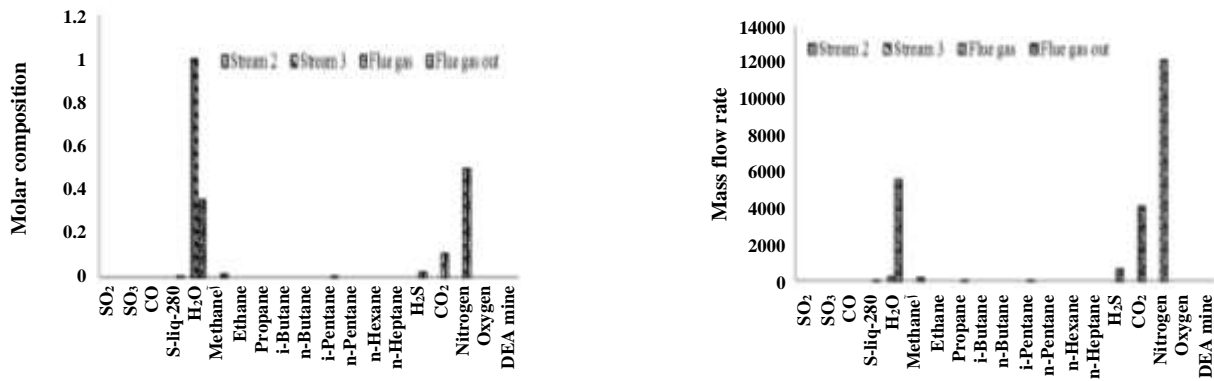


Fig. 14: Rankine with water (a) Molar compositions (b) Mass flow rate.

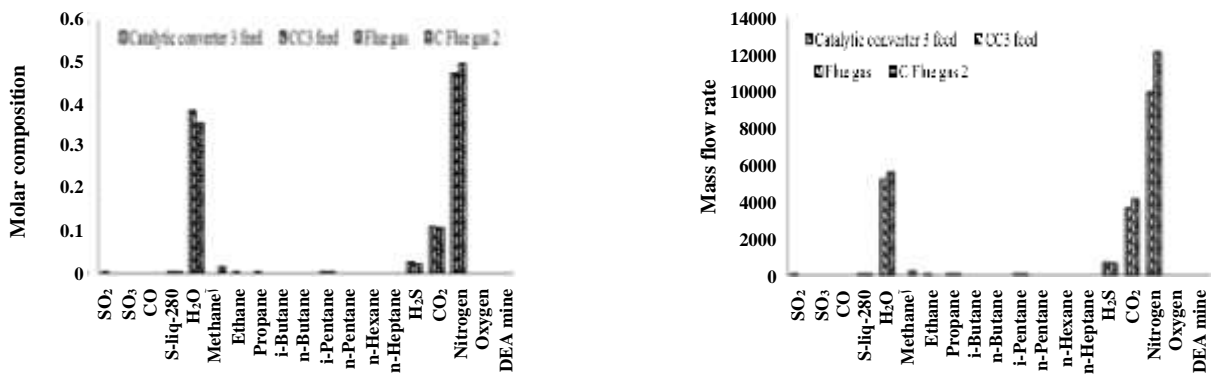


Fig. 15: Catalytic converter 3 with feed (a) Molar compositions (b) Mass flow rate.

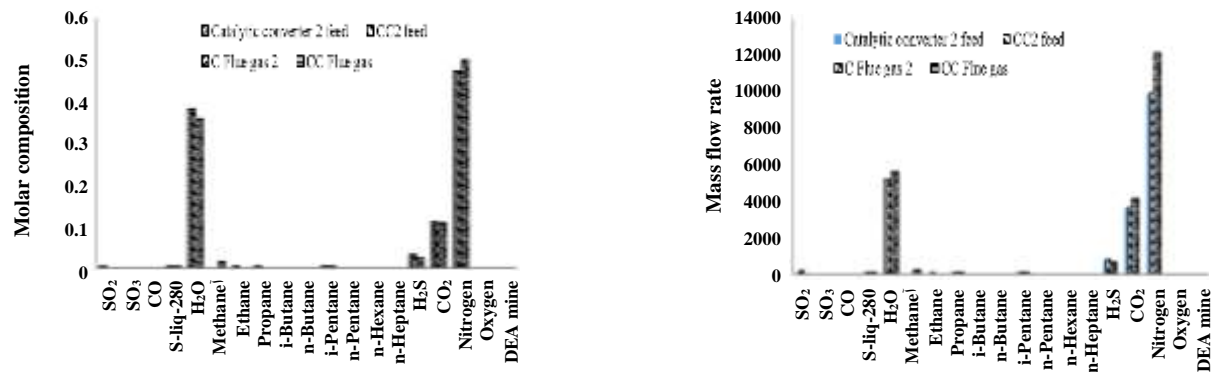


Fig. 16: Catalytic converter 2 with feed (a) Molar compositions (b) Mass flow rate.

Fig. 16 (a, b) shows molar compositions and mass flow rate of catalytic converter 2 with feed.

The mass flow rate in  $\square$  Mass flow rate out

$$m_{(c \text{ flue gas } 2)} + m_{(\text{catalytic converter 2 feed})} =$$

$$m_{(\text{cc flue gas})} + m_{(\text{cc2 feed})}$$

$$22728.3295 \text{ lb/h} + 19719.27077 \text{ lb/h} \square 22728.3295 \text{ lb/h} + 19719.27077 \text{ lb/h}$$

**Energy balance of Rankine with water**

Fig. 17 (a) shows the Rankine with water. Boiler duty;  $Q_{in} = H_2 - H_3 = 1.5 \cdot 10^6 - 1.219 \cdot 10^6 = 283 \cdot 10^3 \text{ Btu/h}$ .

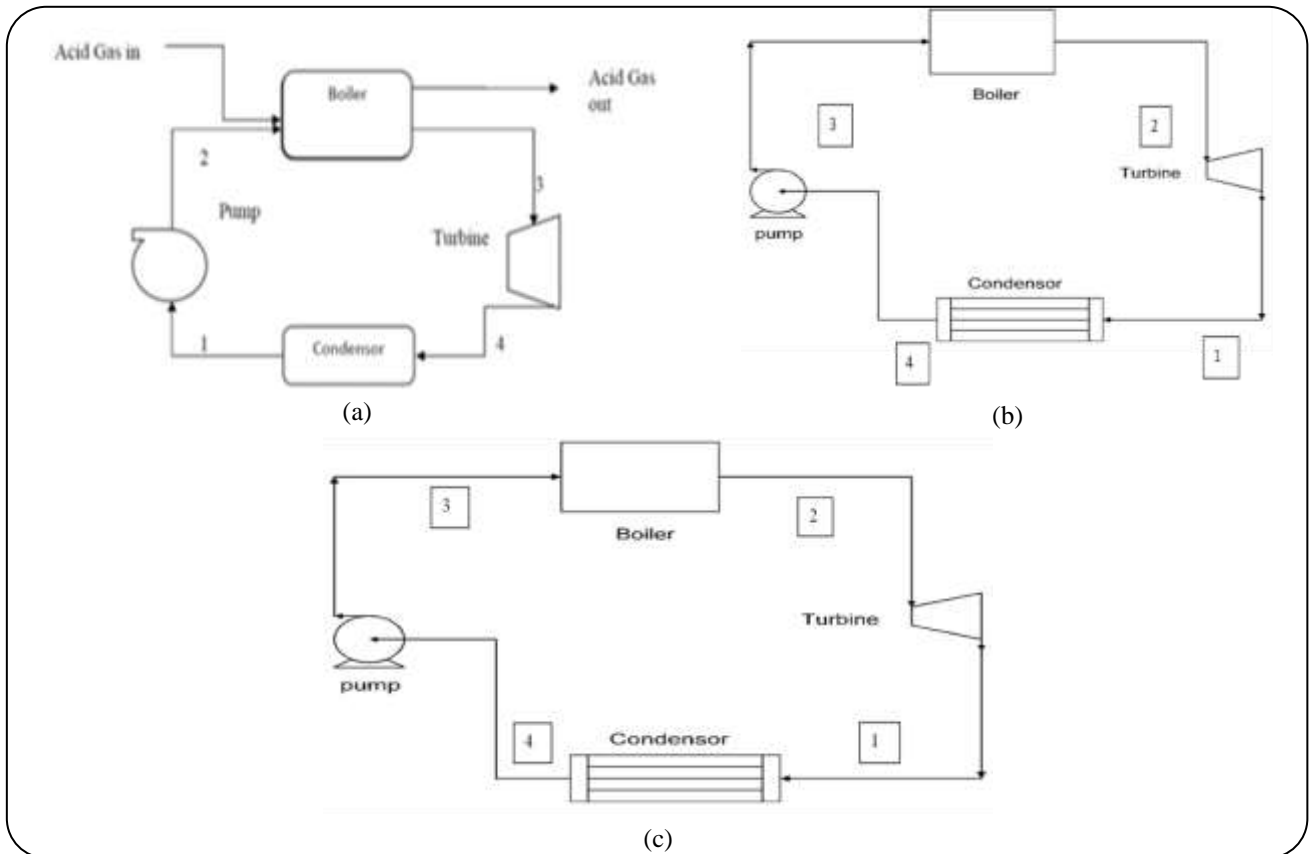


Fig. 17: Energy balance (a) Rankine with water (b) Rankine with n-butane (c) Rankine with i-pentane

Condenser duty;  $Q_{out} = H_1 - H_4 = 1.501 \cdot 10^6 - 1.297 \cdot 10^6 = 204 \cdot 10^3 \text{ Btu/h}$ .

Efficiency;  $\eta = 1 - Q_{out}/Q_{in} = 0.279 = 27.9\%$

Table 13 shows the different energy balances on Rankine with water.

Fig. 17 (b) shows Rankine with n-butane. Evaporator duty;  $Q_{in} = H_2 - H_3 = 2.551 \cdot 10^5 - 1.612 \cdot 10^5 = 131.1 \cdot 10^3 \text{ Btu/h}$

Condenser duty;  $Q_{out} = H_1 - H_4 = 2.567 \cdot 10^5 - 1.612 \cdot 10^5 = 95.5 \cdot 10^3 \text{ Btu/h}$ .

Efficiency;  $\eta = 1 - Q_{out}/Q_{in} = 0.2715 = 27.15\%$

Fig. 9 (c) shows Rankine with i-pentane. Boiler duty;  $Q_{in} = H_2 - H_3 = 2.441 \cdot 10^5 - 1.210 \cdot 10^5 = 123.1 \cdot 10^3 \text{ Btu/h}$ .

Condenser duty;  $Q_{out} = H_1 - H_4 = 2.454 \cdot 10^5 - 1.509 \cdot 10^5 = 94.5 \cdot 10^3 \text{ Btu/h}$ .

Efficiency;  $\eta = 1 - Q_{out}/Q_{in} = 0.2323 = 23.23\%$

In conclusion, the turbine produces a power of 31.48hp or  $8.01 \cdot 10^4 \text{ Btu/h}$ . Boiler duty is  $283 \cdot 10^3 \text{ Btu/h}$ . The turbine produces a power of 14.62hp, and the boiler duty is  $131.1 \cdot 10^3 \text{ Btu/h}$ . The turbine produces a power of 11.75hp, and the boiler duty is  $123.2 \cdot 10^3 \text{ Btu/h}$ .

#### Sulfur in heat exchangers

Exchanger 01;  $Q$  (catalytic converter feed) =  $Q$  (flue gas)

$mC_p \Delta T = mC_p \Delta T$

$$(1.953 \cdot 10^4) \cdot (0.3092) \cdot (500 - 302) =$$

$$2.273 \cdot 10^4 \cdot (0.32215) \cdot (500 - 302)$$

$$1.196 \cdot 10^6 \text{ Btu/h} = 1.196 \cdot 10^6 \text{ Btu/h}$$

Exchanger 02;  $Q$  (catalytic converter feed) =  $Q$  (flue gas)

$mC_p \Delta T = mC_p \Delta T$

$$(1.972 \cdot 10^4 \text{ lb/h}) \cdot (0.3101 \text{ Btu/lb}^\circ\text{F}) \cdot (599 - 302)^\circ\text{F} =$$

$$2.273 \cdot 10^4 \text{ lb/h} \cdot (0.3132 \text{ Btu/lb}^\circ\text{F}) \cdot (616.4 - 361.3)^\circ\text{F}$$

$$1.816 \cdot 10^6 \text{ Btu/h} = 1.816 \cdot 10^6 \text{ Btu/h}$$

#### Material Balance of Benzene production unit

This research has designed five alternatives to optimize the plant. Three of the alternatives are based on Rankine cycles which used different fluids. Rankine with water in this alternative exchange flue gas heat for water in the heat exchanger. All the other equipment except the heat exchanger contains water, so component-wise mole fractions and mass flow rates are not mentioned as they all

Table 13: Energy balance on Rankine with water.

Stream	Rankine with water		Rankine with n-butane		Rankine with i-pentane	
	Mass Enthalpy (Btu/lb)	Enthalpy 1*10 <sup>6</sup> (Btu/h)	Mass Enthalpy (Btu/lb)	Enthalpy 1*10 <sup>6</sup> (Btu/h)	Mass Enthalpy (Btu/lb)	Enthalpy 1*10 <sup>6</sup> (Btu/h)
1	6809	1.501	1164	2.567	1113	2.454
2	6804	1.5	1157	2.551	1107	2.441
3	5520	1.217	562.3	1.24	548.7	1.210
4	5883	1.297	731	1.612	684.3	1.509

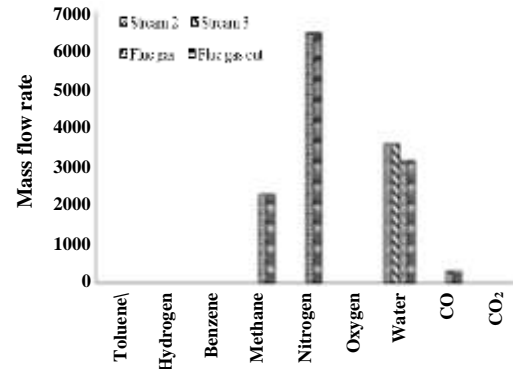
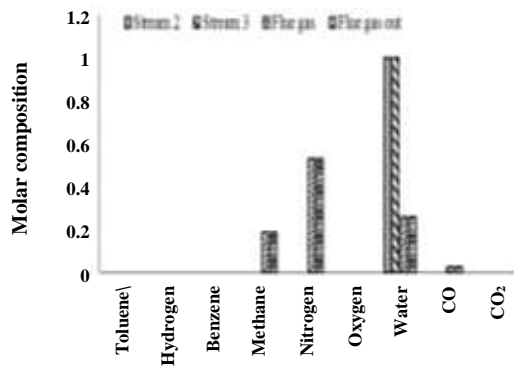


Fig. 18: Rankine with water (a) Molar composition (b) Mass flow rates.

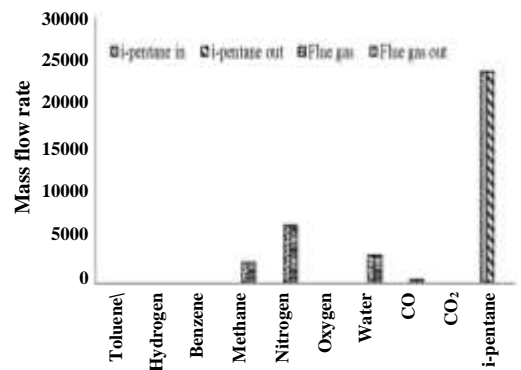
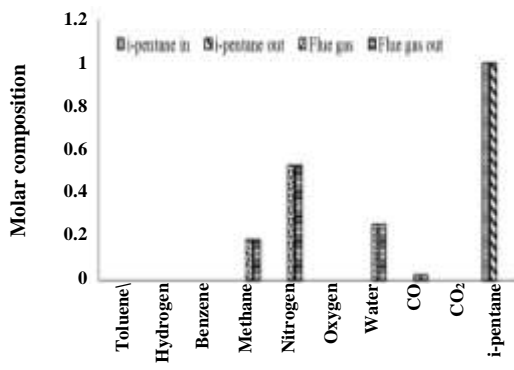


Fig. 19: Rankine with i-pentane (a) Molar composition (b) Mass flow rates.

equal to zero. That is why only the mass balance on the heat exchanger is shown in Fig. 18 (a, b).

The mass flow rate in □ Mass flow rate out

$$m_{(flue\ gas)} + m_{(2)} \square m_{(flue\ gas\ out)} + m_{(3)}$$

$$12167.156lb/h + 3574.44lb/h \square 12167.156b/h + 3574.44lb/h.$$

$$15741.596lb/h = 15741.596lb/h.$$

Rankine with i-pentane in this alternative exchanged flue gas heat for i-pentane in the heat exchanger. All

the other equipment except the heat exchanger contains i-pentane, so component-wise mole fractions and mass flow rates are not mentioned as they are all equal to zero, as shown in Fig. 19 (a, b) [3, 11, 14].

The mass flow rate in □ Mass flow rate out

$$m_{(flue\ gas)} + m_{(i-pentane\ in)} \square m_{(flue\ gas\ out)} + m_{(i-pentane\ out)}$$

$$12168lb/h + 23860lb/h. \square 12168b/h + 23860lb/h.$$

$$36028/h = 36028lb/h$$

Rankine with n-butane in this alternative has been exchanged



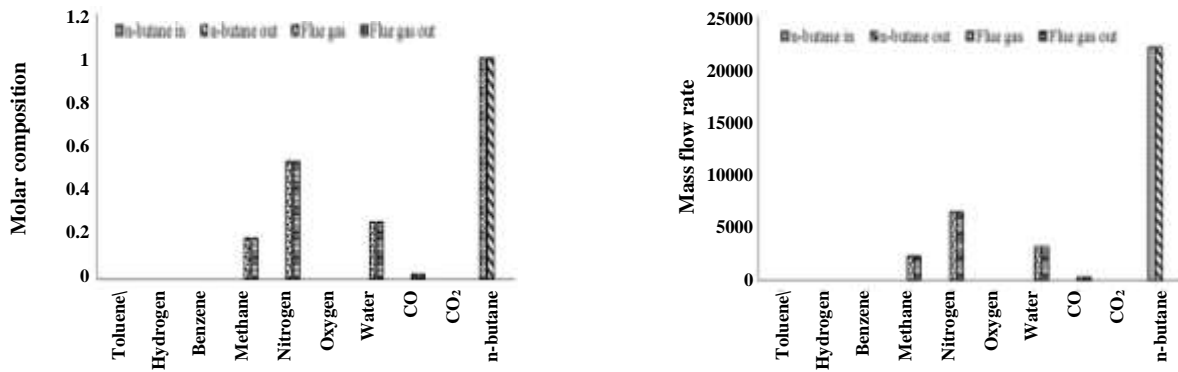


Fig. 20: Rankine with n-butane (a) Mole fractions (b) Mass flow rate.

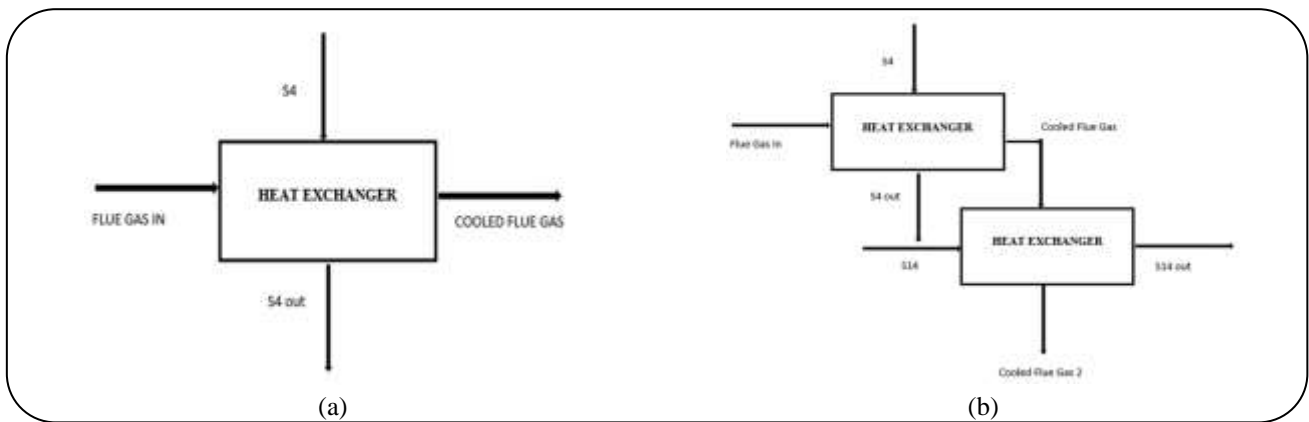


Fig. 21: Heat exchanger (a) Between flue gas and S4 (b) Between flue gas and S14.

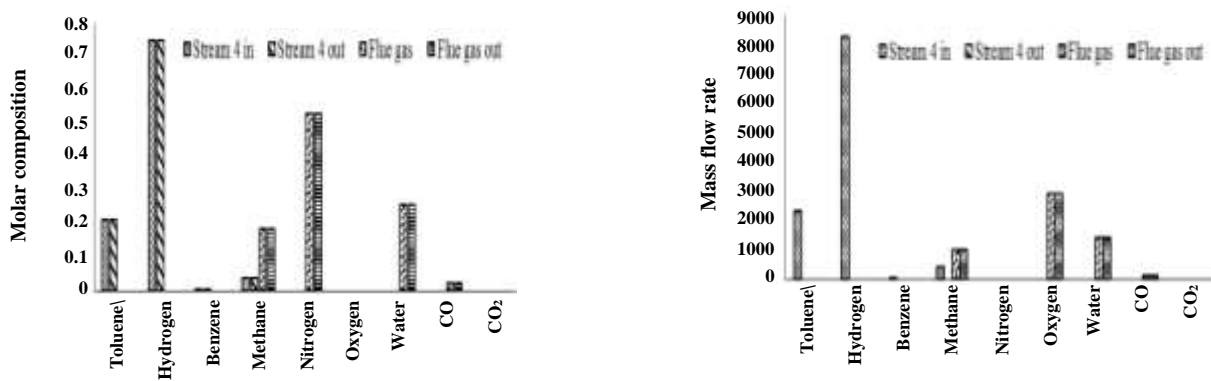


Fig. 22: Heat exchanger between S4 and flue gas (a) Mole fractions (b) Mass flow rate.

flue gas heat to n-butane in the heat exchanger. All the other equipment except the heat exchanger contains n-butane, so component-wise mole fractions and mass flow rates are not mentioned as they are all equal to zero.

That is why only the mass balance on the heat exchanger is in Fig. 20 (a, b).

$$\text{The mass flow rate in } \square \text{ Mass flow rate out}$$

$$m(\text{flue gas}) + m(\text{n-butane in}) \square m(\text{flue gas out}) + m(\text{n-butane out})$$

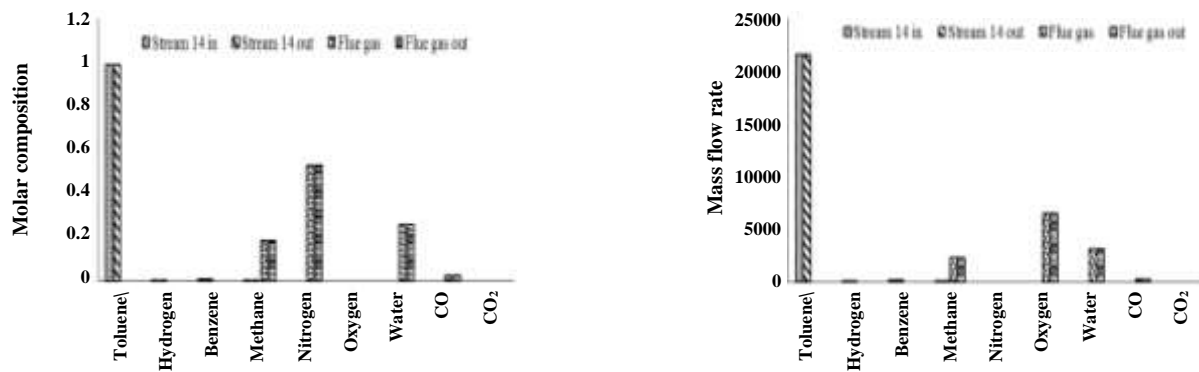


Fig. 23: Heat exchanger between S14 and flue gas (a) Mole fractions (b) Mass flow rate.

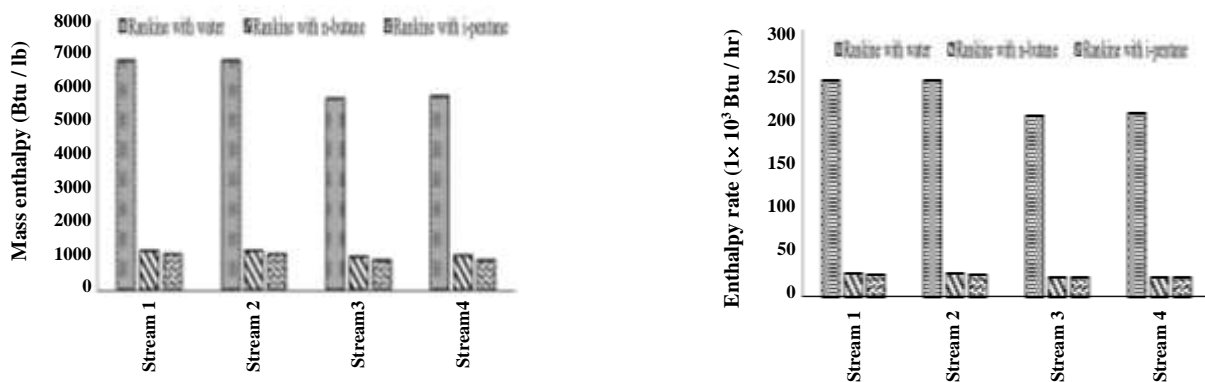


Fig. 24: Energy balance on Rankine with water, n-butane, and i-pentane in the benzene production unit (a) Mass Enthalpy (b) Enthalpy rate.

$22078.3684 \text{ lb/h} + 12168.7785 \text{ lb/h} \square 22078.3684 \text{ lb/hr} + 12168.7785 \text{ lb/h.}$

$34247.147 \text{ lb/h} = 34247.147 \text{ lb/h.}$

Heat exchanger between S4 and flue gas; heat exchanger is placed to preheat the stream named S4, the material balance shown in Fig. 21 (a) and Fig. 22 (a, b).

Heat exchanger between S14 and cooled flue gas; as heat from flue gas which is the outlet of 1<sup>st</sup> heat exchanger can further be recovered, so heat exchanger is placed to preheat the stream named S14 the material balance of which is shown in Fig. 21 (b) and Fig. 23 (a, b).

The mass flow rate in  $\square$  Mass flow rate out

$$m(\text{flue gas}) + m(\text{S14}) \square m(\text{flue gas out}) + m(\text{S14 out})$$

$12167.1615 \text{ lb/hr} + 21724.3533 \text{ lb/h} = 12167.1615 \text{ lb/h} + 21724.3533 \text{ lb/h.}$

$33891.5 \text{ lb/h} = 33891.5 \text{ lb/h.}$

The benzene production unit of energy balances Rankine with water, as shown in Fig. 24 (a, b).

Rankine with water boiler duty;  $Q_{in} = H_2 - H_3 = 24.426 \times 10^7 - 20.416 \times 10^7 = 4.01 \times 10^6 \text{ Btu/h.}$

Condenser duty;  $Q_{out} = H_1 - H_4 = 24.427 \times 10^7 - 20.627 \times 10^7 = 3.8 \times 10^6 \text{ Btu/h.}$

Efficiency;  $\eta = 1 - Q_{out}/Q_{in} = 0.05 = 5.2\%$

Rankine with i-pentane boiler duty;  $Q_{in} = H_2 - H_3 = 25.308 \times 10^6 - 21.328 \times 10^6 = 3.98 \times 10^6 \text{ Btu/h.}$

Condenser duty;  $Q_{out} = H_1 - H_4 = 25.312 \times 10^6 - 21.537 \times 10^6 = 3.774 \times 10^6 \text{ Btu/h.}$

Efficiency;  $\eta = 1 - Q_{out}/Q_{in} = 0.0517 = 5.17\%$

Rankine with n-butane boiler duty;  $Q_{in} = H_2 - H_3 = 26.242 \times 10^6 - 22.230 \times 10^6 = 4.01 \times 10^6 \text{ Btu/h.}$

Condenser duty;  $Q_{out} = H_1 - H_4 = 24.426 \times 10^6 - 22.567 \times 10^6 = 3.699 \times 10^6 \text{ Btu/h.}$

Efficiency;  $\eta = 1 - Q_{out}/Q_{in} = 0.079 = 7.9\%$

In conclusion, Rankine with water turbine produces 83.14hp or  $2.116 \times 10^5 \text{ Btu/h}$ , and boiler duty is

**Table 14: Case 01 of the amine treatment unit.**

Streams	Inlet Temperature of Flue Gas °F	Outlet Temperature of Flue Gas °F	Energy Saved BTU/h	Energy Saved kW
Acid Gas	180.6	172.4	40*10 <sup>3</sup>	11.72

**Table 15: Case 02 of the amine treatment unit.**

Streams	Inlet Temperature of Flue Gas °F	Outlet Temperature of Flue Gas °F	Heat Saved in Reboiler BTU/h	Heat Saved in Condenser BTU/h	Total Heat Saved BTU/h	Total Heat Saved kW
Acid Gas	180.5	161.4	5.35*10 <sup>6</sup>	5.31*10 <sup>6</sup>	10.66*10 <sup>6</sup>	3123.97

**Table 16: Cases of the amine treatment unit**

Streams	Inlet Temperature of Flue Gas °F	Outlet Temperature of Flue Gas °F	Work Produced by Turbine kW	Work Produced by Turbine KWh/yr
	Case 03 of the amine treatment unit			
Acid Gas	180.4	174	14.07	117,174.96
	Case 04 of the amine treatment unit			
	180.4	169.9	18.87	14.07

**Table 17: Comparative study of the amine treatment unit.**

Cases	Temperature In °F	Temperature Out °F	Total Energy Saved kW
Case 01	180.6	172.4	11.72
Case 02	180.5	161.4	3123.97
Case 03	180.4	174	-
Case 04	180.4	169.9	5.568

4.01\*10<sup>6</sup> Btu/h. Rankine with i-pentane turbine produces 82.33hp or 2.095\*10<sup>5</sup> Btu/h, and boiler duty is 3.98\*10<sup>6</sup> Btu/h. Rankine with n-butane turbine produces the power of 131.2hp and boiler duty is 4.01\*10<sup>6</sup> Btu/h [13, 14, 16, 18].

Heat exchanger between S4 and flue gas; Q (S4) = Q (flue gas)  
 $mC_p \Delta T = mC_p \Delta T$

$$1.004*10^6 \text{ Btu/h} = 1.004*10^6 \text{ Btu/h.}$$

Heat exchanger between S4 and flue gas and S14 and outlet of exchanger 1; Q (S4) = Q (flue gas)

$$mC_p \Delta T = mC_p \Delta T$$

$$3.98*10^6 \text{ Btu/h} = 3.98*10^6 \text{ Btu/h.}$$

Exchanger 02; Q (catalytic converter feed) = Q (flue gas)

$$mC_p \Delta T = mC_p \Delta T$$

$$(2.172*10^4 \text{ lb/h}) * (0.4109 \text{ Btu/lb}^\circ\text{F}) * (143.9 - 102.8)^\circ\text{F} =$$

$$(1.217*10^4 \text{ lb/h}) * (0.33495 \text{ Btu/lb}^\circ\text{F}) * (248 - 158)^\circ\text{F}$$

$$2.821*10^4 \text{ Btu/h} = 2.821*10^4 \text{ Btu/h.}$$

Total energy would be;

$$4.01*10^6 \text{ /h} = 4.01*10^6 \text{ Btu/h.}$$

### Comparative study of different alternatives

#### Amine treatment unit

Table 14 shows case 01 of the heat exchanger between acid gas and rich amine.

Table 15 shows case 02 of a heat exchanger between rich amine (outlet of exchanger 1) and lean amine.

Table 16 shows case 03 of the Rankine cycle and case 04 of the modified Rankine cycle.

Table 17 shows the comparative results from cases.

#### Sulfur removal or desulphurization

Table 18 shows case 01 of Rankine with water, case 02 of Rankine with n-butane, and case 03 of Rankine with i-pentane.

Table 19 shows case 04 of sulfur in heat exchangers.

Table 20 shows case 05 of Rankine with water in series with heat exchangers.

Table 21 shows comparative results from cases.

**Table 18: Cases of sulfur removal or desulphurization.**

Streams	Inlet Temperature °F	Outlet Temperature °F	Work Produced by Turbine hp	Work Produced by Turbine kW
	Case 01 of sulfur removal or desulphurization			
	778.7	740.4	31.48	23.47
	Case 02 of sulfur removal or desulphurization			
Flue gas	778.7	761	14.62	10.906
	Case 03 of sulfur removal or desulphurization			
	778.7	762.1	11.75	8.7655

**Table 19: Case 19 of sulfur removal or desulphurization.**

Streams	Inlet Temperature °F	Outlet Temperature °F	Heat Saved in Heater BTU/h	Heat Saved in Second Heater BTU/h	Total Heat Saved BTU/h	Total Heat Saved kW
Flue gas	778.7	361.3	1.196*10 <sup>6</sup>	1.816*10 <sup>6</sup>	3.012*10 <sup>6</sup>	882.6

**Table 20: Case 05 of sulfur removal or desulphurization.**

Streams	Inlet Temperature °F	Outlet Temperature °F	Work Produced by Turbine Hp	Heat Saved in Heater BTU/h	Heat Saved in Second Heater BTU/h	Total Heat Saved + Work produced By Turbine BTU/h	Total Heat Saved + Work produced by Turbine kW
Flue gas	778.7	345.5	31.48	1.196*10 <sup>6</sup>	1.816*10 <sup>6</sup>	3.09*10 <sup>6</sup>	905.54

**Table 21: Comparative study of sulfur removal or desulphurization.**

Cases	Temperature In °F	Temperature Out °F	Total Energy Saved (kW)
Case 01	778.7	740.4	82.939
Case 02	778.7	761	38.421
Case 03	778.7	762.1	36.077
Case 04	778.7	361.3	882.6
Case 05	778.7	345.5	965.59

### Benzene production unit

Table 22 shows case 01 of Rankine with water, case 02 of Rankine with i-butane, and case 03 of Rankine with i-pentane.

Table 23 shows case 04 of single heat exchangers.

Table 24 shows case 05 of two heat exchangers.

Table 25 shows comparative results from cases.

### Simulation and convergence process terminologies

#### Gas Sweetening

Gas sweetening is a method that must be performed to eliminate hydrogen-sulfide that is H<sub>2</sub>S from gasses. Gas sweetening is at times stated as amine treating. Amine treating is usually employed in petrochemical plants, NG treating plants, refineries, and other industries. In oil

refineries or chemical plants, it is usually described as gas sweetening. It is so because they involve a product in which the sour and rotten smell of hydrogen sulfide is not present. As this treatment process does not recover the sulfur in its elemental form, it must be followed by a sulfur recovery unit [19, 26].

#### Regenerator column

This column is a stripper column. Rich amine solution is entered from the top of the column, and the lean amine is sent to the reboiler, which uses the steam to strip CO<sub>2</sub> and H<sub>2</sub>S.

#### Combustor

Combustors transform the fuel's chemical energy, which is denoted by its heating value, off, into thermal

**Table 22: Cases of benzene production.**

Streams	Inlet Temperature °F	Outlet Temperature °F	Work Produced by Turbine hp	Work Produced by Turbine KW
Case 01 of benzene production				
Flue gas	1039.4	150.3	83.14	62
	Case 02 of benzene production			
	1039.4	150.3	131.2	97.82
	Case 03 of benzene production			
	1039.4	158	82.33	61.39

**Table 23: Case 04 of benzene production.**

Streams	Inlet Temperature °F	Outlet Temperature °F	Heat Saved in Heater BTU/h	Heat Saved KW
Flue gas	1039.4	158	3.98*10 <sup>6</sup>	1166.4

**Table 24: Case 05 of benzene production.**

Streams	Inlet Temperature °F	Outlet Temperature °F	Heat Saved in Heater BTU/h	Heat Saved in Second Heater BTU/h	Total Heat Saved BTU/h	Total Heat Saved KW
Flue gas	1039.4	151	3.98*10 <sup>6</sup>	2.82*10 <sup>4</sup>	4.008*10 <sup>6</sup>	1174.6

**Table 25: Comparative study of benzene production.**

Cases	Temperature In °F	Temperature Out °F	Total Energy Saved (KW)
Case 01	1039.4	150.3	1175.2
Case 02	1039.4	150.3	1175.2
Case 03	1039.4	158	1166.4
Case 04	1039.4	158	1166.4
Case 05	1039.4	151	1174.6

energy, which is symbolized by the stagnation of temperature  $T_t$ .

#### Stream Cutter

In Hysys, the fluid properties are calculated based on the defined property packages example Peng Robinson, acid gas solvent. There are seldom occurrences where the fluid is characterized as one property package. Still, there is equipment specified to take a particular property package as they cannot take stream inputs of another fluid package. Stream cutter is thus used in those before that equipment for altering the stream's fluid package for property estimation. For example, if the fluid properties are estimates based on chemical solvent and there is a separator that is defined by Peng Robinson so the steam cutter should be used before that separator to convert the

#### Convergence criteria

One of the essential rules for using simulation software is to keep in mind that if the input data is correct, it will lead to correct results based on the software's specifications.

Similarly, one cannot rely on the simulation results if the input is incorrect or unrealistic concerning practical application or if tolerance is increased to an unacceptable level. Due to these reasons, it is important to know the basic criteria of convergence of equipment used and have a basic check on input data to ensure that the results of simulations are reliable.

#### Separator (V-100, V-101, V-102, V-103, V104, V-105)

Initially, connect streams to the separator and define the inlet and outlet. The temperature and flow rate of respective streams are defined after that. If joining the separator in a process line, then there is no need to define the temperature and flow rate because they are predefined as the streams are connected. The pressure and temperature of the inlet and outlet streams remain the same. However, the flow rate is an inlet in the sum of two flow rates, leaving the column. For convergence, the active specs include pressure specifications and flow specifications, as shown in Fig. 25 (a, b, c, d, e, f).

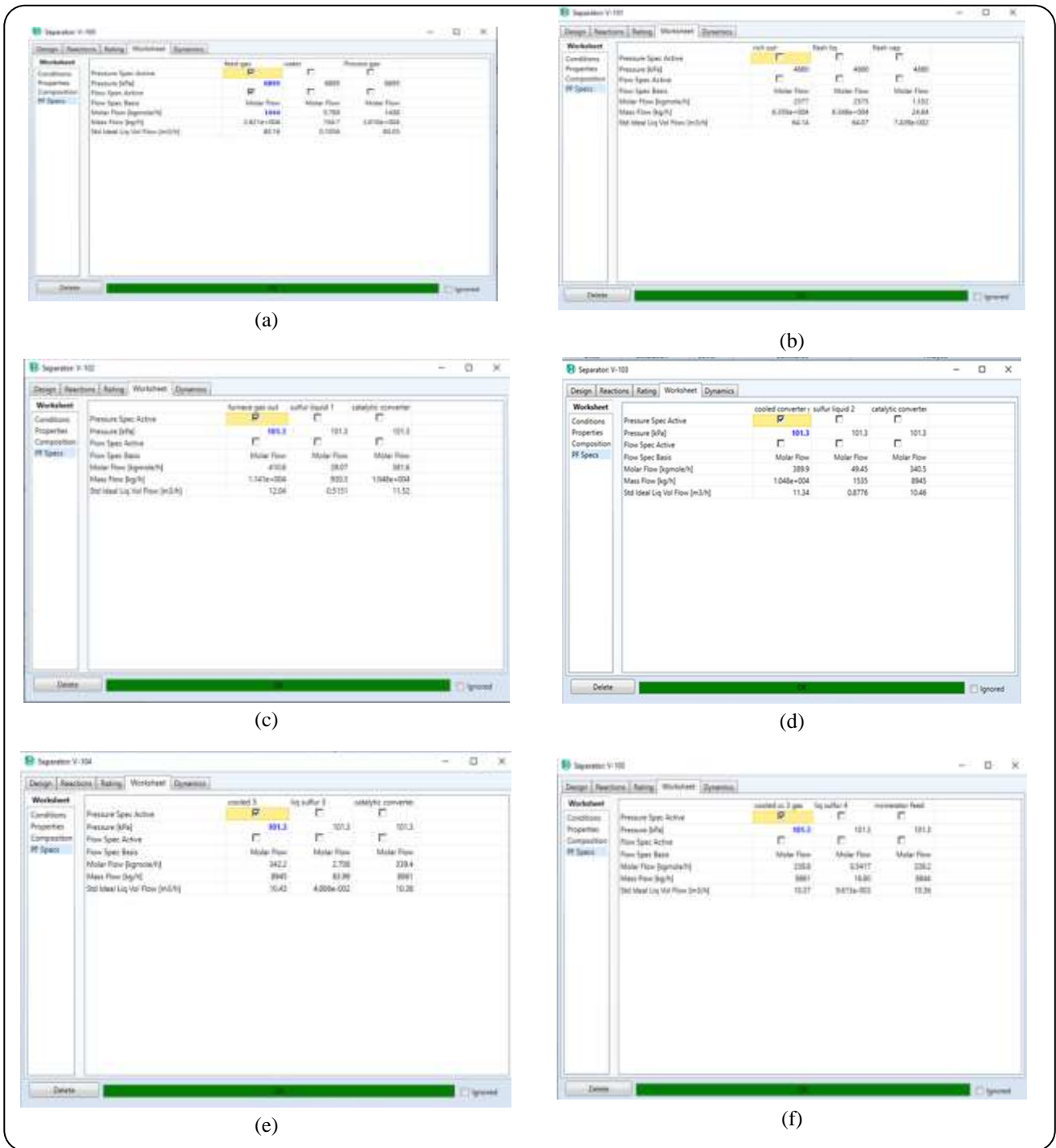


Fig. 25: PF specification of separator (a) V-100 (b) V-101 (c) V-102 (d) V-103 (e) V-104 (f) V-105

**Absorber**

The first step is to connect the inlet and outlet streams. Define pressures for the first and last stages; define the temperatures and flow rate.

During the process, there is no need to define temperature and flow rate if the absorber is connected

in the streamline. Absorbers always have one stream connected, which is from the regenerator. Several equations are used in the column including the Francis Weir equation, which calculates the liquid flow rate leaving the tray. The monitor tab in the rating shows the tolerance and step size of the column, as given in Fig. 26.

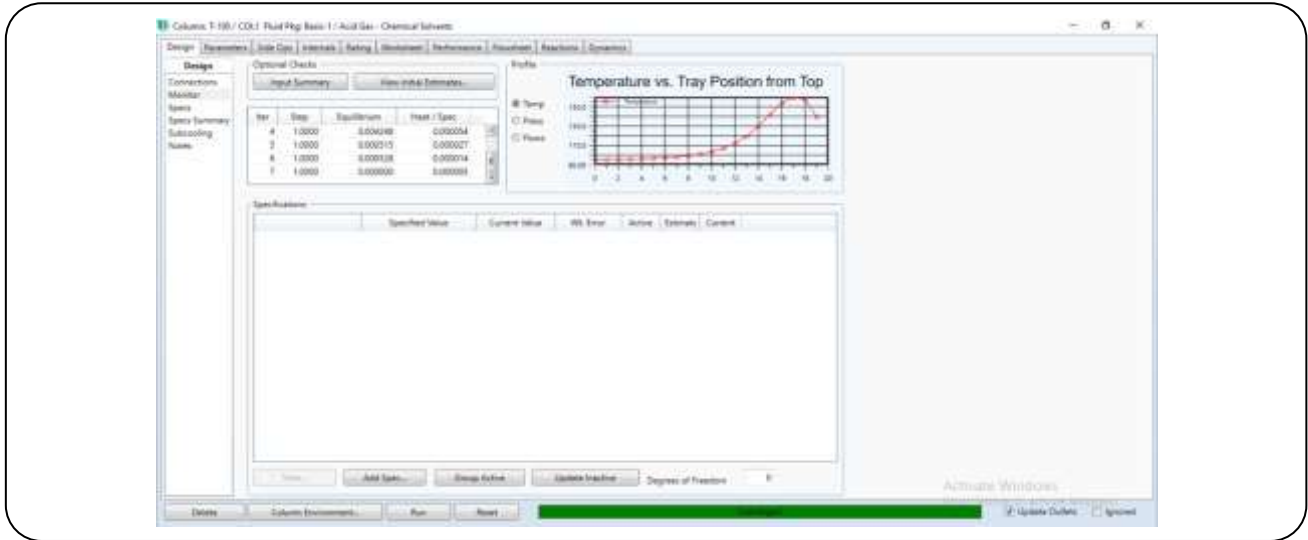


Fig. 26: Monitor of column T-100.

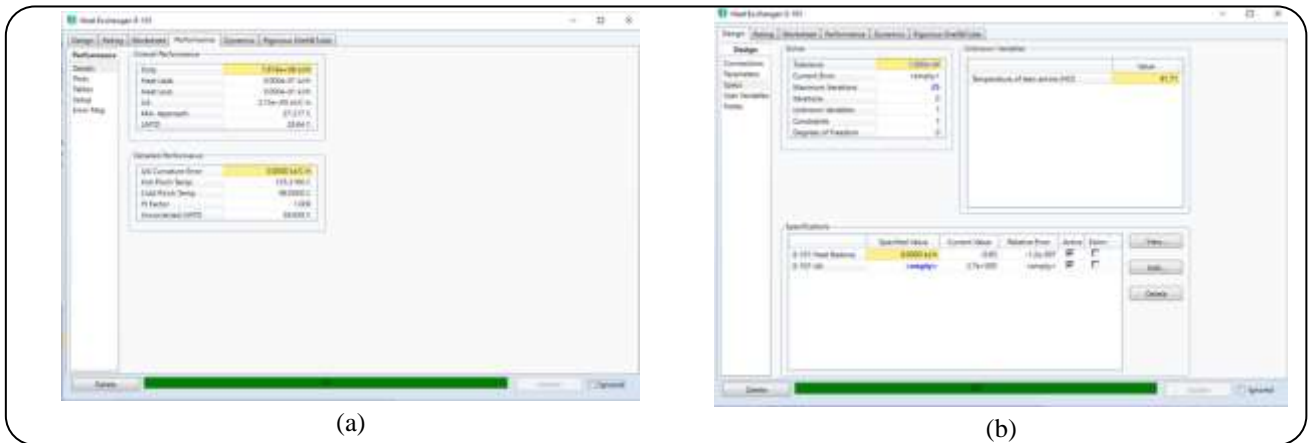


Fig. 27: (a) Details of heat exchanger E-101 (b) Specs of heat exchanger E-101.

**Heat exchanger**

The shell and tube side pressure drops must be specified to specify the pressure drop for heat exchangers. Heaters and coolers are heat exchangers that use the same basic equation but with different sign conventions. In a cooler, the energy stream's heat is subtracted from the inlet, whereas in a heater, it is added to the inlet stream. Heat Exchangers in Aspen HYSYS are modeled by the following equation [2, 11, 15, 17].

$$Q = UA (T_f - T_{amb}) \tag{11}$$

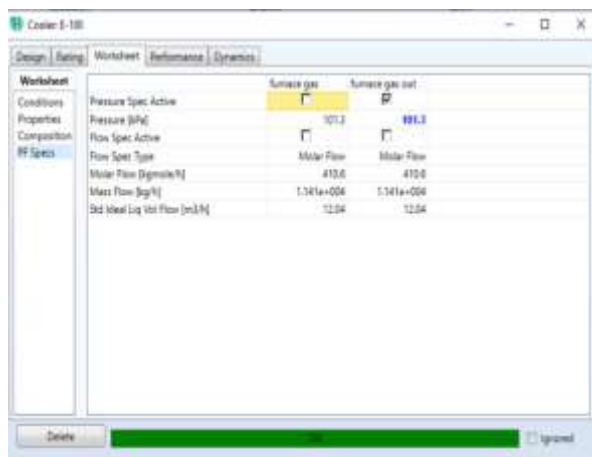
If the temperatures of both inlet streams are defined and only define the temperature of only one outlet stream, then HYSYS calculates the outlet temperature of the other stream by itself. But for this, we have to enter the pressure drop for the tube side and the shell side.

**Heater/Cooler**

To converge the heater and cooler, define the temperature to want the fluid is heated or cooled. And duty is evaluated. The working principle or heater/cooler is the same as the heat exchanger and the same equations are used for modeling these two pieces of equipment. The function of both heater and the cooler is the same, the only difference is in the sign convention. The heater takes the energy, and in the cooler, the energy is extracted. Apart from that, energy balance and enthalpy change are the same for both of the equipment. If the pressure drop is not zero, it is always desired to have a pressure drop of less than 10 bar if the pressure drop exists, as shown in Figs. 27 to 29.

**Regenerator column**

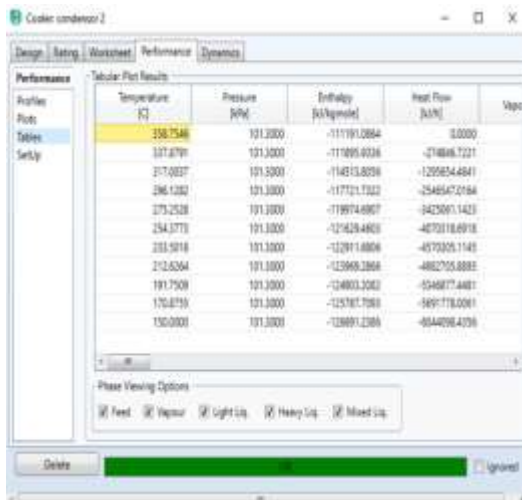
The sub-flow sheet of the distillation column is different from the parent's environment. But to define



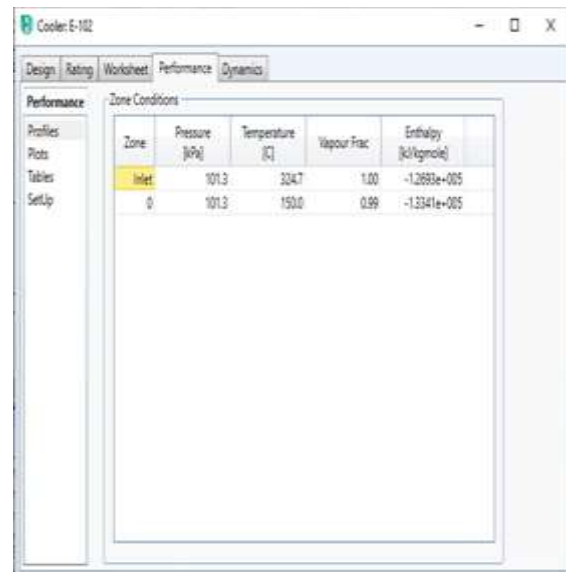
(a)



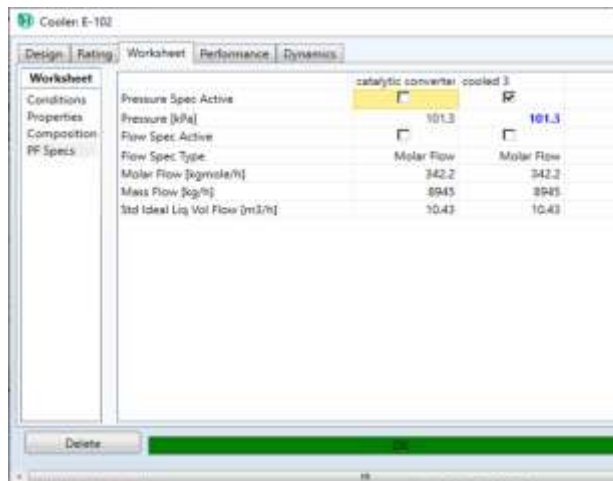
(b)



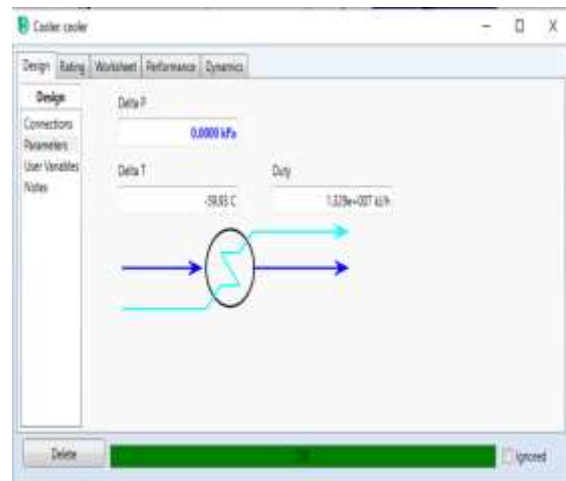
(c)



(d)



(e)



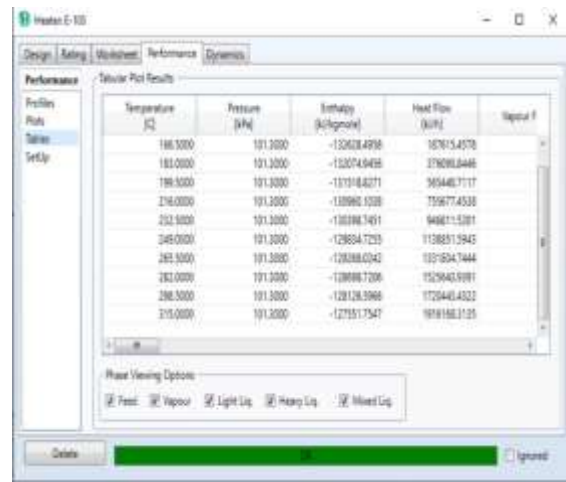
(f)

Fig. 28: (a) PF Specs of cooler E-100 (b) PF Specs cooler condenser 2 (c) Tables of cooler condenser 2 (d) Profiles of cooler E-102 (e) PF Specs of cooler E-102 (f) Parameters of cooler.





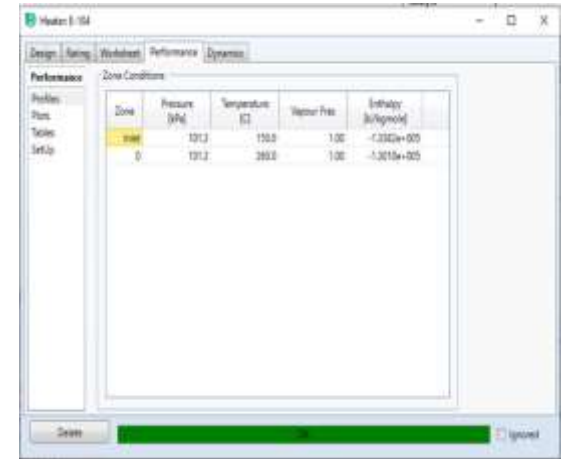
(a)



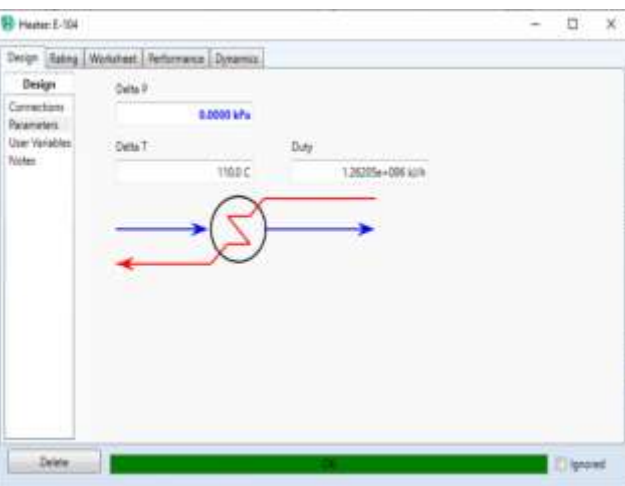
(b)



(c)



(d)



(e)



(f)

Fig. 29: (a) PF Specs of heater E-103 (b) Tables of heater E-103 (c) PF Specs of heater E-104 (d) Profiles of heater E-104 (e) Parameters of heater E-104 (f) Specs of heater E-101.

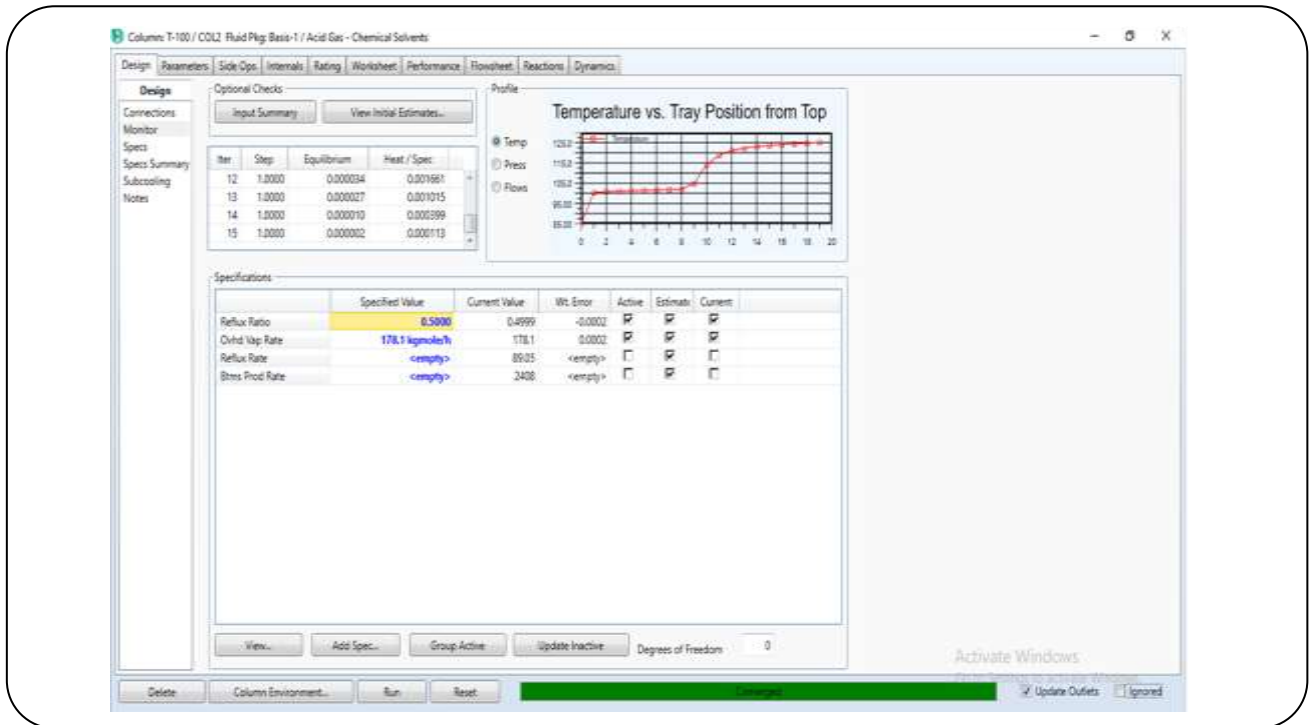


Fig. 30: Monitor of column T-100.

the specifications in the parent environment, the sub-flow sheet has to be fully defined. To converge the column to define a bulk of parameters and for a column, a different fluid package can also be used. Even sometimes, apply a different fluid package only for the distillation column to have efficient results. In other equipment, the interface shows less information that includes only inlet-outlet temperatures and some of their specifications. The distillation column interface shows various parameters at which stage enters the inlet and from where reflux enters again. Also, add additional equipment like a pump around or the strippers through the same Interface. It can conclude that the property view of the column is a mini-environment that performs various functions.

The Francis weir equation is the initial step for calculating the flow rate of the liquid or fluid that is leaving the tray. And the vapor flow rate leaving the tray is determined by the residence equation. For the column, the conductance is proportional to the square of the column's diameter. And the pressure drop across the column is determined by summing the static head and frictional losses. To make the degree of freedom zero, define some specifications, and then HYSYS estimates other parameters. And once the column is converged, it generates various profiles concerning temperature, pressure, and flow rate.

It tells information about the amount of iteration which was done to obtain the convergence, as shown in Fig. 30. Tolerance is also defined in this interface [2, 17].

#### Conversion Reactor (Catalytic Converter Number 1, Catalytic Converter Number 2, Catalytic Converter Number 3, CR-103)

As the name defines, it is a reactor used for conversion reactions. If somehow, add different reactions like kinetic or equilibrium reactions to it, it will not execute the functions. Initially, the reaction set defined the reactor before entering any stream; after that, the reaction set is added to the reactor. Then, temperature, pressure, and flowrates are entered into the flowsheet, and eventually, the outlet stream is automatically defined as given in Fig. 31 (a, b, c, d, e, f, g, h).

#### Pump

The pump should know its duty of the pump. After that, estimate to which extent increases the pressure. But here, got pump converged by defining the outlet temperature because of the conditions of the inlet stream. Iterations, pump evaluated have much temperature at the outlet, how much duty is required as represented in Figs. 32 (a, b) and set the efficiency of 0.75.

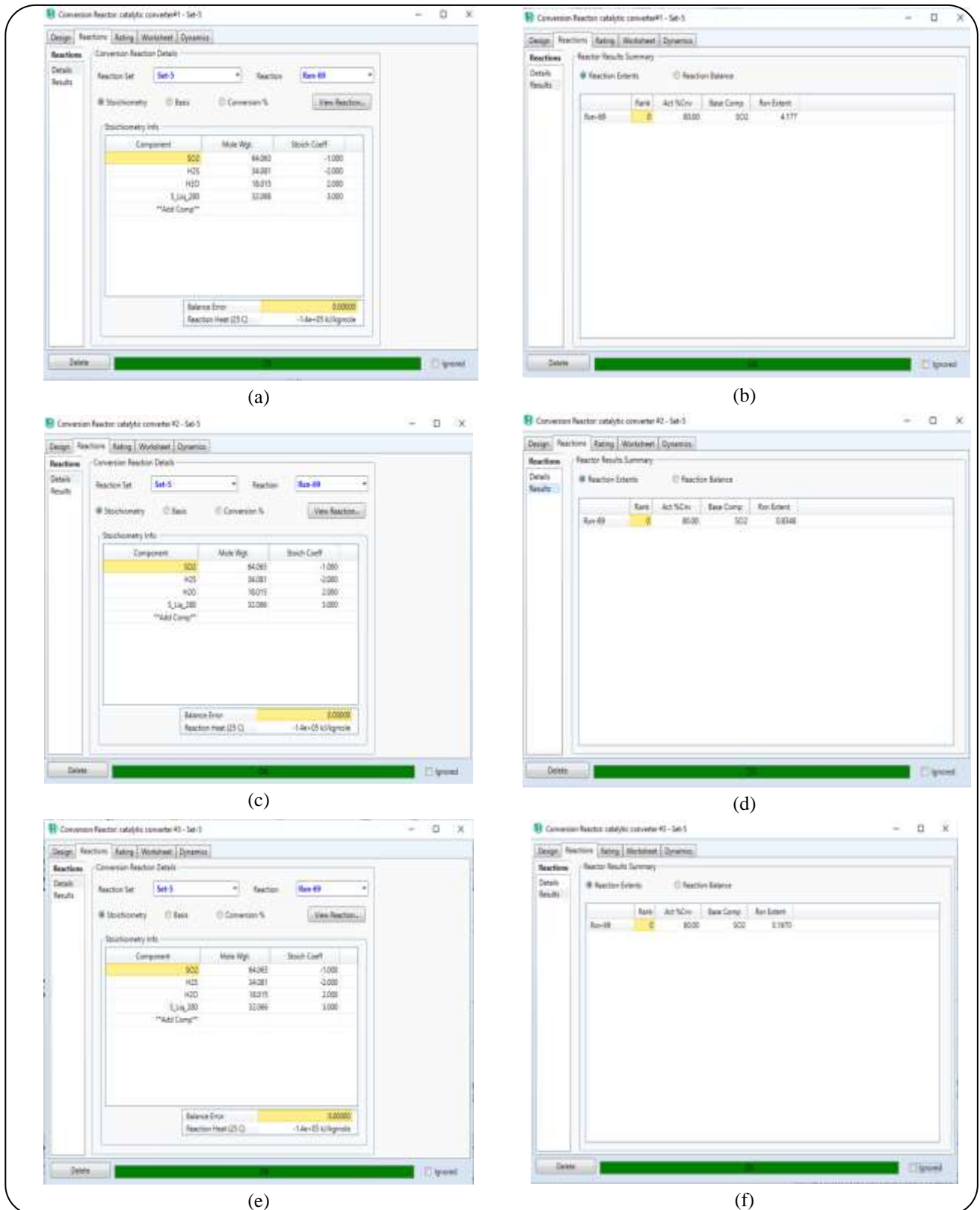


Fig. 31: (a) Details of catalytic converter number1 (b) Results of catalytic converter number1 (c) Details of catalytic converter number 2 (d) Results of catalytic converter number 2 (e) Details of catalytic converter number 3 (f) Results of catalytic converter number 3 (g) Details of conversion reactor CRV-103 (h) Results of conversion reactor CRV-103.

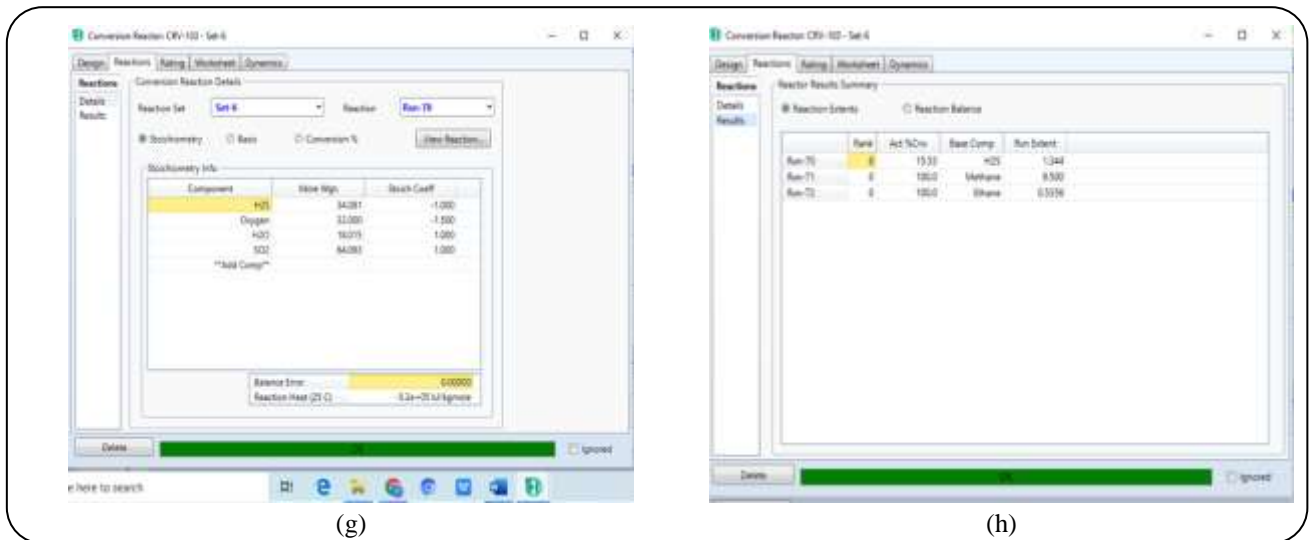


Fig. 31: (a) Details of catalytic converter number 1 (b) Results of catalytic converter number 1 (c) Details of catalytic converter number 2 (d) Results of catalytic converter number 2 (e) Details of catalytic converter number 3 (f) Results of catalytic converter number 3 (g) Details of conversion reactor CRV-103 (h) Results of conversion reactor CRV-103. Continuation

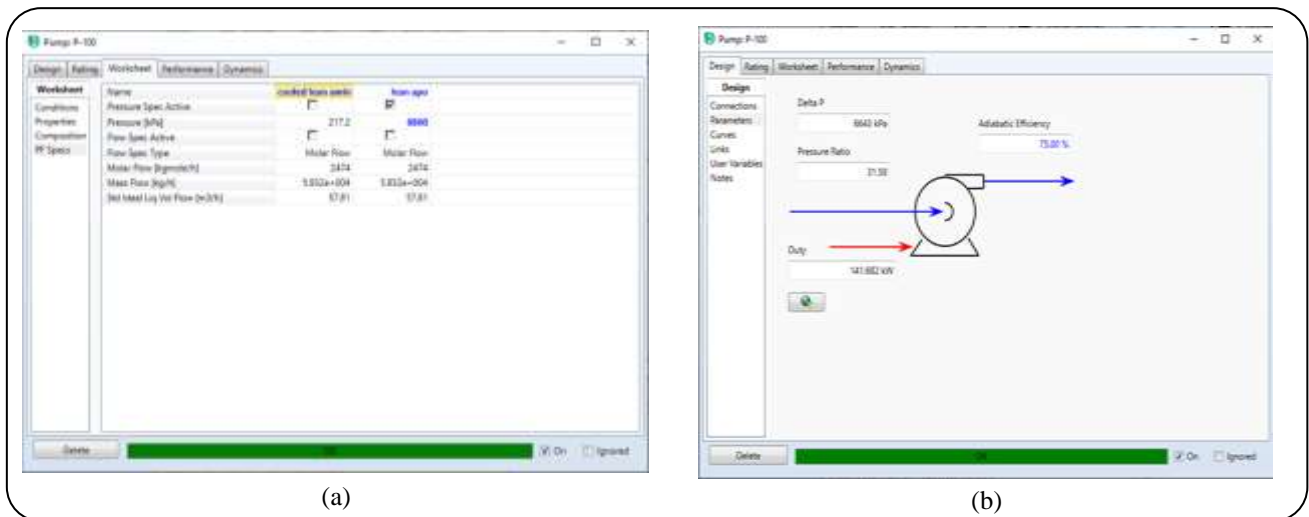


Fig. 32: (a) PF Specs of pump P-100 (b) Parameters of pump P-100.

**Cost optimization case study:**

**Amin treatment unit**

*Case 1: Single Heat Exchanger*

Cost saved in reboiler =  $40 \times 10^3$  BTU/h =  $(40 \times 10^3 \text{ BTU/h}) (1055 \text{ J/BTU}) (1 \text{ h}/3600 \text{ s}) = 11.72 \text{ kW}$   
 =  $(11.72 \text{ kW}) (8328 \text{ h/yr}) (1 \text{ unit}/1 \text{ kWh})$   
 =  $(9.406 \text{ PKR/unit}) = 918064 \text{ PKR/yr.}$   
 =  $(918064 \text{ PKR/yr}) (1 \$/168.5 \text{ PKR}) = \$5448.45/\text{yr.}$

*Case 2: Double Heat Exchanger*

Cost saved in reboiler (Qr) =  $(5.35 \times 10^6 \text{ BTU/h}) (1055 \text{ J/BTU}) (1 \text{ h}/3600 \text{ s}) = 1.567 \times 10^6 \text{ W}$

=  $(1567 \times 10^3 \text{ kW}) (8328 \text{ h/yr}) (1 \text{ unit}/1 \text{ kWh})$   
 =  $(9.06 \text{ PKR/unit})$   
 =  $(118.296 \times 10^6 \text{ PKR/yr}) (1 \$/168.5 \text{ PKR}) =$   
 =  $\$701678.23/\text{yr.}$   
 Cost Saved In Condenser (Qc) =  $(5.31 \times 10^6 \text{ BTU/h}) (1055 \text{ J/BTU}) (1 \text{ h}/3600 \text{ s})$   
 =  $(1556.1 \text{ kW}) (8328 \text{ h/yr}) (1 \text{ unit}/1 \text{ kWh})$   
 =  $(9.06 \text{ PKR/unit})$   
 =  $(117.41 \times 10^6 \text{ PKR/yr}) (1 \$/168.5 \text{ PKR}) =$   
 =  $\$696797.38/\text{yr.}$   
 Total Cost Saved per year =  $\$701678.23/\text{yr} +$   
 =  $\$696797.38/\text{yr} = \$1398475.6/\text{yr.}$

**Case 3: Rankine cycle**

The cost saved in a boiler, if use electricity to operate the boiler, then cost saved is given by:

$$\begin{aligned} \text{Cost saved per year} &= (1.729 \times 10^5 \text{ BTU/h}) \\ &(1055 \text{ J/1 BTU}) (1 \text{ h}/3600 \text{ s}) = 50.66 \text{ kW} \\ &= (50.66 \text{ kW}) (8328 \text{ h/yr}) (1 \text{ unit}/1 \text{ kWh}) \\ &(9.06 \text{ PKR}/1 \text{ unit}) (1 \$/168.5 \text{ PKR}) = \$22684.76/\text{yr}. \end{aligned}$$

The above is the estimated cost saved by using electricity, now the cost which can be saved if coal is used as a fuel is calculated as Fuel = Anthracite Coal, Calorific Value = 24 MJ/kg, Energy Obtained = 182.40995 MJ/h.

$$\text{Mass flowrate} = (182.40995 \text{ MJ/h}) (24 \text{ MJ/kg}) = 7.296 \text{ kg/h}.$$

$$\text{Cost of fuel per kg} = \$0.03338/\text{kg}$$

$$\text{The cost saved per year} = (7.296 \text{ kg/h}) (\$0.03338/\text{kg}) (8328 \text{ h/yr}) = \$2028.20/\text{yr}.$$

**Case 4: Modified Rankine Cycle**

The electricity used to operate the boiler, then the cost saved is given by:

$$\begin{aligned} \text{Cost saved per year} &= (1.542 \times 10^5 \text{ BTU/h}) \\ &(1055 \text{ J/1 BTU}) (1 \text{ h}/3600 \text{ s}) = 45.2 \text{ kW} \\ &= (45.2 \text{ kW}) (8328 \text{ h/yr}) (1 \text{ unit}/1 \text{ kWh}) (9.06 \text{ PKR}/1 \text{ unit}) \\ &(1 \$/168.5 \text{ PKR}) = \$20239.85/\text{yr}. \end{aligned}$$

The above is the estimated cost saved by using electricity, now the cost which can be saved if coal is used as a fuel is calculated as; Fuel = Anthracite coal, Calorific value = 24 MJ/Kg, Energy obtained = 162.681 MJ/h.

$$\text{Mass flowrate} = (162.681 \text{ MJ/h}) (24 \text{ MJ/kg}) = 6.78 \text{ kg/h}.$$

$$\text{Cost of fuel per kg} = \$0.03338/\text{kg}$$

$$\text{Cost saved per year} = (6.78 \text{ kg/h}) (\$0.03338/\text{kg}) (8328 \text{ h/yr}) = \$1884.76/\text{yr}.$$

**Case study: Sulfur removal/desulfurization****Rankine with water**

The electricity used to operate the boiler then the cost saved is given by:

$$\begin{aligned} \text{Cost saved per year} &= (2.832 \times 10^5 \text{ BTU/h}) \\ &(1055 \text{ J/1 BTU}) (1 \text{ h}/3600 \text{ s}) = 82.99 \text{ kW} \\ &= (82.99 \text{ kW}) (8328 \text{ h/yr}) (1 \text{ unit}/1 \text{ kWh}) \\ &(9.06 \text{ PKR}/1 \text{ unit}) (1 \$/168.5 \text{ PKR}) = \$38580/\text{yr}. \end{aligned}$$

The above is the estimated cost saved by using electricity, now the cost which can be saved if coal is used as a fuel is calculated as; Fuel = Anthracite coal, Calorific value = 24 MJ/kg, Energy obtained = 298.776 MJ/h.

$$\text{Mass flowrate} = (298.776 \text{ MJ/h}) (24 \text{ MJ/kg}) = 12.45 \text{ kg/h}.$$

$$\text{Cost of fuel per kg} = \$0.03338/\text{kg}$$

$$\text{Cost saved per year} = (12.45 \text{ kg/h}) (\$0.03338/\text{kg}) (8328 \text{ h/yr}) = \$3460.95/\text{yr}.$$

**Rankine with n-Butane**

The electricity used to operate the boiler then the cost saved is given by:

$$\begin{aligned} \text{Cost saved per year} &= (1.311 \times 10^5 \text{ BTU/h}) \\ &(1055 \text{ J/1 BTU}) (1 \text{ h}/3600 \text{ s}) = 38.42 \text{ kW} \\ &= (38.42 \text{ kW}) (8328 \text{ h/yr}) (1 \text{ unit}/1 \text{ kWh}) \\ &(9.06 \text{ PKR}/1 \text{ unit}) (1 \$/168.5 \text{ PKR}) = \$17860.89/\text{yr}. \end{aligned}$$

The above is the estimated cost saved by using electricity, now the cost which can be saved if coal is used as a fuel is calculated as; Fuel = Anthracite coal, Calorific value = 24 MJ/kg, Energy obtained = 138.31 MJ/h.

$$\text{Mass flowrate} = (138.31 \text{ MJ/h}) (24 \text{ MJ/kg}) = 5.763 \text{ kg/h}.$$

$$\text{Cost of fuel per kg} = \$0.03338/\text{kg}$$

$$\text{Cost saved per year} = (5.763 \text{ kg/h}) (\$0.03338/\text{kg}) (8328 \text{ h/yr}) = \$1602.04/\text{yr}.$$

**Rankine with i-Pentane**

The electricity used to operate the boiler then the cost saved is given by:

$$\begin{aligned} \text{Cost saved per year} &= (1.232 \times 10^5 \text{ BTU/h}) \\ &(1055 \text{ J/1 BTU}) (1 \text{ h}/3600 \text{ s}) = 36.10 \text{ kW} \\ &= (36.10 \text{ kW}) (8328 \text{ h/yr}) (1 \text{ unit}/1 \text{ kWh}) \\ &(9.06 \text{ PKR}/1 \text{ unit}) (1 \$/168.5 \text{ PKR}) = \$16782.35/\text{yr}. \end{aligned}$$

The above is the estimated cost saved by using electricity, now the cost which can be saved if coal is used as a fuel is calculated as; Fuel = Anthracite coal, Calorific value = 24 MJ/Kg, Energy obtained = 129.97 MJ/h.

$$\text{Mass flowrate} = (129.97 \text{ MJ/h}) (24 \text{ MJ/kg}) = 5.415 \text{ kg/h}.$$

$$\text{Cost of fuel per kg} = \$0.03338/\text{kg}$$

$$\text{Cost saved per year} = (5.415 \text{ kg/h}) (\$0.03338/\text{kg}) (8328 \text{ h/yr}) = \$1505.308/\text{yr}.$$

**Sulfur in heat exchangers (1\$/168.5PKR)**

$$\begin{aligned} \text{Cost saved in first heater} &= (1.196 \times 10^6 \text{ BTU/h}) \\ &(1055 \text{ J/1 BTU}) (1 \text{ h}/3600 \text{ s}) = 350.5 \text{ kW} \\ &= (350.5 \text{ kW}) (8328 \text{ h/yr}) (1 \text{ unit}/1 \text{ kWh}) \\ &(9.06 \text{ PKR}/1 \text{ unit}) = \$162942.28/\text{yr}. \end{aligned}$$

$$\begin{aligned} \text{Cost saved in second heater} &= (1.816 \times 10^6 \text{ BTU/h}) \\ &(1055 \text{ J/1 BTU}) (1 \text{ h}/3600 \text{ s}) = 532.18 \text{ kW} \\ &= (532.18 \text{ kW}) (8328 \text{ h/yr}) (1 \text{ unit}/1 \text{ kWh}) \\ &(9.06 \text{ PKR}/1 \text{ unit}) = \$247402/\text{yr}. \end{aligned}$$

Total cost saved per year = \$162942.28/yr + \$247402/yr = \$410344.28/yr.

#### Rankine with water in series with heat exchangers

The electricity used to operate the boiler then the cost saved is given by:

$$\begin{aligned} \text{Cost saved per year} &= (2.832 \cdot 10^5 \text{ BTU/h}) \\ &(1055 \text{ J/1 BTU}) (1 \text{ h}/3600 \text{ s}) = 82.99 \text{ kW} \\ &= (82.99 \text{ kW}) (8328 \text{ h/yr}) (1 \text{ unit}/1 \text{ kWh}) \\ &(9.06 \text{ PKR}/1 \text{ unit}) (1 \$/168.5 \text{ PKR}) = \$38580/\text{yr}. \end{aligned}$$

The above is the estimated cost saved by using electricity, now the cost which can be saved if coal is used as a fuel is calculated as; Fuel = Anthracite coal, Calorific value = 24 MJ/kg, Energy obtained = 298.776 MJ/h.

$$\text{Mass flowrate} = (298.776 \text{ MJ/h}) (24 \text{ MJ/kg}) = 12.45 \text{ kg/h}.$$

$$\text{Cost of fuel per kg} = \$0.03338/\text{kg}$$

$$\text{Cost saved per year} = (12.45 \text{ kg/h}) (\$0.03338/\text{kg}) (8328 \text{ h/yr}) = \$3460.95/\text{yr}.$$

Now, the costs by heat exchangers are given by:

$$\begin{aligned} \text{Cost saved in first heater} &= (1.196 \cdot 10^6 \text{ BTU/h}) \\ &(1055 \text{ J/1 BTU}) (1 \text{ h}/3600 \text{ s}) = 350.5 \text{ kW} \\ &= (350.5 \text{ kW}) (8328 \text{ h/yr}) (1 \text{ unit}/1 \text{ kWh}) \\ &(9.06 \text{ PKR}/1 \text{ unit}) (1 \$/168.5 \text{ PKR}) = \$162942.28/\text{yr}. \end{aligned}$$

$$\begin{aligned} \text{Cost saved in second heater} &= (1.816 \cdot 10^6 \text{ BTU/h}) \\ &(1055 \text{ J/1 BTU}) (1 \text{ h}/3600 \text{ s}) = 532.18 \text{ kW} \\ &= (532.18 \text{ kW}) (8328 \text{ h/yr}) (1 \text{ unit}/1 \text{ kWh}) \\ &(9.06 \text{ PKR}/1 \text{ unit}) = \$247402/\text{yr}. \end{aligned}$$

$$\text{Now, cost obtained by both heaters} = \$162942.28/\text{yr} + \$247402/\text{yr} = \$410344.28/\text{yr}.$$

For total cost, total cost saved per year when electricity is used = \$38580/yr + 410344.28/yr = \$448924.28/yr.

The total cost saved per year when coal is used = \$3460.95/yr + \$410344.28/yr = \$413805.23/yr.

#### Case Study: Benzene Cycle

##### Benzene with *i*-pentane

If using electricity to operate the boiler then the cost saved is given by:

$$\begin{aligned} \text{Cost saved per year} &= (3.98 \cdot 10^6 \text{ BTU/h}) (1055 \text{ J/1 BTU}) \\ &(1 \text{ h}/3600 \text{ s}) = 1166.36 \text{ kW} \\ &= (1166.36 \text{ kW}) (8328 \text{ h/yr}) (1 \text{ unit}/1 \text{ kWh}) \\ &(9.06 \text{ PKR}/1 \text{ unit}) (1 \$/168.5 \text{ PKR}) = \$542223.58/\text{yr}. \end{aligned}$$

The above is the estimated cost saved by using electricity, now the cost which can be saved if coal is used as a fuel is calculated as; Fuel = Anthracite coal, Calorific value = 24 MJ/kg, Energy obtained = 4198.9 MJ/h.

$$\text{Mass flowrate} = (4198.9 \text{ MJ/h}) (24 \text{ MJ/Kg}) = 174.95 \text{ kg/h}.$$

$$\text{Cost of fuel per kg} = \$0.03338/\text{kg}$$

$$\text{Cost saved per year} = (174.95 \text{ kg/h}) (\$0.03338/\text{kg}) (8328 \text{ h/yr}) = \$48634.11/\text{yr}.$$

##### Benzene with *n*-Butane

If using electricity to operate the boiler then the cost saved is given by:

$$\begin{aligned} \text{Cost saved per year} &= (4.012 \cdot 10^6 \text{ BTU/h}) \\ &(1055 \text{ J/1 BTU}) (1 \text{ h}/3600 \text{ s}) = 1175.74 \text{ kW} \\ &= (1175.74 \text{ kW}) (8328 \text{ h/yr}) (1 \text{ unit}/1 \text{ kWh}) \\ &(9.06 \text{ PKR}/1 \text{ unit}) (1 \$/168.5 \text{ PKR}) = \$546584.2/\text{yr}. \end{aligned}$$

The above is the estimated cost saved by using electricity, now the cost which can be saved if coal is used as a fuel is calculated as; Fuel = Anthracite coal, Calorific value = 24 MJ/Kg, Energy obtained = 4232.66 MJ/h.

$$\text{Mass flowrate} = (4232.66 \text{ MJ/h}) (24 \text{ MJ/kg}) = 176.36 \text{ kg/h}$$

$$\text{Cost of fuel per kg} = \$0.03338/\text{kg}$$

$$\text{Cost saved per year} = (176.36 \text{ kg/h}) (\$0.03338/\text{kg}) (8328 \text{ h/yr}) = \$49026.3/\text{yr}.$$

##### Benzene with Water

If using electricity to operate the boiler then the cost saved is given by:

$$\begin{aligned} \text{Cost saved per year} &= (4.011 \cdot 10^6 \text{ BTU/h}) \\ &(1055 \text{ J/1 BTU}) (1 \text{ h}/3600 \text{ s}) = 1175.44 \text{ kW} \\ &= (1175.44 \text{ kW}) (8328 \text{ h/yr}) (1 \text{ unit}/1 \text{ kWh}) \\ &(9.06 \text{ PKR}/1 \text{ unit}) (1 \$/168.5 \text{ PKR}) = \$546444.74/\text{yr}. \end{aligned}$$

The above is the estimated cost saved by using electricity, now the cost which can be saved if coal is used as a fuel is calculated as; Fuel = Anthracite coal, Calorific value = 24 MJ/Kg, Energy obtained = 4231.605 MJ/h.

$$\text{Mass flowrate} = (4231.605 \text{ MJ/h}) (24 \text{ MJ/Kg}) = 176.31 \text{ kg/h}.$$

$$\text{Cost of fuel per kg} = \$0.03338/\text{kg}$$

$$\text{The cost saved per year} = (176.31 \text{ kg/h}) (\$0.03338/\text{kg}) (8328 \text{ h/yr}) = \$49012.17/\text{yr}.$$

##### Single heat exchanger

$$\begin{aligned} \text{Cost saved in heater} &= (3.98 \cdot 10^6 \text{ BTU/h}) \\ &(1055 \text{ J/1 BTU}) (1 \text{ h}/3600 \text{ s}) = 1166.36 \text{ kW} \\ &= (1166.36 \text{ kW}) (8328 \text{ h/yr}) (1 \text{ unit}/1 \text{ kWh}) \\ &(9.06 \text{ PKR}/1 \text{ unit}) (1 \$/168.5 \text{ PKR}) = \$542223.58/\text{yr}. \end{aligned}$$

**Table 26: Energy and cost of amine treatment plant alternatives.**

Case	Alternative	Energy Saved (kW)	Power Generation (kW)	Boiler Operated with	Revenue (\$/yr)
1	Single heat exchanger	11.72	-	-	5448.45
2	Double heat exchanger	1568	-	-	139847
3	Rankine cycle	-	14.07	Electricity	22684.76
				Anthracite coal	2028.20
4	Modified Rankine cycle	5.568	14.07	Electricity	20239.85
				Anthracite coal	1884.76

### Two heat exchangers

Cost saved in first heater =  $(3.98 \times 10^6 \text{ BTU/h})$  (1055J/1BTU) (1h/3600s) = 1166.36KW

Cost saved per year = (1166.36KW) (8328h/yr) (1unit/1KWh) (9.06PKR/1unit) (1\$/168.5PKR) = \$542223.58/yr.

Cost saved in second heater =  $(2.82 \times 10^4 \text{ BTU/h})$  (1055J/1BTU) (1h/3600s) = 8.264kW

Cost saved per year = (8.264kW) (8328h/yr) (1unit/1KWh) (9.06PKR/1unit) (1\$/168.5PKR) = \$3839.95/yr.

Total cost saved per year = \$542223.58/yr + \$3839.95/yr = \$546063/yr.

### Amine Treatment Plant

These were some of the alternatives applied to the most targeted energy optimization area, as shown in Table 26. Here can observe the cost which will be saved after the application of these ideas. So, the application of double heat exchangers saves most of the energy, and it has the greatest revenue of all. Before this, some heaters were employed to heat different streams. Applied the approach of using the heat of any other stream to heat the desired one just by replacing the heater, and luckily this approach saved a great amount of energy and generated great revenue as well. Each alternative has proved to be fruitful as far as the economy and environmental safety are concerned. Rankine Application is of a different advantage as generating power by utilizing its heat [5, 17, 22, 24].

### Sulfur recovery unit

In this sulfur recovery unit, flue gas was the material that had so much potential for thermal waste recovery, as given in Table 27. So, used the thermal energy of flue gas in various ways to extract the maximum amount of

revenue out of it. Out of those alternatives, here too, the double heat exchanger proved to be the most energy-efficient approach. It cut the need for a heater which was consuming a great amount of energy to run. So, by cutting, they gathered a great amount of revenue. Other alternatives are also proved to be good choices in terms of saving the environment as well. Talking about power generation from the Rankine cycle, so it can observe that the cycle which is being run by water is the most efficient of all [25, 27].

### Benzene production unit

Now in this last alternative, applied the same approach to reducing the waste heat. Exchanged heat of different streams where to cut the need for coolers and heaters. Apart from that, power generation opportunities were also clear were had great sources of waste heat. Power generation from those points was the best alternative to use up the heat that is being wasted. In terms of heat exchangers, the best alternative found was the use of two heat exchangers. Revenue generated by both exchangers is nearly the same, but two exchangers' application is better than only one, as given in Table 28. Rankine, when operated with water where the boiler is run by electricity, is the most efficient way of reusing heat and generating power simultaneously [8, 27].

Looking at the alternatives, not considering the revenue generated by them is not enough, cannot estimate that either it is a fruitful alternative or not. The application of these ideas not only saves a great amount of money but also saves the environment by stopping harmful and highly thermally active gases from exhausting the atmosphere freely. So, going economically smart and saving the environment was the main goal achieved by using the simulation of various chemical plants. Cost optimization detailed insights into whether the alternative is feasible or not.

**Table 27: Energy and cost of sulfur removal plant alternatives.**

Case	Alternative	Energy Saved (kW)	Power generation (kW)	Boiler Operated with	Revenue (\$/yr)
1	Double heat exchangers	882.68	-	-	410344.28
2	Rankine with water	82.99	31.48	Electricity	38580
				Anthracite Coal	3460.95
3	Rankine with n-butane	38.42	14.62	Electricity	17860.89
				Anthracite coal	1602.04
4	Rankine with i-pentane	36.10	11.75	Electricity	16782.35
				Anthracite Coal	1505.308
5	Rankine in series with heat exchanger	965.59	31.48	Electricity	448924.28
				Anthracite Coal	413805.23

**Table 28: Energy and cost of benzene production plant alternatives.**

Case	Alternative	Energy Saved (kW)	Power generation (kW)	Boiler Operated with	Revenue (\$/yr)
1	Heat exchanger	1166.36	-	-	542223.58
2	Two heat exchangers	1174.62	-	-	546063
3	Rankine with water	1175.44	61.99	Electricity	546444.74
				Anthracite coal	49012.17
4	Rankine with n-butane	1175.74	97.82	Electricity	546584.2
				Anthracite coal	49026.3
5	Rankine with i-pentane	1166.4	61.39	Electricity	542223.58
				Anthracite coal	48634.11

And by comparing, they select the best alternative somewhere. It is the application of double heat exchangers, or Rankine is being run by water that is saving most of the thermal energy [10, 14]. Also, people are looking at energy cost as a way of energy management which is okay. That's one way of reducing the cost but the problem, of course, becomes when you have energy that costs less but still produces the same amount of carbon. So, energy management, energy cost, and energy efficiency are not necessarily always the same things. So, buy energy cheaper, but still, you will make the same amount of GHG, but if you reduce the amount of energy that you were using regardless of the price you pay, you reduce the GHGs. So, the only simple and easy method left is the optimization of the plants. Energy efficiency or thermal waste recovery has a great scope and can be done by numerous methods. Either could have designed any equipment which could store or reuse the thermal energy or could have applied different alternatives that could save cost and environment simultaneously.

The beauty of optimization is that it has no endpoints, and there is no specific solution for it. Different approaches will result in different answers, but the positive thing is all answers will have something fruitful in it. So, found few alternatives for the plants, so it does not mean that the study ends here. More focus will liberate different other possibilities for recovering the heat. And this is what the world needs the most right now. Optimization of existing plants means saving money and the earth. So, it can't be other people as well continue this effort to make more alternatives. And this can be done by the basic technical knowledge of engineering students. Maybe new alternatives are more efficient than these, which are mentioned in the above study.

## CONCLUSIONS

In this industrial era, industries' demands cannot be neglected means cannot simply remove them from our lives. It is a need in such a way that it makes the commodities that are



essential for every human being. Various chemical process involves high heat rejection and toxic exhaust of gases which indirectly pollutes the environment and increase global warming. Here the focus was mainly on optimizing heat that is usually wasted in various chemical plants. This research chose the approach of providing alternatives as it deals with every part of the plant and has many different solutions. Many chemical plants were observed and tried to find the spot for maximum wastage of heat, and three were selected which found a great opportunity to recover the heat that is being wasted. These include an amine treatment plant, a sulfur recovery unit, and a benzene production plant. In all three plants, flue gas was being exhausted at a very high temperature, so there was a pool of approaches applied to save the plant's economy. This work mainly utilized the heat of flue gas and used the flue gas in the heat exchanger, and cut the demand for heaters.

Similarly, for the demand for coolers, used any other stream which was cold, and a chiller was employed to heat it. After the application of these heat exchangers and fully exhausting the demand of the plant found that there is more energy left in the flue gas, which further can be utilized fruitfully. Application of organic Rankine cycle proved to be a good approach as it not only saved the heat but also produced some amount of power which is a plus anyhow. The Rankine cycle used different organic fluids, including water, n-butane, and i-pentane, and observed their behavior for the same temperature demand. This work chooses only these fluids as they are considered good refrigerating agents. So, the results were quite different by using different fluids for the purpose. The cost optimization gave detailed insights into what alternatives are good to be applied. So, in this manner, gathered a complete approach to how to cut extra costs, save the environment, and make any chemical plant efficient. If different equipment could be designed to reuse flue gas or to utilize its thermal energy to the maximum, more alternatives will have a strong comparative study which will yield maximum results in terms of energy management. The continuation of this study will create a safe environment and stop all sorts of pollution. Side by side, fuel use will also reduce as the power requirement will eventually be decreased when using equipment smartly.

#### Acknowledgments

The authors would like to acknowledge the Department of Chemical Engineering and Department of Polymer and

Petrochemical Engineering, NED University of Engineering & Technology, Karachi, Pakistan, for supporting this research work.

#### Abbreviations

BCS	Bardeen-Cooper-Schrieffer
BTU	British Thermal Unit
DEA	Diethylamine
DoE	Department of Energy
GHG	Green House Gas
HINT	Heat Integration Network
HEN	Heat Exchanger Network
HRGS	Heat Recovery Generator System
MW	Mega Watt
ORC	Organic Rankine Cycle
PNNL	Pacific Northwest National Laboratory
TEG	Thermo Electric Generator
T-H	Temperature-Enthalpy
TOE	Tons of Oil Equivalent
WHR	Waste Heat Recovery

Received : Nov. 18, 2020 ; Accepted : June 17, 2021

#### REFERENCES

- [1] Zhang J., Kan X., Shen Y., Loh K.-C., Wang C.-H., Dai Y., Tong Y.W., [A Hybrid Biological and Thermal Waste-To-Energy System with Heat Energy Recovery and Utilization for Solid Organic Waste Treatment](#), *Energy*, **152**: 214-222 (2018).
- [2] Reddy K.A., [A Critical Review of Thermal Waste Heat Recovery Systems](#), *Int. J. Res. Trends Innov.*, **2**: 113-117 (2017).
- [3] Hoseinzadeh S., Stephan Heyns P., [Advanced Energy, Exergy, and Environmental \(3E\) Analyses and Optimization of a Coal-Fired 400 MW Thermal Power Plant](#), *J. Energy Resour. Technol.*, **143**: 1-9 (2021).
- [4] Wemhoff A.P., Dai T., Fleischer A.S., [The Potential for Thermal Waste Energy Recovery in Industrial Kitchens](#), *Int. Mech. Eng. Congress Expo.*, **58417**: 1-9 (2017).
- [5] Sebastian R.M., Alappat B., [Thermal Properties of Indian Municipal Solid Waste over the Past, Present and Future Years and Its Effect on Thermal Waste to Energy Facilities](#), *Civil Eng. Urban Planning: Int. J.*, **3**: 97-106 (2016).

- [6] Hoseinzadeh S., Sohani A., Samiezadeh S., Kariman H., Ghasemi M.H., Using Computational Fluid Dynamics for Different Alternatives Water Flow Path in a Thermal Photovoltaic (PVT) System, *Int. J. Numer. Methods Heat Fluid Flow*, **1**: 1-20 (2020).
- [7] Miró L., Gasia J., Cabeza L.F., Thermal Energy Storage (TES) for Industrial Waste Heat (IWH) Recovery: A Review, *Appl. Energy*, **179**: 284-301 (2016).
- [8] Yu X., Li Z., Lu Y., Huang R., Roskilly A.P., Investigation of Organic Rankine Cycle Integrated With Double Latent Thermal Energy Storage for Engine Waste Heat Recovery, *Energy*, **170**: 1098-1112 (2019).
- [9] Hoseinzadeh S., Ghasemi M.H., Heyns S., Application of Hybrid Systems In Solution of Low Power Generation at Hot Seasons for Micro Hydro Systems, *Renewable Energy*, **160**: 323-332 (2020).
- [10] Moyaa D., Aldásb C., Lópeza G., Kaparajuc P., Municipal Solid Waste as a Valuable Renewable Energy Resource: a Worldwide Opportunity of Energy Recovery by Using Waste-To-Energy Technologies, *Energy Procedia*, **134**: 286-295 (2017).
- [11] Hoseinzadeh S., Moafi A., Shirkhani A., Chamkha A.J., Numerical Validation Heat Transfer of Rectangular Cross-Section Porous Fins, *J. Thermophys. Heat Transfer*, **33**: 698-704 (2019).
- [12] Tian Z., Gu B., Gao W., Zhang Y., Performance Evaluation of an Electric Vehicle Thermal Management System With Waste Heat Recovery, *Appl. Therm. Eng.*, **169**: 114976 (2020).
- [13] Xu Z.Y., Wang R.Z., Yang C., Perspectives for Low-Temperature Waste Heat Recovery, *Energy*, **176**: 1037-1043 (2019).
- [14] Ju X., Xu C., Liao Z., Du X., Wei G., Wang Z., Yang Y., A Review of Concentrated Photovoltaic-Thermal (CPVT) Hybrid Solar Systems with Waste Heat Recovery (WHR)", *Sci. Bull.*, **62**: 1388-1426 (2017).
- [15] Sohani A., Hoseinzadeh S., Berenjkari K., Experimental Analysis of Innovative Designs for Solar Still Desalination Technologies; An In-Depth Technical and Economic Assessment, *J. Energ. Storag.*, **33**: 1-15 (2021).
- [16] Tian Z., Gan W., Zhang X., Gu B., Yang L., Investigation on an Integrated Thermal Management System With Battery Cooling and Motor Waste Heat Recovery for Electric Vehicle, *Appl. Therm. Eng.*, **136**: 16-27 (2018).
- [17] Moreira L.F., Arrieta F.R.P., Thermal and Economic Assessment of Organic Rankine Cycles for Waste Heat Recovery in Cement Plants, *Renewable Sustainable Energy Rev.*, **114**: 1-20 (2019).
- [18] Xu H., Romagnoli A., Sze J.Y., Py X., Application of Material Assessment Methodology in Latent Heat Thermal Energy Storage for Waste Heat Recovery, *Appl. Energy*, **187**: 281-290 (2017).
- [19] Merlin K., Soto J., Delaunay D., Traonvouez L., Industrial Waste Heat Recovery Using An Enhanced Conductivity Latent Heat Thermal Energy Storage, *Appl. Energy*, **183**: 491-503 (2016).
- [20] Yih J., Wang H., Experimental Characterization of Thermal-Hydraulic Performance of a Microchannel Heat Exchanger for Waste Heat Recovery, *Energy Convers. Manage.*, **204**: 1-15 (2020).
- [21] Hussam J., Olabi A.G., Industrial Waste Heat Recovery, *Energy*, **1**: 1-2 (2018).
- [22] Serrano A., Feroso F.G., Alonso-Fariñas B., Rodríguez-Gutierrez G., Fernandez-Bolaños J., Borja R., Olive Mill Solid Waste Biorefinery: High-Temperature Thermal Pre-Treatment for Phenol Recovery and Biomethanization, *J. Cleaner Prod.*, **148**: 314-323 (2017).
- [23] Elankovan R., Suresh S., Karthick K., Hussain M.M.M.D., Chandramohan V.P., Evaluation of Thermoelectric Power Generated Through Waste Heat Recovery From Long Ducts and Different Thermal System Configurations, *Energy*, **185**: 477-491 (2019).
- [24] Mumbach G.D., Alves J.L.F., Da Silva J.C.G., De Sena R.F., Marangoni C., Machado R.A.F., Bolzan A., Thermal Investigation of Plastic Solid Waste Pyrolysis Via the Deconvolution Technique Using the Asymmetric Double Sigmoidal Function: Determination of the Kinetic Triplet, Thermodynamic Parameters, Thermal Lifetime and Pyrolytic Oil Composition for Clean Energy Recovery, *Energy Convers. Manage.*, **200**: 1-12 (2019).
- [25] Zabek D., Morini F., Solid State Generators and Energy Harvesters for Waste Heat Recovery and Thermal Energy Harvesting, *Therm. Sci. Eng. Prog.*, **9**: 235-247 (2019).
- [26] Ji C., Qin Z., Dubey S., Choo F.H., Duan F., Three-Dimensional Transient Numerical Study on Latent Heat Thermal Storage for Waste Heat Recovery from a Low Temperature Gas Flow, *Appl. Energy*, **205**: 1-12 (2017).

- [27] Shu G., Zhao M., Tian H., Wei H., Liang X., Huo Y., Zhu W., [Experimental Investigation on Thermal OS/ORC \(Oil Storage/Organic Rankine Cycle\) System for Waste Heat Recovery from Diesel Engine](#), *Energy*, **107**: 693-706 (2016).