Heat Integration using Pinch Analysis of Ammonia Plant

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Certificate

This is to certify that work in this thesis has been carried out by **Mr. Maqbool Khalid, Miss Sumiah Naveed** and **Mr. Mohammad Talal** and completed under my supervision in School of Chemical and Materials Engineering, National University of Sciences and Technology, H-12, Islamabad, Pakistan.

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Dedication

We dedicate our work to our parents and teachers, who helped us constantly to reach this point in our lives, and to Lt Col Retd. Nadeem Ehsan and Dr. Muhammad Nouman Aslam Khan for continuous support and love that they bestowed upon us.

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Thanks to **Almighty Allah** for His countless blessings, He gave us courage to finish this project.

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We are able to achieve this milestone due to love and prayers of our beloved **parents**.

Nomenclature

ΔT_{LM}	Log Mean Temperature Difference
T	Temperature
ΔT_{min}	Minimum Temperature Difference
ΔT	Temperature Difference
T_S	Supply Temperature
T_T	Target Temperature
c_p	Specific Heat Capacity
C_P	Heat Capacity
M	Mass Flowrate
Q	Heat
W	Work
ΔH	Change in Enthalpy
h	Heat Transfer Coefficient
U	Overall Heat Transfer Coefficient
A	Area
Wt%	Weight Percentage
L	Length of Coil

p	Pitch of coil
R	Helix Radius
d_o	Coil Outside Diameter
N	Number of Turn in Coil
^oC	Degree Celsius
m	Metre
m^2	Square Metre
mm	Millimetre
kJ	Kilojoule
kg	Kilogram
hr	Hour
W	Watt
kW	Kilowatt
Gcal	Giga calorie
PCE	Purchase Cost of Equipment
PCC	Total Physical Plant Cost

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Abstract

Around 80% of ammonia produced is being used for the manufacturing of urea-based fertilizers, however its production requires extensive amount of energy.

Retrofitting of an existing ammonia plant for improving energy efficiency is done by using Pinch Analysis. This method relates the core process with the utility system to reduce the energy consumption and heat waste of the plant. The technique is applied on the convective section of the primary reformer by changing the arrangement of heating coils and by doing this significant waste heat was recovered from the flue gases. Otherwise, this heat would have lost in atmosphere. The analysis was performed using Aspen Energy Analyser. The problem came out be threshold problem i.e. the system requires only one utility (cold utility) to meet the desired temperatures of different streams. Using thermodynamics principles, pinch rules and limitation imposed by the current plant design, new modifications are suggested to improve the plant efficiency. As a result of these suggestions, around 7% more high pressure steam can be generated with only slight modifications.

Chapter 1
Introduction

1.1 Background

Fauji Fertilizer Company Limited (FFC) was established back in 1978 and is one of the largest chemical fertilizer producer in Pakistan. The major products of FFC are urea and ammonia and have a capacity to produce more than 6000 metric tons of urea per day.

Ammonia is one of the most important chemical in the field of agriculture, as 80% of the produced ammonia is being used for manufacturing of fertilizers like urea, ammonium nitrate and ammonium phosphate, etc. Ammonia is also used in industrial refrigeration systems as coolant, for water and wastewater treatment, for making urea nitrate and for several metallurgical process. These mentioned uses of ammonia show its importance but the production of ammonia gas requires large amount of energy. Hence we need to design energy efficient plant with minimum heat wastage and maximum recovery.

The technique through which FFC used to produce ammonia in 1978 is now obsolete as new methods are developed using modern technology. The aim of using modern technology is to develop efficient plants that give maximum output for the same amount of inputs.

1.2 Problem Statement

Currently in Plant II of FFC, the flue gases from primary reformer and auxiliary boiler are leaving the system at sufficiently high temperature into atmosphere. This shows the plant is not being operated to its full potential along with high adverse effects on the environment. The flue gases are the source for heat recovery and by using modern engineering tools this recoverable energy can be used to heat other process stream or to produce steam.

1.3 Purpose of Study

No matter how carefully the process is designed, there is always some room to improve the process efficiency through different modifications. Pinch Analysis is one of the advance technique through which industry can modify their process for heat integration. It is a well-defined method to relate plant's core streams with the utility system, such that maximum energy is recovered.

1.4 What is Pinch Analysis?

It was first introduced by Bodo Linnhoff and Vredeveld. Pinch analysis is a methodology for minimising energy consumption of chemical processes by calculating thermodynamically feasible energy targets (or minimum energy consumption).

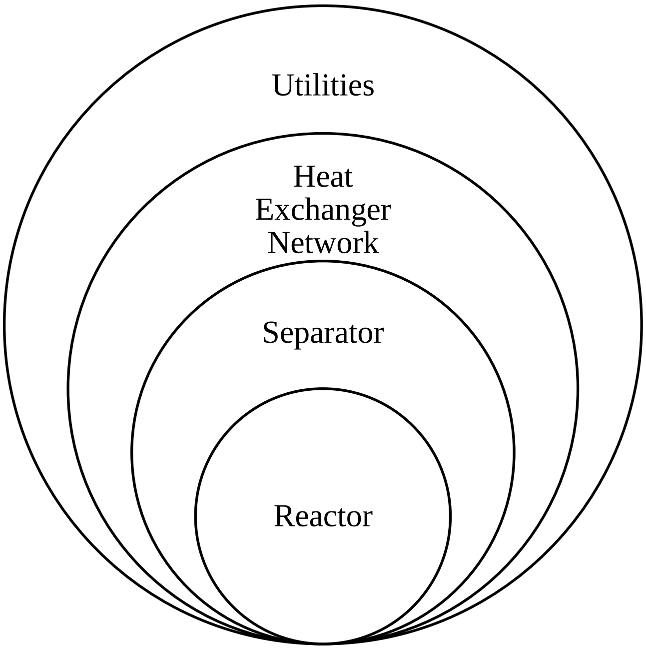


Figure 1.1 Onion Diagram

Designing of new process start with the reactors i.e. the core of the onion. Separators, the second layer of onion, are designed once feed, product, flowrates and recycle concentrations are known. After complete material and energy balance is performed on the new system, heat exchanger network is designed. This is where pinch analysis comes in, it provides a systematic approach to relate core process (first and second layer of onion) with the utility system (fourth layer of onion), while saving energy. Pinch analysis defines targets including minimum energy requirement, exchanger area and minimum cost prior to heat exchanger network design. The heat exchanger network designed through pinch rules ensure that these targets are meet and energy is saved. Thus, the prime objective of the technique is to achieve financial saving by maximizing process to process heat recovery and reducing the external utility loads, i.e. by better process heat integration.

Now a days, number of software have been developed to perform pinch analysis on industries complex process with speed and efficiency. Aspen Energy Analyzer is one of the famous product of Aspen One that is used to perform pinch analysis. The problem in this thesis is done by using Aspen Energy Analyzer.

Chapter 2
Literature Review

2.1 Process Integration

Process integration means to optimize a process for reducing the amount of raw material, minimize waste and utility requirement. It can be done by recycling the useful waste materials and recover waste heat that is being generated by different processes.

Process integration has two divisions.

1. Mass Integration
2. Heat Integration

2.1.1 Mass Integration

Objective of mass integration are:

- To achieve maximum product yield.
- Minimizing waste material.
- Decreasing the requirement of fresh feed.

2.1.2 Heat Integration

Objective of heat integrations:

- Calculating minimum heating and cooling duties to operate plant.
- Designing heat exchanger network to meet the initially set targets.
- Converting waste heat into valuable energy.

Different techniques used to perform process integration are:

- Pinch Analysis
- Process Graph theory Process
- State space approach
- Genetic Algorithm

Out of these, pinch analysis is one of the most common technique and widely used technique to perform process (heat) integration.

2.2 Pinch Analysis

Pinch analysis is based on the principles of thermodynamics. The First Law of thermodynamics helps to determine the change in enthalpy (ΔH) of the streams passing through heat exchanger. The Second Law determines the direction of heat flow that is from hot stream to cold stream. Temperature crossovers of the hot and cold stream profile (temperature vs enthalpy) is restricted by second law. Temperature crossover means that hot stream cannot be cooled to the temperature lower than cold stream supply temperature and vice versa. Temperature approach of heat exchanger defines the limit to which a hot stream can be cooled to. This is the minimum allowable temperature difference (ΔT_{min}) between hot and cold stream. The point at which minimum temperature difference (ΔT_{min}) occur in known as pinch point in the profile. The pinch defines the minimum driving force allowed in the exchanger unit.

The goal of any process industry is to minimize the utility requirement and maximize process to process stream heat recovery. An optimize heat exchanger network is required for minimum energy requirement. To design the most suitable heat exchanger network, pinch analysis provides well defined steps. These can be used for a new project as well as for retrofitting and existing one.

2.3 Steps of Pinch Analysis

Steps used to perform pinch analysis are:

- Identification of hot, cold & utility streams in the process.
- Thermal data extraction for the process.
- Selection of initial minimum temperature difference (ΔT_{min}) value.
- Construction of composite curve & grand composite curve.
- Estimation of minimum energy cost targets.

- Estimation of heat exchanger network capital cost targets.
- Estimation of optimum minimum temperature difference (ΔT_{min}) value.
- Estimation of practical targets of heat exchanger network design.
- Design of heat exchanger network.

Details of these steps are discussed below:

2.3.1 Identification of Hot, Cold & Utility Streams in the Process

This step is the simplest yet the most crucial step in which process flow diagram of the plant is carefully analysed to identify hot, cold and utility streams. Hot streams are those which need to be cooled while cold stream are those which must be heated. Reactant stream normally require preheating before entering a reactor and hence are cold stream. If heat exchange between the process streams is not feasible or economical then utility streams are used as coolant to heating agents. Hot utilities include steam, hot oil or flue gases, etc. Cold utilities include cooling water, steam generation or air, etc.

2.3.2 Thermal Data Extraction for the Process Streams

Following stream data is required to perform pinch analysis:

- **Supply Temperature (T_S ^oC)**
- **Target Temperature (T_T ^oC)**
- **Heat Capacity Flow Rate (C_P kW/^oC)**

  Heat capacity flow rate is the product of mass flow rate (M kg/s) and specific heat capacity (c_p kJ/kg ^oC).

$$C_P = M \times c_p$$

- **Enthalpy Change (ΔH kW)**

  According to first law of thermodynamics:

$$\Delta H = Q \pm W$$

For heat exchangers,

$$W = 0$$

So,

$$\Delta H = Q$$

Moreover,

$$Q = C_P \, (T_S - T_T)$$

- **Heat Transfer Coefficients (h W/m^2 ^{o}C)**

  Heat transfer coefficients are required to calculate over heat transfer coefficient (U W/m^2 ^{o}C**)** between the hot and cold streams in the heat exchanger. The overall heat transfer coefficient are used to calculate the area of heat exchangers.

$$\frac{1}{U} = \frac{1}{h_{Hot}} + \frac{1}{h_{Cold}}$$

$$A = \frac{Q}{U \times \Delta T_{LM}}$$

$$\Delta T_{LM} = \frac{(T_{Hin} - T_{Cout}) - (T_{Hout} - T_{Cin})}{\ln \frac{(T_{Hin} - T_{Cout})}{(T_{Hout} - T_{Cin})}}$$

Where ΔT_{LM} is log mean temperature difference.

2.3.3 Selection of Initial ΔTmin Value

For feasible heat transfer design, minimum driving force is required that is ΔTmin has to be maintained between the streams. Value of minimum temperature difference (ΔT_{min}) will depend on the physical properties of the exchanging streams and the type of heat exchanger that is being used.

Importance of selecting suitable minimum temperature difference (ΔT_{min}) value can be determine for the fact that it is the main controlling factor for both capital and energy costs. For a particular value of heat load (Q), if smaller value of minimum temperature difference (ΔT_{min}) is selected, then area requirement by the exchangers rises. For the higher value, heat recovery will be reduced and utilities demand increase to meet the desired results.

2.3.4 Construction of Composite & Grand Composite Curve

2.3.4.1 Composite curves

Composite curves are temperature vs enthalpy (T-H) plot that is used to determine energy target prior to design. Combine composite curve is plot between the hot and cold stream. The minimum distance between these two composite curves is ΔT_{min}. The point at which ΔT_{min} occur is known as pinch point. Hot composite curve represents the heat source in the system while cold composite curve are the heat sink. Overlapping area of the two curves are maximum recoverable energy. The hot composite curve portion extended on left side of the plot beyond the cold composite curve represents minimum cold utility requirement while extended cold composite curve on right side of the plot represents minimum hot utility requirement. However, there is always a chance of error while using graphical method. To eliminate these error, Linnhoff developed a numerical approach known as the problem table algorithm to determine energy requirement.

2.3.4.2 Grand Composite curve

Grand composite curve is used to determine the type of utility required to satisfy supply temperatures of the process streams. It represents the variation of heat supply and demand within the process. GCC are used to maximize the use of cheaper utilities and minimize the requirement of expensive utility.

Method to plot the composite and grand composite will be discussed in later chapters while solving the project.

2.3.4.3 Threshold Problem

If the plant’s streams require only one utility that is either hot or cold utility then such problem is known as threshold problem. The pinch point for these type of problem lies on the one side of the plot rather than in the middle. If the plant requires only cold utility, then pinch will lie on the right end of the plot and if it requires hot utility, then pinch will lie in the left end of the plot. However, the problem will remain single utility dependent till a value of ΔT_{min} known as threshold temperature. After this, plant will require both hot and cold utilities.

2.3.5 Estimation of Minimum Energy Cost Targets

After determining the minimum utilities requirement and the type of utility to be used, energy cost can be determined by using following equation.

$$Total\ Energ\ Cost = \sum_{i=1}^{i} Q_i \times C_i$$

Where

Q_i = Duty of utility i (kW)

C_i = Unit cost of utility i ($/kW yr)

i = Total number of utilities used

2.3.6 Estimation of Heat Exchanger Network Capital Cost Targets

Capital cost of the heat exchanger networks is dependent upon:

- Number of exchangers
- Overall network area
- Distribution of area between the exchangers.

While targeting the area before the network design, it is assumed that area is evenly distributed between the units.

2.3.6.1 Area Targeting

Minimum heat exchanger network area is calculated through following relation:

$$HEN\ Area_{min} = \sum_i \left[\frac{1}{\Delta T_{LM}} \sum_j \frac{q_j}{h_j} \right]$$

2.3.6.2 Number of Units Targeting

Minimum number of exchanger units are calculated by:

$$N_{min} = [N_h + N_c + N_u - 1]_{Above\ Pinch} + [N_h + N_c + N_u - 1]_{Below\ Pinch}$$

Where:

N_h = Number of hot streams

N_c = Number of cold streams

N_u = Number of utility streams

2.3.6.3 Heat Exchanger Network Capital Cost Targeting

After determining minimum area heat exchanger area and minimum number of units, capital cost of the heat exchanger can be determined by:

$$C(\$)_{HEN} = \left[N_{min}\left\{a + b\ \left({}^{A_{min}}/_{N_{min}}\right)\right\}^{c}\right]_{Above\ Pinch}$$

$$+ \left[N_{min}\left\{a + b\ ({}^{A_{min}}/_{N_{min}})\right\}^{c}\right]_{Below\ Pinch}$$

$$where\ a, b\ and\ are\ constants\ in\ exchanger\ cost\ law$$

$$Exchanger\ cost\ (\$) = a + b\ (Area)^{c}$$

2.3.7 Estimation of Optimum ΔT_{min} Value

ΔT_{min} can affect the system in following ways:

- Increasing minimum temperature difference (ΔT_{min}) value will give higher energy costs but lower capital cost.
- Decreasing minimum temperature difference (ΔT_{min}) value will give lower energy costs but higher capital cost.
- Optimum minimum temperature difference (ΔT_{min}) exists where both total annual cost of energy and capital costs are minimized.

2.3.8 Estimation of Practical Targets of Heat Exchanger Network Design

In some cases optimum value of minimum temperature difference (ΔT_{min}) might not be suitable for appropriate design. If the value small that means requires area will be large and network design will be much complicated. In these situations, higher value of minimum temperature difference (ΔT_{min}) is preferred if the effect on marginal cost is low.

The pinch temperature divides the system in two parts such that each portion is in enthalpy balance with its utility. Following points must be kept in while designing a network

- No external heating below the pinch.
- No external cooling above the pinch.
- No heat transfer across the pinch.

If any of above mentioned points are not followed while designing a network, then more energy will be required than theoretically calculated using combine composite curves.

2.3.8.1 Plus/Minus Principle

There are several factors in the plant system that can be changed for changing the energy consumption of plant. This will result changes in the energy and material balance. The change in these factor may have a favourable impact on energy consumption, this is known as plus/minus principle.

Following are the guidelines that can be used to reduce utility loads.

- Increasing (+) in hot stream duty or decreasing (-) in cold stream duty above the pinch will reduce utility load.
- Increasing (+) in cold stream duty or decreasing (-) in hot stream duty below the pinch will reduce utility load.

2.3.9 Design of Heat Exchanger Network

Heat exchanger network is designed using Pinch Design Method (PDM). The pinch temperature divides the network grid diagram into two parts that are above and below the pinch. Heat exchanger networks for both part are designed separately. When no more process to process heat exchange is possible then utility streams are added in the grid. Some of the basic design rules to identify between which two stream an exchange is possible are given below:

- Divide the problem at the pinch.
- Design away from the pinch.
- Above the pinch,
    - match streams adjacent to the pinch if,

      $$CP_{Hot} \leq CP_{Cold}$$

    - $N_{Hot} \leq N_{Cold}$

      Where:

      N_{Hot} = Number of hot streams

      N_{Cold} = Number of Cold Streams
    - Use hot utility
- Below the pinch,
    - match streams adjacent to the pinch if,

      $$CP_{Cold} \leq CP_{Hot}$$

    - $N_{Cold} \leq N_{Hot}$

      Where:

      N_{Hot} = Number of hot streams

      N_{Cold} = Number of Cold Streams
    - Use cold utility
- If the stream matching criteria is not satisfied then split a stream.
- Maximize the exchanger heat loads.

In the following chapters, these steps are used to perform the pinch analysis on the ammonia plant.

Chapter 3
Process Description

In this chapter, the existing network of coils in discussed in detail followed by extraction of data required to perform pinch analysis.

3.1 Process Flow Diagram

In the convective section of primary reformer, series of coils are install to preheat natural gas, process air, boil feed water and to generate steam to recover heat from the flue gas.

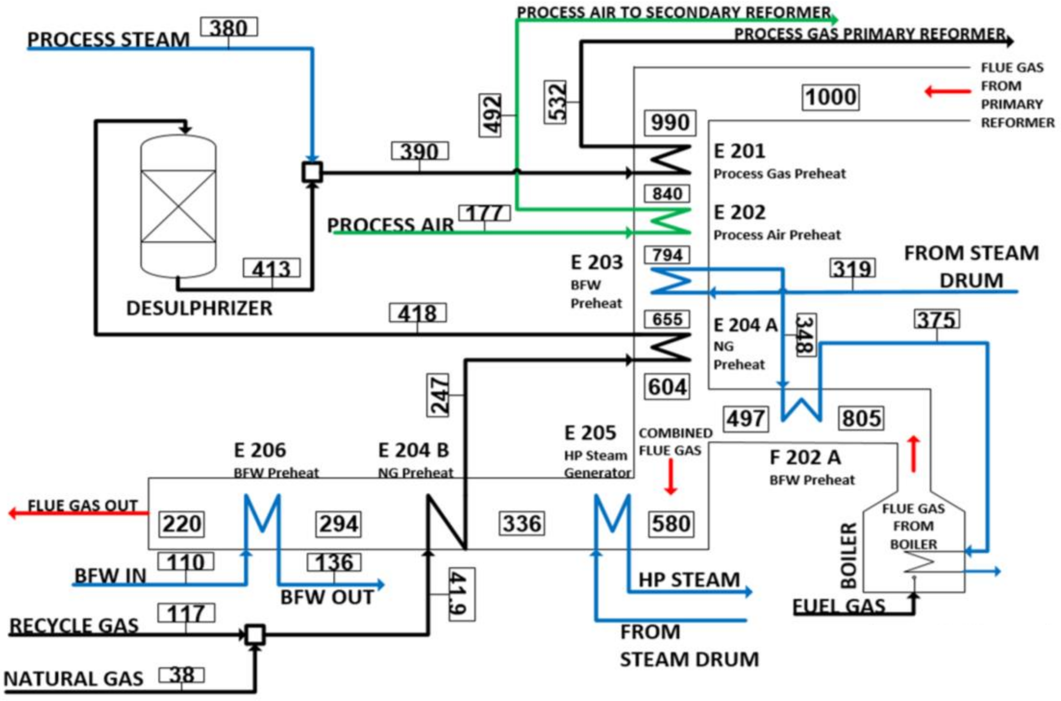


Figure 3.1: Existing Process Flow Diagram

Legend

Temperature	□
Natural Gas	—
Air	—
Steam/Water	—
Flue Gas	—

3.1.1 Natural Gas

Natural Gas at 38 °C is mixed with Recycled Gas coming at 117 °C and passed through two preheaters E 204 B and E 204 A, in respective order. Mixture of these gases then entered the desulphurizer at 418°C. Sulphur content of natural gas is reduced from 0.9 ppm to 5 ppb.

The Desulphurized Gas is mixed with process steam and heated in coil E 201 to 532 °C before entering into the primary reformer.

3.1.2 Process Air

Process Air after being compressed in preheated in coil E 202 from 177 °C to 492°C.

3.1.3 Steam from Steam Drum

Steam from steam drum is first heated in coil E 203 through flue gas from primary reformer and then in F 202 A through flue gas coming from boiler.

3.1.4 Boiler Feed Water

Boiler Feed Water is heated in coil E 206 from 110 °C to 136 °C through stack gases.

3.1.5 Flue Gases

Flue gas coming from primary reformer is passed through four series of coils, namely E 201, E 202, E 203 and E 204 A, its temperature drops from 990 °C to 604 °C and then mixed with flue gas from the boiler. These combine flue gases are used to produce high pressure steam in coil E 205. Before exhausting into the environment at 220 °C, they preheat natural gas and boiler feed water in coil E 204 B and E 206, respectively.

Now we will perform data extraction of the process.

3.2 Data Extraction

For data extraction, complete material and energy balance of the system is required. Following data is required for pinch analysis:

- Streams supply temperatures (T_S °C)
- Streams target temperatures (T_T °C)
- Streams mass flow rates (M kg/h)
- Streams specific heat capacities (c_p kJ/kg °C)
- Streams head load (Q kW)
- Heat transfer coefficients (h W/m^2 °C)

3.3 Material Balance

Material balance is performed to calculate the mass flow rates and well the composition of streams. Detail material balance is given below:

3.3.1 Mass Flowrates and Compositions of Known Streams

Following data is provided by the industry

Natural Gas

Components	M (kg/hr)	Wt%
N_2	9435	24.67%
CO_2	6588.214	17.23%
CH_4	22104.286	57.80%
C_2H_6	112.5	0.29%
Flow (Total)	**38240**	**100%**

Table 3-1: Natural Gas Composition

Recycled Gas

Components	**M (kg/hr)**	**Wt%**
H_2	172.260	17.30%
N_2	798.750	80.22%
CO	0.000	0.00%
CO_2	0.000	0.00%
Ar	8.929	0.90%
CH_4	15.000	1.51%
C_2H_6	0.000	0.00%
H_2O	0.804	0.08%
Flow (Total)	**995.742**	**100%**

Table 3-2: Recycle Gas Composition

Process Air

Components	**M (kg/hr)**	**Wt%**
O_2	10585.71	23.12%
N_2	34436.25	75.20%
CO_2	21.61	0.05%
Ar	592.86	1.29%
H_2O	153.48	0.34%
Flow (Total)	**45789.91**	**100%**

Table 3-3: Process Gas Composition

Process Steam

Components	**M (kg/hr)**	**Wt%**
H_2O	**103950**	**100%**

Table 3-4: Process Steam Composition

Steam from Steam Drum

Components	**M (kg/hr)**	**Wt%**
H_2O	**255379**	**100%**

Table 3-5: Steam from Steam Drum Composition

Boiler Feed Water

Components	**M (kg/hr)**	**Wt%**
H_2O	257959	100%

Table 3-6: Boiler Feed Water Composition

Flue Gas from Primary Reformer

Components	**M (kg/hr)**	**Wt%**
O_2	4305.25	1.70%
N_2	180006.77	71.19%
CO_2	37765.50	14.94%
Ar	3007.13	1.19%
H_2O	27776.39	10.98%
Flow (Total)	**252861.03**	**100%**

Table 3-7: Primary Reformer Flue Gas Composition

Flue Gas from Boiler

Components	**M (kg/hr)**	**Wt%**
O_2	1944.16	2.60%
N_2	53362.16	71.37%
CO_2	10702.51	14.31%
Ar	892.63	1.19%
H_2O	7870.61	10.53%
Flow (Total)	**74772.08**	**100%**

Table 3-8: Boiler Flue Gas Composition

However, information about combine Natural and Recycled Gas, Process Gas and Combine Flue Gas is need to be calculated used material balance techniques.

3.3.2 Mass Flowrates and Compositions of Unknown Streams

3.3.2.1 Mass Flowrate Calculations

Material Balance around Mixer of Natural and Recycled Gas

Let X be the flow rate of inlet stream Natural Gas = 38240 kg/hr

Let Y be the flow rate of inlet stream Recycle Gas = 995.742 kg/hr

Let Z be the flow rate of outlet stream (Natural and Recycled Gas) =?

$$X + Y = Z$$

$$38240 + 995.742 = Z$$

So, Z = 39235.742 kg/hr

Component Balance

N_2 Balance

Let x be the mass fraction of N_2 in inlet stream Natural Gas = 0.2467

Let y be the mass fraction of N_2 in inlet stream Recycled Gas = 0.8022

Let z be the mass fraction of N_2 in outlet stream = ?

$$38240 \text{ x } 0.2467 + 995.742 \text{ x } 0.8022 = 39235.742 \text{ x } z$$

So, **z = 0.2608**

H_2 Balance

Let x be the mass fraction of H_2 in inlet stream Natural Gas = 0

Let y be the mass fraction of H_2 in inlet stream Recycled Gas = 0.1730

Let z be the mass fraction of H_2 in outlet stream = ?

$$38240 \text{ x } 0 + 995.742 \text{ x } 0.1730 = 39235.742 \text{ x } z$$

So, **z = 0.0044**

CO_2 Balance

Let x be the mass fraction of CO_2 in inlet stream Natural Gas = 0.1723

Let y be the mass fraction of CO_2 in inlet stream Recycled Gas = 0

Let z be the mass fraction of CO_2 in outlet stream = ?

$$38240 \times 0.1723 + 995.742 \times 0 = 39235.742 \times z$$

So, **z** = **0.1679**

Ar Balance

Let x be the mass fraction of Ar in inlet stream Natural Gas = 0

Let y be the mass fraction of Ar in inlet stream Recycled Gas = 0.0090

Let z be the mass fraction of Ar in outlet stream = ?

$$38240 \times 0 + 995.742 \times 0.0090 = 39235.742 \times z$$

So, **z** = **0.0002**

CH_4 Balance

Let x be the mass fraction of CH_4 in inlet stream Natural Gas = 0.5780

Let y be the mass fraction of CH_4 in inlet stream Recycled Gas =0.0151

Let z be the mass fraction of CH_4 in outlet stream = ?

$$38240 \times 0.5780 + 995.742 \times 0.0151 = 39235.742 \times z$$

So, **z** = **0.5638**

C_2H_6 Balance

Let x be the mass fraction of C_2H_6 in inlet stream Natural Gas = 0.0029

Let y be the mass fraction of C_2H_6 in inlet stream Recycled Gas = 0

Let z be the mass fraction of C_2H_6 in outlet stream = ?

$$38240 \text{ x } 0.0029 + 995.742 \text{ x } 0 = 39235.742 \text{ x } z$$

So, **z = 0.0029**

H_2O Balance

Let x be the mass fraction of H_2O in inlet stream Natural Gas = 0

Let y be the mass fraction of H_2O in inlet stream Recycled Gas = 0.0008

Let z be the mass fraction of H_2O in outlet stream = ?

$$38240 \text{ x } 0.7380 + 995.742 \text{ x } 0.0081 = 39235.742 \text{ x } z$$

So, **z = 0**

Components	**Natural and Recycle Gas Mass Fraction**	**Wt%**
N_2	0.1839	18.39%
H_2	0.00430	0.43%
CO_2	0.00754	0.75%
Ar	0.0001	0.01%
CH_4	0.6957	69.75%
C_2H_6	0.0019	0.19%
H_2O	0	0%
Total	**1**	**100%**
M (kg/hr)	**39235.7**	

Table 3-9: Natural & Recycle Gas Composition

Material Balance around Mixer of Process Steam and Desulphurized Gas

Let X be the flow rate of inlet stream Process Steam = 103950 kg/hr

Let Y be the flow rate of inlet stream Desulphurized Gas = 39235.4 kg/hr

Let Z be the flow rate of outlet stream Process Gas = ?

$$X + Y = Z$$

$$103950 + 39235.4 = Z$$

So, Z = **105701.6 kg/hr**

Component Balance

O_2 Balance

Let x be the mass fraction of O_2 in inlet stream Process Steam = 0

Let y be the mass fraction of O_2 I inlet stream Desulphurized Gas = 0.0044

Let z be the mass fraction of O_2 in outlet stream Process Gas = ?

$$10395 \text{ x } 0 + 39235.4 \text{ x } 0.0044 = 105701.6 \text{ x } z$$

So, z = 0

H_2 Balance

Let x be the mass fraction of H_2 in inlet stream Process Steam = 0

Let y be the mass fraction of H_2 in inlet stream Desulphurized Gas = 0.2608

Let z be the mass fraction of H_2 in outlet stream Process Gas = ?

$$103950 \text{ x } 0 + 39235.4 \text{ x } 0.2608 = 105701.6 \text{ x } z$$

So, **z =0.00007**

N_2 Balance

Let x be the mass fraction of N_2 in inlet stream Process Steam = 0

Let y be the mass fraction of N_2 in inlet stream Desulphurized Gas = 0

Let z be the mass fraction of N_2 in outlet stream Process Gas = ?

$$103950 \times 0 + 39235.4 \times 0 = 105701.6 \times z$$

So, **z = 0.00432**

CO Balance

Let x be the mass fraction of CO in inlet stream Process Steam = 0

Let y be the mass fraction of CO in inlet stream Desulphurized Gas = 0.1679

Let z be the mass fraction of CO in outlet stream Process Gas = ?

$$103950 \times 0 + 39235.4 \times 0.1679 = 105701.6 \times z$$

So, z = 0

CO_2 Balance

Let x be the mass fraction of CO_2 in inlet stream Process Steam = 0

Let y be the mass fraction of CO_2 in inlet stream Desulphurized Gas = 0.0002

Let z be the mass fraction of CO_2 in outlet stream (Process Gas) = ?

$$103950 \times 0 + 39235.4 \times 0.0002 = 105701.6 \times z$$

So, z = 0.00278

Ar Balance

Let x be the mass fraction of Ar in inlet stream Process Steam = 0

Let y be the mass fraction of Ar in inlet stream 2 Desulphurized Gas = 0.5638

Let z be the mass fraction of Ar in outlet stream Process Gas = ?

$$103950 \text{ x } 0 + 39235.4 \text{ x } 0.5638 = 105701.6 \text{ x } z$$

So, z = 0

CH_4 Balance

Let x be the mass fraction of CH_4 in inlet stream Process Steam = 0

Let y be the mass fraction of CH_4 in inlet stream Desulphurized Gas = 0.0029

Let z be the mass fraction of CH_4 in outlet stream Process Gas = ?

$$103950 \text{ x } 0 + 39235.4 \text{ x } 0.0029 = 105701.6 \text{ x } z$$

So, **z = 0.00934**

C_2H_6 Balance

Let x be the mass fraction of C_2H_6 in inlet stream Process Steam = 0

Let y be the mass fraction of C_2H_6 in inlet stream Desulphurized Gas = 0

Let z be the mass fraction of C_2H_6 in outlet stream Process Gas = ?

$$103950 \text{ x } 0 + 39235.4 \text{ x } 0 = 105701.6 \text{ x } z$$

So, **z = 0.00005**

H_2O Balance

Let x be the mass fraction of H_2O in inlet stream Process Steam = 1

Let y be the mass fraction of H_2O in inlet stream Desulphurized Gas = 0

Let z be the mass fraction of H_2O in outlet stream Process Gas = ?

$$103950 \times 1 + 39235.4 \times 0 = 105701.6 \times z$$

So, **z = 0.9834**

Components	**Process Gas Mass Fraction**	**Wt%**
O_2	0	0%
H_2	0.00007	0.007%
N_2	0.00432	0.432%
CO	0	0%
CO_2	0.00278	0.278%
Ar	0	0%
CH_4	0.00934	0.934%
C_2H_6	0.00005	0.005%
H_2O	0.9834	98.34%
Total	**1**	**100%**
M (kg/hr)	**105701.6**	

Table 3-10: Process Gas Composition

Material balance around Mixer of Flue Gas from Primary Reformer and Boiler

Let X be the flow rate of inlet stream of Primary Reformer Flue Gas = 252861.03 kg/hr

Let Y be the flow rate of inlet stream of Boiler Flue Gas = 74772.08 kg/hr

Let Z be the flow rate of outlet stream Combine Flue Gas = ?

$$X + Y = Z$$

$$252861.03 + 74772.08 = Z$$

So, **Z = 327633.11 kg/hr**

Component balance

O_2 Balance

Let x be the mass fraction of O_2 in inlet stream of Primary Reformer Flue Gas = 0.0170

Let y be the mass fraction of O_2 in inlet stream of Boiler Flue Gas = 0.0260

Let z be the mass fraction of O_2 in outlet stream of Combine Flue Gas = ?

$$252861.03 \text{ x } 0.0170 + 74772.08 \text{ x } 0.0260 = 327633.11 \text{ x } z$$

So, **z = 0.0191**

H_2 Balance

Let x be the mass fraction of H_2 in inlet stream of Primary Reformer Flue Gas = 0

Let y be the mass fraction of H_2 in inlet stream of Boiler Flue Gas = 0

Let z be the mass fraction of H_2 in outlet stream of Combine Flue Gas = ?

$$252861.03 \text{ x } 0 + 74772.08 \text{ x } 0 = 327633.11 \text{ x } z$$

So, **z = 0**

N_2 Balance

Let x be the mass fraction of N_2 in inlet stream of Primary Reformer Flue Gas = 0.7119

Let y be the mass fraction of N_2 in inlet stream of Boiler Flue Gas = 0.7137

Let z be the mass fraction of N_2 in outlet stream Combine Flue Gas = ?

$$252861.03 \times 0.7119 + 74772.08 \times 0.7137 = 327633.11 \times z$$

So, **z** = **0.7123**

CO Balance

Let x be the mass fraction of CO in inlet stream of Primary Reformer Flue Gas = 0

Let y be the mass fraction of CO in inlet stream of Boiler Flue Gas = 0

Let z be the mass fraction of CO in outlet stream Combine Flue Gas = ?

$$252861.03 \times 0 + 74772.08 \times 0 = 327633.11 \times z$$

So, **z** = **0**

CO_2 Balance

Let x be the mass fraction of CO_2 in inlet stream of Primary Reformer Flue Gas = 0.1494

Let y be the mass fraction of CO_2 in inlet stream of Boiler Flue Gas = 0.1431

Let z be the mass fraction of CO2 in outlet stream Combine Flue Gas = ?

$$252861.03 \times 0.1494 + 74772.08 \times 0.1431 = 327633.11 \times z$$

So, **z** = **0.1479**

Ar Balance

Let x be the mass fraction of Ar in inlet stream of Primary Reformer Flue Gas = 0.0119

Let y be the mass fraction of Ar in inlet stream of Boiler Flue Gas = 0.0119

Let z be the mass fraction of Ar in outlet stream Combine Flue Gas = ?

$$252861.03 \text{ x } 0.0119 + 74772.08 \text{ x } 0.0119 = 327633.11 \text{ x } z$$

So, **z** = **0.0191**

CH_4 Balance

Let x be the mass fraction of CH_4 in inlet stream Primary Reformer Flue Gas = 0.2467

Let y be the mass fraction of CH_4 in inlet stream of Boiler Flue Gas = 0

Let z be the mass fraction of CH_4 in outlet stream Combine Flue Gas =?

$$252861.03 \text{ x } 0.2467 + 74772.08 \text{ x } 0 = 327633.11 \text{ x } z$$

So, **z** = **0**

C_2H_6 Balance

Let x be the mass fraction of C_2H_6 in inlet stream Primary Reformer Flue Gas = 0

Let y be the mass fraction of C_2H_6 in inlet stream of Boiler Flue Gas = 0

Let z be the mass fraction of C_2H_6 in outlet stream Combine Flue Gas = ?

$$252861.03 \text{ x } 0 + 74772.08 \text{ x } 0 = 327633.11 \text{ x } z$$

So, **z** = **0**

H_2O Balance

Let x be the mass fraction of H_2O in inlet stream Primary Reformer Flue Gas = 0.1098

Let y be the mass fraction of H_2O in inlet stream of Boiler Flue Gas = 0.1053

Let z be the mass fraction of H_2O in outlet stream Combine Flue Gas = ?

$$252861.03 \text{ x } 0.1098 + 74772.08 \text{ x } 0.1053 = 327633.11 \text{ x } z$$

So, **z** = **0.1088**

Components	**Combine Flue Gas Mass Fraction**	**Wt%**
O_2	0.0190	1.9%
H_2	0	0%
N_2	0.7123	71.23%
CO	0	0%
CO_2	0.1479	14.79%
Ar	0.019	1.9%
CH_4	0	0%
C_2H_6	0	0%
H_2O	0.1088	10.88%
Total	**1**	**100%**
Mass Flow (kg/hr)	**327633.11**	

Table 3-11: Combine Flue Gas Composition

3.4 Energy Balance

Reference Temperature = 25 °C

Balance on Natural and Recycle Gas Mixer

Stream	**M (kg/hr)**	**c_p (kJ/kg °C)**	**T_S (°C)**	**H_{In} (kJ/hr)**
Natural Gas	38240	1.92	38	954470.4
Recycle Gas	996	3.87	117	354615.84
Natural & Recycle Gas	39210	1.99	41.9	1318671.51

$$\Delta H_{In} = \Delta H_{Natural\ Gas} + \Delta H_{Recycle\ Gas}$$

$$\Delta H_{In} = 954470.4 + 354615.84$$

$$\Delta H_{In} = 1309086.24\ kJ/hr$$

$$\Delta H_{Out} = 1318671.51\ kJ/hr$$

So, $\Delta H_{In} \cong \Delta H_{Out}$

Balance on Desulphurized Gas and Process Steam Mixer

Stream	**M (kg/hr)**	**c_p (kJ/kg °C)**	**T_S (°C)**	**H_{In} (kJ/hr)**
Process Steam	103950	2.52	380	92993670
Desulphurized Gas	39210	2.42	413	36816621.6
Process Steam & Desulphurized Gas	143071	2.49	390	130030078.4

$$\Delta H_{In} = \Delta H_{Process\ Steam} + \Delta H_{Desulphurized\ Gas}$$

$$\Delta H_{In} = 92993670 + 36816621.6$$

$$\Delta H_{In} = 129810291.6\ kJ/hr$$

$$\Delta H_{Out} = 130030078\ kJ/hr$$

So, $\Delta H_{In} \cong \Delta H_{Out}$

Balance on Primary Reformer Flue Gas and Boiler Flue Gas Mixer

Stream	**M (kg/hr)**	**c_p (kJ/kg °C)**	**T_S (°C)**	**H_{In} (kJ/hr)**
Primary Reformer Flue Gas	252861	1.3	604	190328499.5
Boiler Flue Gas	74772	1.26	497	44468452.52
Combined Flue Gas	327633	1.19	580	216385214.9

$$\Delta H_{In} = \Delta H_{Primary\ Reformer\ Flue\ Gas} + \Delta H_{Boiler\ Flue\ Gas}$$

$$\Delta H_{In} = 170328499.5 + 44468452.52$$

$$\Delta H_{In} = 214796952\ kJ/hr$$

$$\Delta H_{Out} = 234796952\ kJ/hr$$

$$\text{So, } \Delta H_{In} \cong \Delta H_{Out}$$

Balance on Coil E 201

Stream	**M (kg/hr)**	**c_p (kJ/kg °C)**	**T_S (°C)**	**T_T (°C)**	**H_{In} (kJ/hr)**	**H_{Out} (kJ/hr)**
Process Gas	143071	2.49	390	532	130030078.4	180617122.5
Primary Reformer Flue Gas	252861	1.33	990	840	324534492.7	274088716.6

Duty E 201 = $Q_{E\ 201}$ = 12.06 Gcal/hr = 50410800 kJ/hr

For Process Gas,

$$Q = \Delta H_{Out} - \Delta H_{In}$$

$$Q = 180617122.5 - 130030078.4 = 50587044.18\ kJ/hr$$

$$\text{So, } Q_{E\ 201} \cong Q$$

For Primary reformer Flue Gas,

$$Q = \Delta H_{In} - \Delta H_{Out}$$

$$Q = 1324534492.7 - 274088716.6 = 50445776.07\ kJ/hr$$

$$\text{So, } Q_{E\ 201} \cong Q$$

Balance on Coil E 202

Stream	**M** **(kg/hr)**	**c_p** **(kJ/kg °C)**	**T_S** **(°C)**	**T_T** **(°C)**	**H_{In}** **(kJ/hr)**	**H_{Out}** **(kJ/hr)**
Process Air	45761	1.07	177	492	130030078.4	22866314.09
Primary Reformer Flue Gas	252861	1.33	840	794	274088716.6	256674177.3

Duty E 202 = $Q_{E\ 202}$ = 3.68 Gcal/hr = 15382400 kJ/hr

For Process Air,

$$Q = \Delta H_{Out} - \Delta H_{In}$$

$$Q = 22866314.09 - 130030078.4 = 15423745.05\ kJ/hr$$

$$\text{So, } Q_{E\ 202} \cong Q$$

For Primary Reformer Flue Gas,

$$Q = \Delta H_{In} - \Delta H_{Out}$$

$$Q = 274088716.6 - 256674177.3 = 17414539.34\ kJ/hr$$

$$\text{So, } Q_{E\ 202} \cong Q$$

Balance on Coil E 203

Stream	**M (kg/hr)**	**c_p (kJ/kg °C)**	**T_S (°C)**	**T_T (°C)**	**H_{In} (kJ/hr)**	**H_{Out} (kJ/hr)**
Steam from Steam Drum	143071	2.49	319	348	316092803.5	345622277.2
Primary Reformer Flue Gas	252861	1.33	794	655	256674177.3	207093186

Duty E 203 = $Q_{E\ 203}$ = 10.9 Gcal/hr = 45562000 kJ/hr

For Steam from Steam Drum,

$$Q = \Delta H_{Out} - \Delta H_{In}$$

$$Q = 345622277.2 - 316092803.5 = 29529473.77\ kJ/hr$$

$$\text{So, } Q_{E\ 203} \cong Q$$

For Primary Reformer Flue Gas,

$$Q = \Delta H_{In} - \Delta H_{Out}$$

$$Q = 256674177.3 - 207093186 = 49580991.34\ kJ/hr$$

$$\text{So, } Q_{E\ 203} \cong Q$$

Balance on Coil E 204 A

Stream	**M (kg/hr)**	**c_p (kJ/kg K)**	**T_S (°C)**	**T_T (°C)**	**H_{In} (kJ/hr)**	**H_{Out} (kJ/hr)**
Natural and Recycle Gas	39210	2.2	247	418	17322193.8	33900966
Primary Reformer Flue Gas	252861	1.3	655	604	207093186	190328499.5

Duty E 204 A = $Q_{E\ 204\ A}$ = 3.87 Gcal/hr = 16176600 kJ/hr

For Natural and Recycle Gas,

$$Q = \Delta H_{Out} - \Delta H_{In}$$

$$Q = 33900966 - 17322193.8 = 16578772.2\ kJ/hr$$

$$\text{So, } Q_{E204\,A} \cong Q$$

For Primary Reformer Flue Gas,

$$Q = \Delta H_{In} - \Delta H_{Out}$$

$$Q = 207093186 - 190328499.5 = 16764686.48\ kJ/hr$$

$$\text{So, } Q_{E\ 204\ A} \cong Q$$

Balance on Coil F 202 A

Stream	**M (kg/hr)**	**c_p (kJ/kg °C)**	**T_S (°C)**	**T_T (°C)**	**H_{In} (kJ/hr)**	**H_{Out} (kJ/hr)**
Steam from Steam Drum	255379	4.19	348	375	345622277.2	374513303.5
Boiler Flue Gas	74772	1.26	805	497	73486002.05	44468452.52

Duty F 202 A = $Q_{F\ 202\ A}$ = 6.93 Gcal/hr = 28967400 kJ/hr

For Steam from Steam Drum,

$$Q = \Delta H_{Out} - \Delta H_{In}$$

$$Q = 374513303.5 - 345622277.2 = 28891026.27\ kJ/hr$$

$$\text{So, } Q_{F\ 202\ A} \cong Q$$

For Boiler Flue Gas,

$$Q = \Delta H_{In} - \Delta H_{Out}$$

$$Q = 73486002.05 - 44468452.52 = 29017549.53\ kJ/hr$$

$$\text{So, } Q_{F\ 202\ A} \cong Q$$

Balance on Heat Exchanger E 205

Stream	**M** **(kg/hr)**	**c_p** **(kJ/kg °C)**	**T_S** **(°C)**	**T_T** **(°C)**	**H_{In}** **(kJ/hr)**	**H_{Out}** **(kJ/hr)**
Combined Flue Gas	327633	1.19	580	336	216385214.9	118196881.1

Duty E 205 = $Q_{E\ 205}$ = 22.81 Gcal/hr = 95345800 kJ/hr

For Combined Flue Gas,

$$Q = \Delta H_{In} - \Delta H_{Out}$$

$$Q = 216385214.9 - 118196881.1 = 98188333.7\ kJ/hr$$

$$\text{So, } Q_{E\ 205} \cong Q$$

Balance on Coil E 204 B

Stream	**M** **(kg/hr)**	**c_p** **(kJ/kg °C)**	**T_S** **(°C)**	**T_T** **(°C)**	**H_{In}** **(kJ/hr)**	**H_{Out}** **(kJ/hr)**
Natural and Recycle Gas	39210	2.49	41.9	247	1318671.51	17322193.8
Combined Flue Gas	327633	1.16	336	294	118196881.1	101353268.6

Duty E 204 B = $Q_{E\ 204\ B}$ = 3.82 Gcal/hr = 15967600 kJ/hr

For Natural and Recycle gas,

$$Q = \Delta H_{Out} - \Delta H_{In}$$

$$Q = 17322193.8 - 1318671.51 = 16003522.29\ kJ/hr$$

$$\text{So, } Q_{E\ 204\ B} \cong Q$$

For Combined Flue Gas,

$$Q = \Delta H_{In} - \Delta H_{Out}$$

$$Q = 118196881.1 - 101353268.6 = 16843612.53\ kJ/hr$$

$$\text{So, } Q_{E\ 204\ B} \cong Q$$

Balance on Coil E 206

Stream	**M (kg/hr)**	**c_p (kJ/kg °C)**	**T_S (°C)**	**T_T (°C)**	**H_{In} (kJ/hr)**	**H_{Out} (kJ/hr)**
Boiler Feed Water	257959	4.2	110	136	92091363	119974151.3
Combined Flue Gas	327633	1.16	294	220	101353268.6	73471700.25

Duty E 206 = $Q_{E\ 206}$ = 6.64 Gcal/hr = 27755200 kJ/hr

For Boiler Feed Water,

$$Q = \Delta H_{Out} - \Delta H_{In}$$

$$Q = 119974151.3 - 92091363 = 27882788.31\ kJ/hr$$

So, $Q_{E\ 206} \cong Q$

For Combined Flue Gas,

$$Q = \Delta H_{In} - \Delta H_{Out}$$

$$Q = 101353268.6 - 7347100.25 = 27881568.3\ kJ/hr$$

So, $Q_{E\ 206} \cong Q$

3.5 Data for Pinch Analysis

Following data is extracted after performing energy and material balance on the primary reformer convective section.

Process Streams	**M (kg/hr)**	**c_p (kJ/kg °C)**	**C_P (kW/ °C)**	**T_S (°C)**	**T_T (°C)**	**ΔH (kJ/hr)**	**h (W/m^2 °C)**
Natural & Recycle Gas	39210	2.18	23.72	41	418	8943.5	116.4
Process Air	45761	1.07	13.59	177	492	4279.8	115.0
Process Gas	143071	2.67	106.26	400	532	14026	170.8
Steam from Steam Drum	255379	5.22	370.29	319	375	20736	70.0
BFW	257959	4.15	297.01	110	136	7722	5721
Primary Reformer Flue Gas	252861	1.31	91.93	990	604	35483	200.0
Boiler Flue Gas	74772	1.26	26.17	805	497	8059.6	200.0
Combine Flue Gas	327633	1.18	107.48	580	220	38693	338.3

Table 3-12: Process Streams Data

In this problem five streams have to be heated to desired temperature while there are three streams that require cooling. Note that intermediate temperatures are not required while performing a pinch analysis. Heat transfer coefficient is required to perform area targeting and calculating the area of new designed coils.

The minimum temperature difference (ΔT_{min}) in this problem is 27 °C. At any point in the heat exchanger, temperature difference between the hot and cold streams cannot be less than 27 °C.

3.6 Energy Targeting

Using pinch analysis, it is possible estimate minimum energy requirement before designing the heat exchanger network. For this purpose problem table algorithm is used. Following calculations will be used to target energy for the process.

3.6.1 Shifted Temperatures

Shifted temperatures are calculated by following equations.

For Cold stream, $Shift\,Temperature = Temperature + \,^1/_2\ \Delta T_{min}$

For Hot stream, $Shift\,Temperature = Temperature - \,^1/_2\ \Delta T_{min}$

Process Streams	T_S (ºC)	T_T (ºC)	Stream Type	Shift T_S (ºC)	Shift T_T (ºC)
Natural & Recycle Gas	41	418	Cold	54.5	431.5
Process Air	177	492	Cold	190.5	505.5
Process Gas	400	532	Cold	413.5	545.5
Steam	319	375	Cold	332.5	388.5
BFW	110	136	Cold	123.5	149.5
Primary Reformer Flue Gas	990	604	Hot	976.5	590.5
Boiler Flue Gas	805	497	Hot	791.5	483.5
Combine Flue Gas	580	220	Hot	566.5	206.5

Table 3-13: Process Streams Shifted Temperatures

3.6.2 Problem Table Algorithm

These shifted temperature are then arrange in descending order and any repeated values are omitted. Temperature difference, net heat capacity and heat loads are calculated for each interval using following equations.

$$\Delta T = T_{(i+1)} - T_i$$

$$dH = C_{P\,Net} \times \Delta T$$

$$C_{P\,Net} = \sum C_{P\,Hot} - \sum C_{P\,Cold}$$

If enthalpy change (dH) of the interval is positive, then it has heat surplus and if the value of enthalpy change (dH) is negative then it demands heat. According to second law of thermodynamics, each upper interval can transfer heat to the lower interval as heat is transferred from higher temperature to lower temperature. First, infeasible cascade is determined by supply zero heat in the start and if the interval has surplus heat then it is added to the net heat load and if it demands heat then it is subtracted from the net heat load. If there is any negative value in the infeasible cascade then this heat is supplied in the start to form a feasible cascade. Surplus heat supplied in start is the minimum hot utility demand and heat left at end is the minimum cold utility. In this way, problem table algorithm is used to target the energy before designing the network.

In this problem, infeasible and feasible cascade are same because hot utility demand of the system is zero and minimum cold utility demand is calculated at the end.

Shift Temperature (^oC)	Interval	$T_{(i+1)}$-T_i (^oC)	Cp $_{Net}$ (kW/^oC)	dH (kW)		Infeasible/ Feasible Cascade
976.5					Pinch	0
	1	185	91.92	17006	S	
791.5						17006
	2	201	118.09	23736	S	
590.5						40743
	3	24	26.16	628	S	
566.5						41371
	4	21	128.25	2693	S	
545.5						44064
	5	40	22.1	880	S	
505.5						44944
	6	22	8.42	185	S	
483.5						45130
	7	52	-17.75	-923	D	
431.5						44207
	8	18	-41.47	-746	D	
413.5						43460
	9	25	64.78	1619	S	
388.5						45080
	10	56	-305.5	-17108	D	
332.5						27971
	11	142	64.78	9199	S	
190.5						37170
	12	3	78.36	235	S	
187.5						37405
	13	38	-23.72	-901	D	
149.5						36504
	14	26	-320.73	-8339	D	
123.5						28165
	15	69	-23.72	-1636	D	
54.5						26528

Table 3-14: Problem Table Algorithm for Existing Plant

Pinch temperature is 976.5 ^{o}C. Hot and cold pinch are 990 ^{o}C and 963 ^{o}C, respectively. They are calculated by

$$\text{Hot Pinch} = \text{Pinch Temperature} + \ {}^{1}\!/_{2}\ \Delta T_{min}$$

$$\text{Hot Pinch} = 976.5 + \ {}^{1}\!/_{2} \times 27 = \ 990\ ^{\circ}\text{C}$$

$$\text{Cold Pinch} = \text{Pinch Temperature}\ - \ {}^{1}\!/_{2}\ \Delta T_{min}$$

$$\text{Cold Pinch} = 976.5 - \ {}^{1}\!/_{2} \times 27 = \ 963\ ^{\circ}\text{C}$$

The minimum hot utility requirement is 0 kW while minimum cold utility requirement is 25268 kW. So, it is a threshold problem.

3.6.3 Composite and Grand Composite Curve

Composite Curve is a plot between the actual temperature and heat flow. Calculations are:

For hot composite curve, arrange the supply and target temperatures of hot streams in descending order and calculate the value of heat capacity (C_P) in each interval. The product of the temperature difference and heat capacity will give the value of heat flow. Plot temperatures against heat flow.

Process Stream	**T_S (^{o}C)**	**T_T (^{o}C)**	**C_P (kW/^{o}C)**
Primary Reformer Flue Gas	990	604	91.9
Boiler Flue Gas	805	497	26.2
Combine Flue Gas	580	220	107.5

Table 3-15: Hot Process Streams Data

Actual Temperature (^oC)	**Interval**	**$T_{(i+1)}$-T_i (^oC)**	**$\Sigma C_{P\ Hot}$ ($kW/^oC$)**	**Heat Flow (kW)**	**Total Head Load (kW)**
220					
497	1	277	107.5	29777.5	29777.5
580	2	83	133.7	11097.1	40874.6
604	3	24	26.2	628.8	41503.4
805	4	201	118.1	23738.1	65241.5
990	5	185	91.9	170002	235243

Table 3-16: Hot Composite Curve Streams Data

Similarly, for cold streams

Process Stream	**T_S (^oC)**	**T_T (^oC)**	**C_P ($kW/^oC$)**
Natural & Recycle Gas	41.9	418	23.72
Process Air	177	492	13.6
Process Gas	390	532	106.3
Steam from Steam Drum	319	375	370.3
Boiler Feed Water	110	136	107.5

Table 3-17: Cold Process Streams Data

Actual Temperature (°C)	**Interval**	**$T_{(i+1)}-T_i$ (°C)**	**$\Sigma C_{P\ Cold}$ (kW/°C)**	**Heat Flow (kW)**	**Total Head Load (kW)**
41.9					
110	1	68.1	23.72	1615.3	1615.3
136	2	23	131.22	3018	4633.3
177	3	41	23.72	972.5	5605.8
319	4	142	37.32	5299	10904.8
375	5	56	407.6	22825.6	33730.4
390	6	15	37.32	559.8	34290.2
418	7	28	143.6	4020.8	38311
492	8	74	119.9	8872.6	47183.6
532	9	40	106.3	4252	51435.6

Table 3-18: Cold Composite Curve Stream Data

Now plot the actual temperatures and heat flow for hot and cold streams on the graph to obtain combine composite curve.

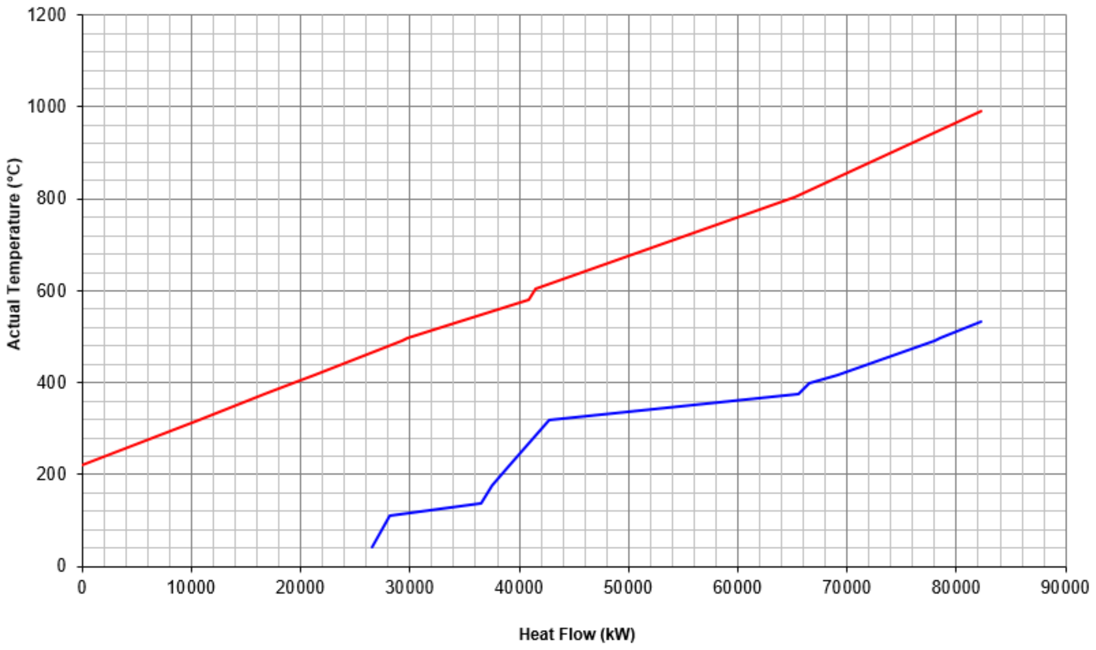


Figure 3.2: Combine Composite Curve

Grand composite curve is a plot between shifted temperature and the net heat flow or the feasible cascade.

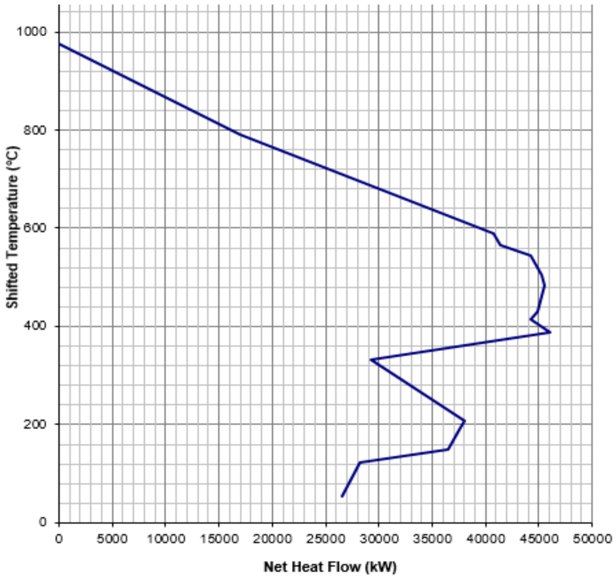


Figure 3.3: Grand Composite Curve

3.6.4 Threshold Temperature

Threshold temperature will only require one utility till a temperature limit known as threshold temperature that is 310 ^{o}C. After this, system requires both utilities. The following table shows the value of hot and cold utilities for different values of minimum temperature difference (ΔT_{min}).

ΔT_{min} (°C)	**Hot Pinch (°C)**	**Cold Pinch (°C)**	**Hot Utility (kW)**	**Cold Utility (kW)**
310	629	319	1670.6942	28198.674
295.75	990	694.25	0	26527.98
281.5	990	708.5	0	26527.98
267.25	990	722.75	0	26527.98
253	990	737	0	26527.98
238.75	990	751.25	0	26527.98
224.5	990	765.5	0	26527.98
210.25	990	779.75	0	26527.98
196	990	794	0	26527.98
181.75	990	808.25	0	26527.98
167.5	990	822.5	0	26527.98
153.25	990	836.75	0	26527.98
139	990	851	0	26527.98
124.75	990	865.25	0	26527.98
110.5	990	879.5	0	26527.98
96.25	990	893.75	0	26527.98
82	990	908	0	26527.98
67.75	990	922.25	0	26527.98
53.5	990	936.5	0	26527.98
39.25	990	950.75	0	26527.98
25	990	965	0	26527.98

Table 3-19: Threshold Temperature

Chapter 4
Aspen Energy Analyzer

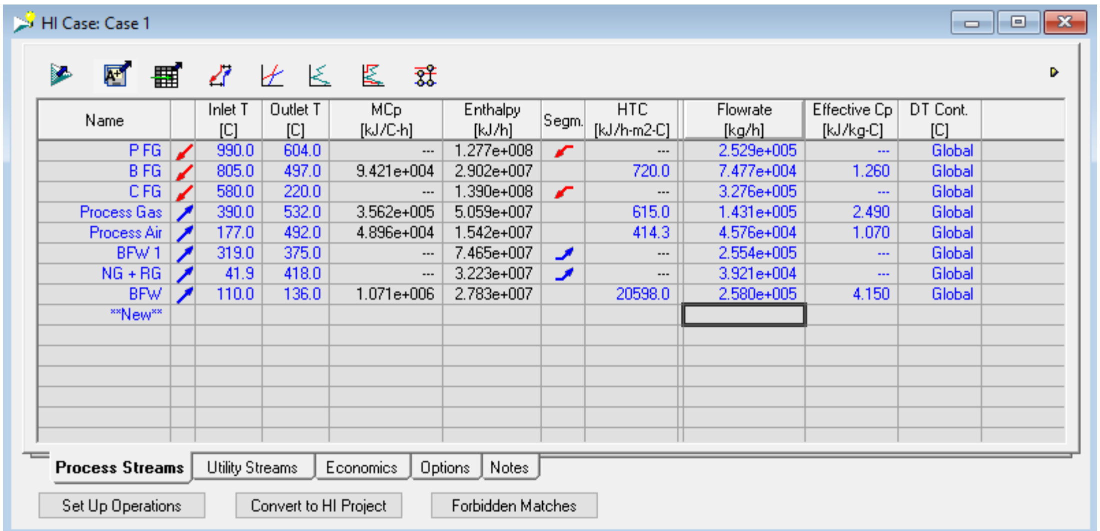
Heat Exchanger Network is designed using Aspen Energy Analyzer. Details to design existing heat exchanger network on Aspen Energy Analyzer is discussed in following section.

4.1 Data Entering

4.1.1 Defining Process Streams

Following process streams data will be entered into aspen energy analyser.

- Name of the stream
- Inlet Temperature
- Outlet Temperature
- Mass Flowrate
- Specific Heat Capacity
- Heat Transfer Coefficient

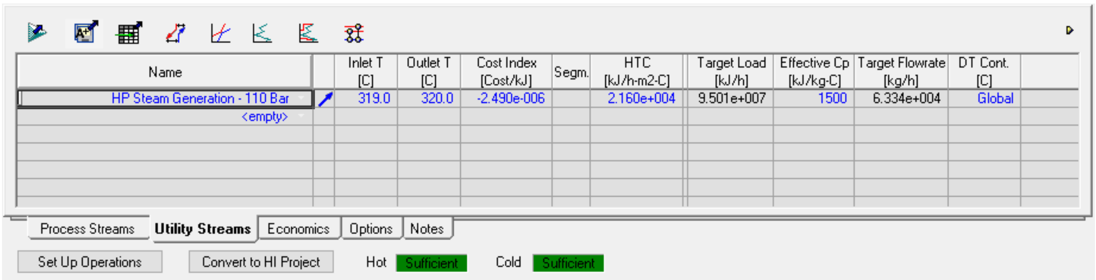


Name		Inlet T [C]	Outlet T [C]	MCp [kJ/C-h]	Enthalpy [kJ/h]	Segm.	HTC [kJ/h-m2-C]	Flowrate [kg/h]	Effective Cp [kJ/kg-C]	DT Cont. [C]
P FG	↙	990.0	604.0	---	1.277e+008	↙	---	2.529e+005	---	Global
B FG	↙	805.0	497.0	9.421e+004	2.902e+007		720.0	7.477e+004	1.260	Global
C FG	↙	580.0	220.0	---	1.390e+008	↙	---	3.276e+005	---	Global
Process Gas	↗	390.0	532.0	3.562e+005	5.059e+007		615.0	1.431e+005	2.490	Global
Process Air	↗	177.0	492.0	4.896e+004	1.542e+007		414.3	4.576e+004	1.070	Global
BFW 1	↗	319.0	375.0	---	7.465e+007	↗	---	2.554e+005	---	Global
NG + RG	↗	41.9	418.0	---	3.223e+007	↗	---	3.921e+004	---	Global
BFW	↗	110.0	136.0	1.071e+006	2.783e+007		20598.0	2.580e+005	4.150	Global
New										

Figure 4.1 Defining Reference Streams

4.1.2 Defining Utility Stream

After defining process streams, utility is defined for the system. Only cold utility is required here. The cold utility is High Pressure Steam Generation of 110 Bars at 320 ^{o}C.



Name		Inlet T [C]	Outlet T [C]	Cost Index [Cost/kJ]	Segm.	HTC [kJ/h-m2-C]	Target Load [kJ/h]	Effective Cp [kJ/kg-C]	Target Flowrate [kg/h]	DT Cont. [C]
HP Steam Generation - 110 Bar		319.0	320.0	-2.490e-006		2.160e+004	9.501e+007	1500	6.334e+004	Global
<empty>										

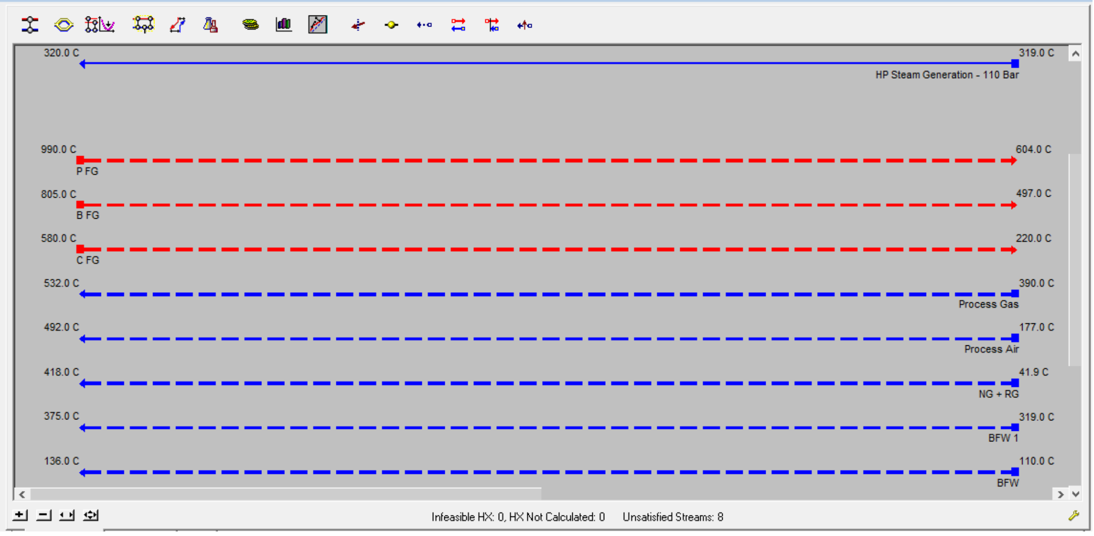
Figure 4.2: Defining Utility Stream

4.2 Existing Heat Exchanger Network

Existing heat exchanger network design is as followed:

4.2.1 Unsatisfied Grid Diagram

In this problem, there are three hot streams while five cold streams. The unsatisfied streams are represented by the dotted line. When the stream is satisfied, it is represented by solid line.

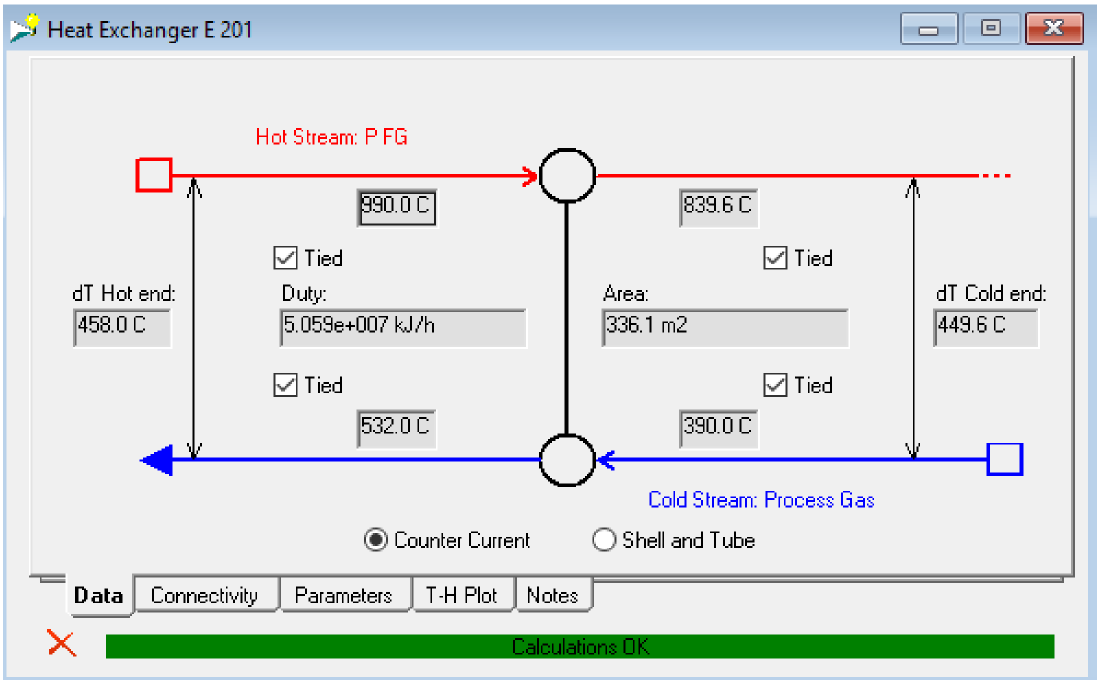


Infeasible HX: 0, HX Not Calculated: 0   Unsatisfied Streams: 8

Figure 4.3: Unsatisfied Grid Diagram

4.2.2 Coil E 201

Now heat exchanger between Primary Reformer Flue Gas and Process Gas placed. Heat exchanger is defined by providing supply or target temperatures of the streams exchanging heat.



Heat Exchanger E 201

Hot Stream: P FG

990.0 C

839.6 C

☑ Tied

☑ Tied

dT Hot end: 458.0 C

Duty: 5.059e+007 kJ/h

Area: 336.1 m2

dT Cold end: 449.6 C

☑ Tied

☑ Tied

532.0 C

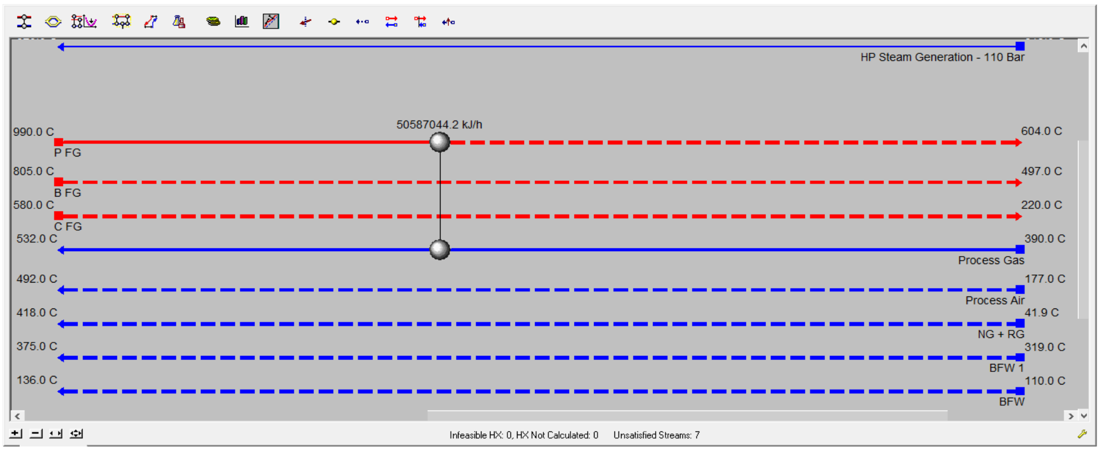
390.0 C

Cold Stream: Process Gas

● Counter Current ○ Shell and Tube

Calculations OK

Figure 4.4: Coil E 201

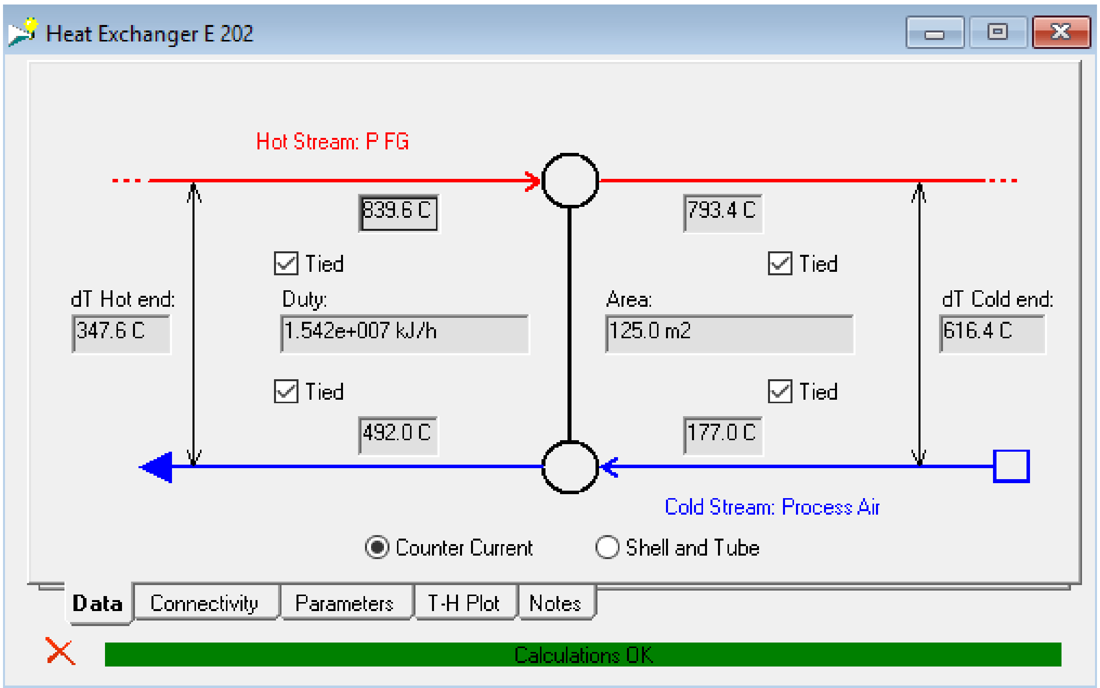


Infeasible HX: 0, HX Not Calculated: 0 Unsatisfied Streams: 7

Figure 4.5: Gird Diagram with E 201

4.2.3 Coil E 202

Now heat exchanger between Primary Reformer Flue Gas and Process Air placed. Heat exchanger is defined by providing supply or target temperatures of the streams exchanging heat.



Heat Exchanger E 202

Hot Stream: P FG

839.6 C

793.4 C

☑ Tied

☑ Tied

dT Hot end: 347.6 C

Duty: 1.542e+007 kJ/h

Area: 125.0 m2

dT Cold end: 616.4 C

☑ Tied

☑ Tied

492.0 C

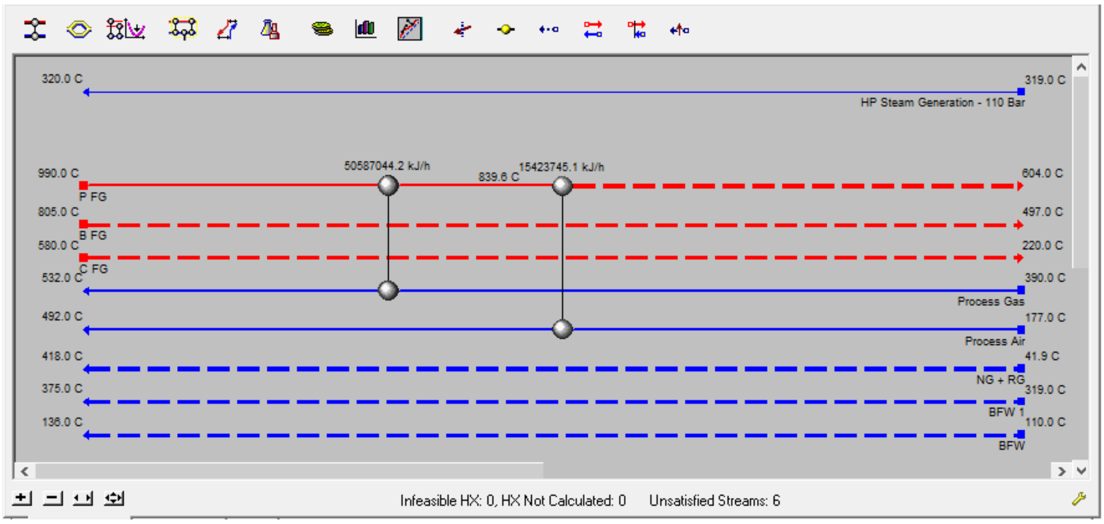
177.0 C

Cold Stream: Process Air

● Counter Current ○ Shell and Tube

Calculations OK

Figure 4.6: Coil E 202

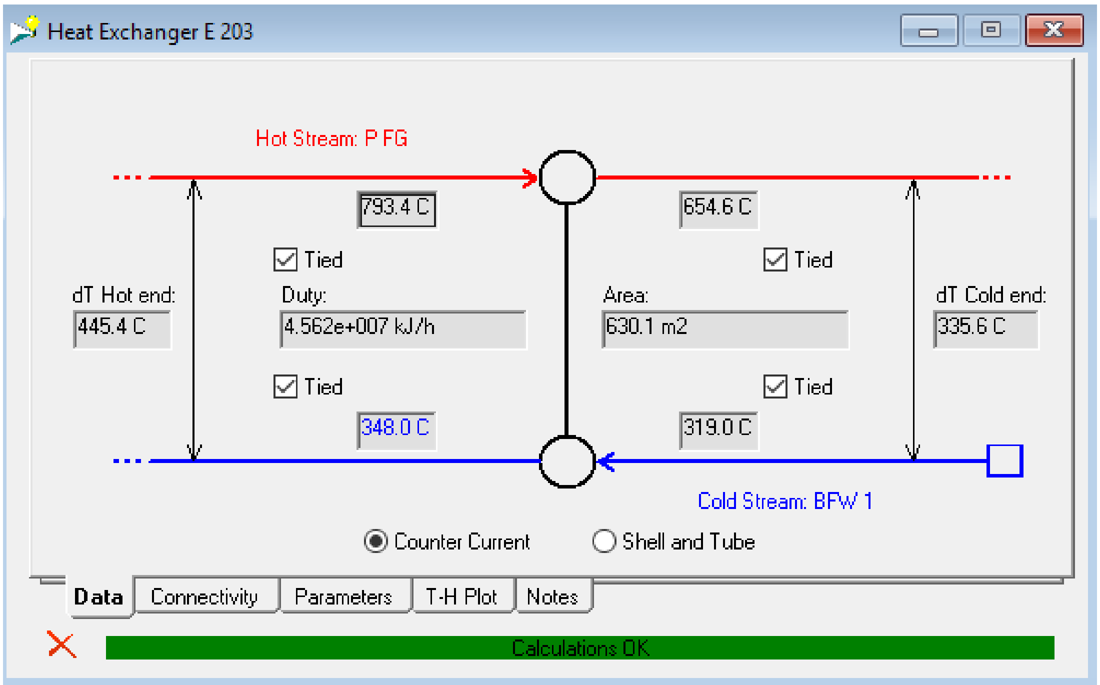


Infeasible HX: 0, HX Not Calculated: 0 Unsatisfied Streams: 6

Figure 4.7: Gird Diagram with E 202

4.2.4 Coil E 203

Now heat exchanger between Primary Reformer Flue Gas and Steam from Steam Drum placed. Heat exchanger is defined by providing supply or target temperatures of the streams exchanging heat.



Heat Exchanger E 203

Hot Stream: P FG

793.4 C

654.6 C

☑ Tied

☑ Tied

dT Hot end: 445.4 C

Duty: 4.562e+007 kJ/h

Area: 630.1 m2

dT Cold end: 335.6 C

☑ Tied

☑ Tied

348.0 C

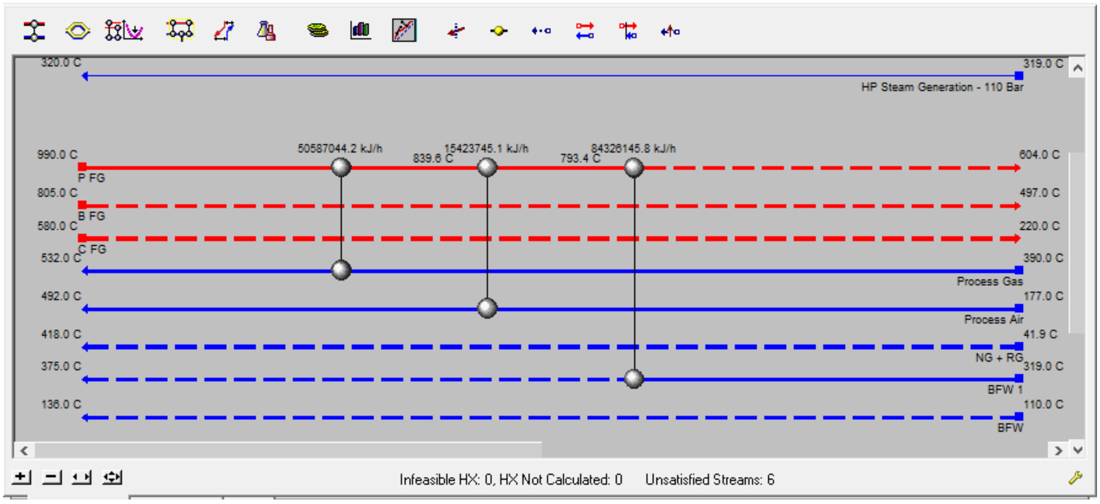
319.0 C

Cold Stream: BFW 1

● Counter Current ○ Shell and Tube

Calculations OK

Figure 4.8: Coil E 203

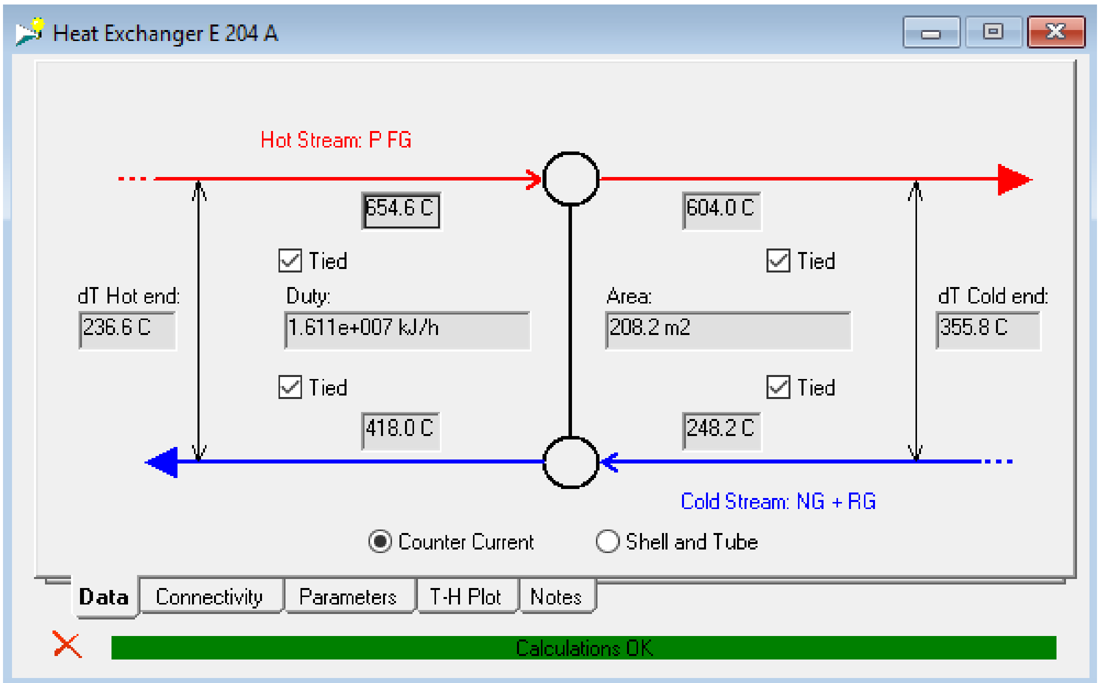


Infeasible HX: 0, HX Not Calculated: 0 Unsatisfied Streams: 6

Figure 4.9: Gird Diagram with E 203

4.2.5 Coil E 204 A

Now heat exchanger between Primary Reformer Flue Gas and Natural & Recycle Gas is placed. Heat exchanger is defined by providing supply or target temperatures of the streams exchanging heat.



Heat Exchanger E 204 A

Hot Stream: P FG

654.6 C

604.0 C

☑ Tied

☑ Tied

dT Hot end: 236.6 C

Duty: 1.611e+007 kJ/h

Area: 208.2 m2

dT Cold end: 355.8 C

☑ Tied

☑ Tied

418.0 C

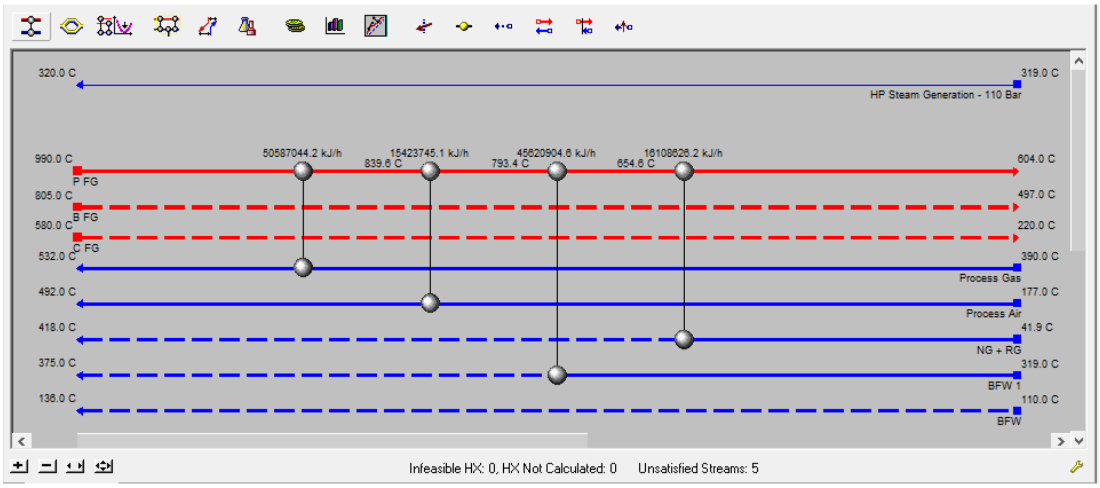
248.2 C

Cold Stream: NG + RG

● Counter Current ○ Shell and Tube

Calculations OK

Figure 4.10: Coil E 204 A

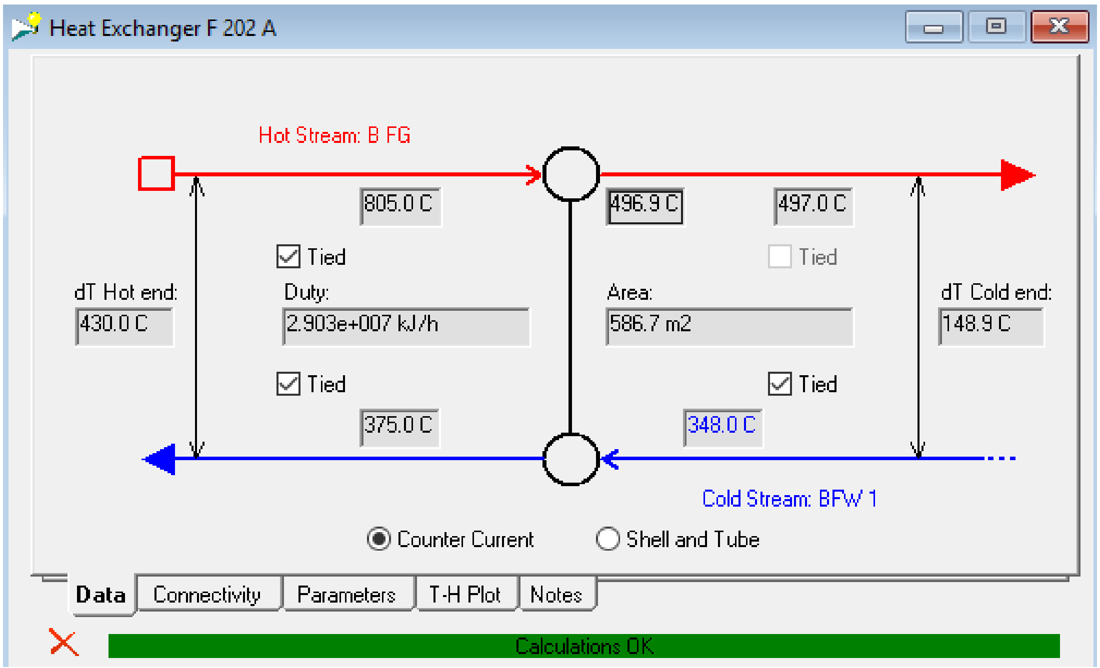


Infeasible HX: 0, HX Not Calculated: 0   Unsatisfied Streams: 5

Figure 4.11: Gird Diagram with E 204 A

4.2.6 Coil F 202 A

Now heat exchanger between Boiler Flue gas and Steam from Steam Drum placed. Heat exchanger is defined by providing supply or target temperatures of the streams exchanging heat.



Heat Exchanger F 202 A

Hot Stream: B FG

805.0 C

496.9 C

497.0 C

☑ Tied

☐ Tied

dT Hot end: 430.0 C

Duty: 2.903e+007 kJ/h

Area: 586.7 m2

dT Cold end: 148.9 C

☑ Tied

☑ Tied

375.0 C

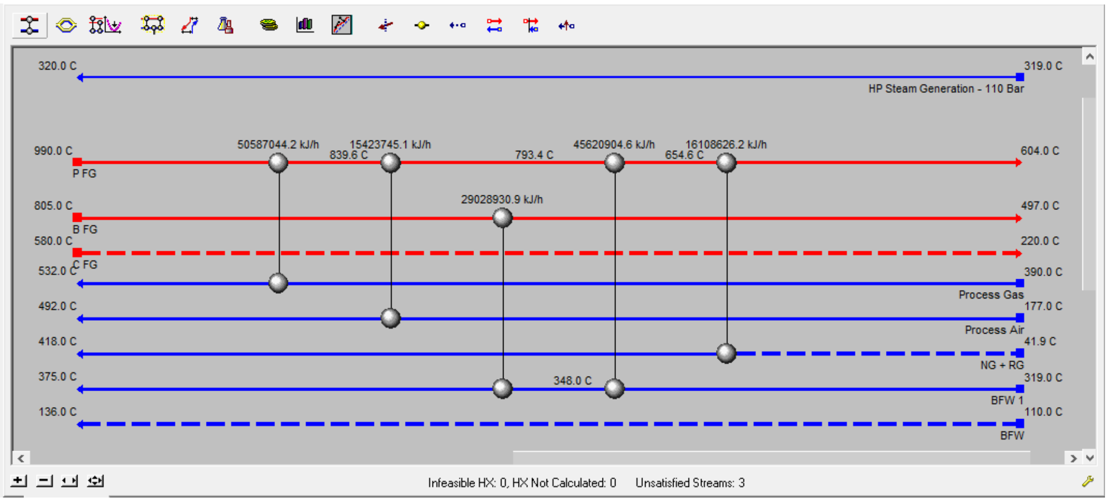
348.0 C

Cold Stream: BFW 1

● Counter Current ○ Shell and Tube

Calculations OK

Figure 4.12: Coil F 202 A

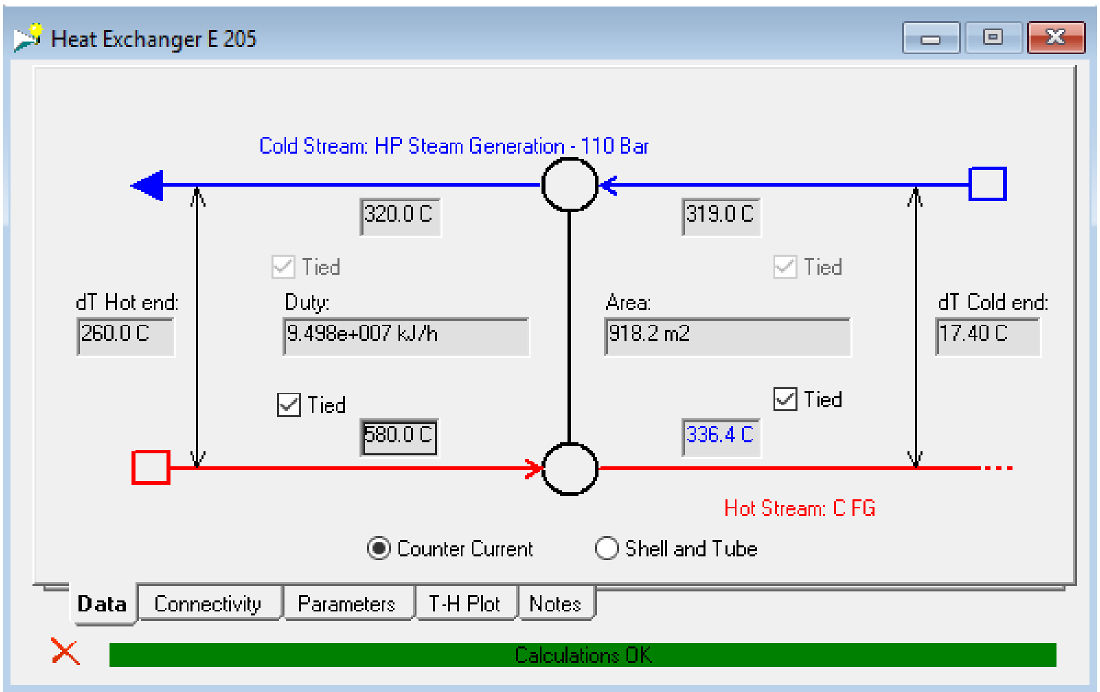


Infeasible HX: 0, HX Not Calculated: 0 Unsatisfied Streams: 3

Figure 4.13: Gird Diagram with F 202 A

4.2.7 Coil E 205

Now high pressure steam generator is added to be heated by combine flue gas.



Heat Exchanger E 205

Cold Stream: HP Steam Generation - 110 Bar

320.0 C

319.0 C

☑ Tied

☑ Tied

dT Hot end: 260.0 C

Duty: 9.498e+007 kJ/h

Area: 918.2 m2

dT Cold end: 17.40 C

☑ Tied

☑ Tied

580.0 C

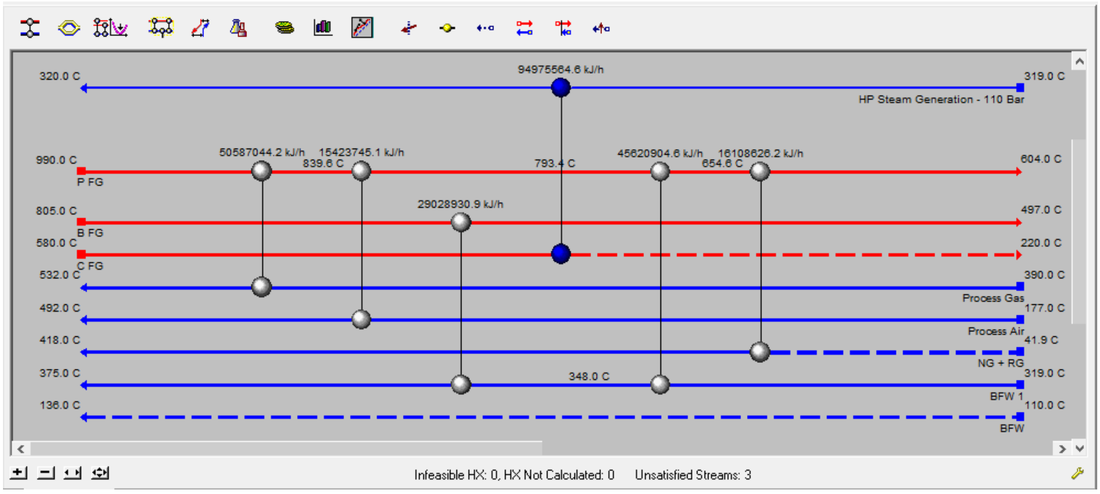
336.4 C

Hot Stream: C FG

● Counter Current ○ Shell and Tube

Calculations OK

Figure 4.14: Coil E 205

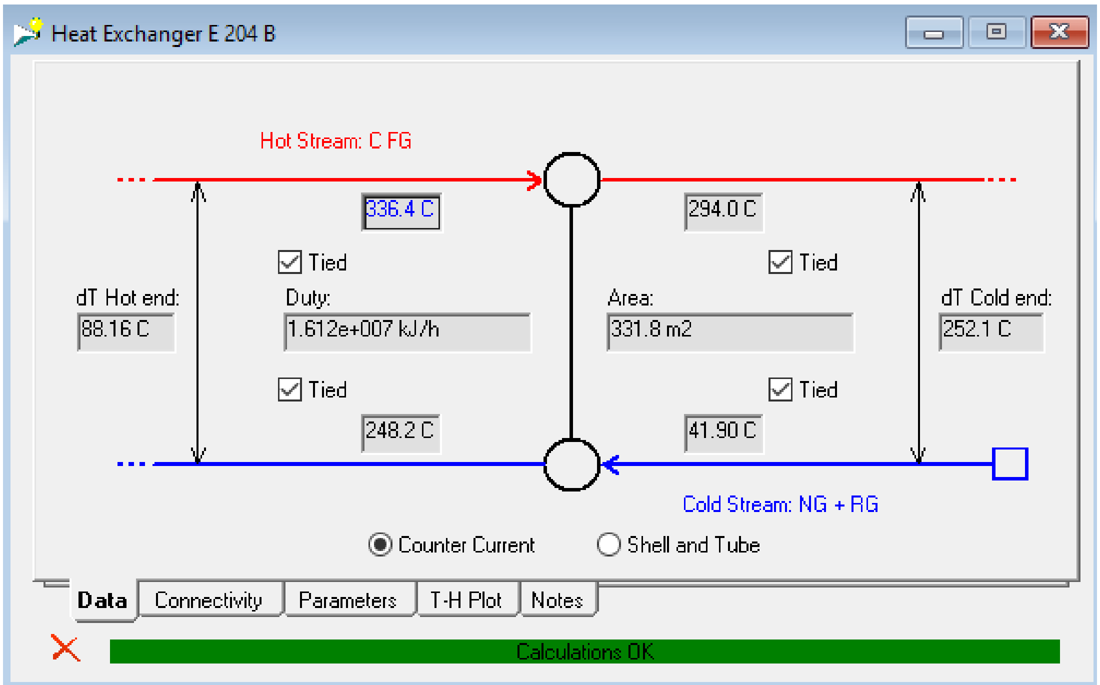


Infeasible HX: 0, HX Not Calculated: 0 Unsatisfied Streams: 3

Figure 4.15: Gird Diagram with E 205

4.2.8 Coil E 204 B

Now heat exchanger between Combine Flue Gas and Boiler Feed Water is placed. Heat exchanger is defined by providing supply or target temperatures of the streams exchanging heat.



Heat Exchanger E 204 B

Hot Stream: C FG

336.4 C

294.0 C

☑ Tied

☑ Tied

dT Hot end: 88.16 C

Duty: 1.612e+007 kJ/h

Area: 331.8 m2

dT Cold end: 252.1 C

☑ Tied

☑ Tied

248.2 C

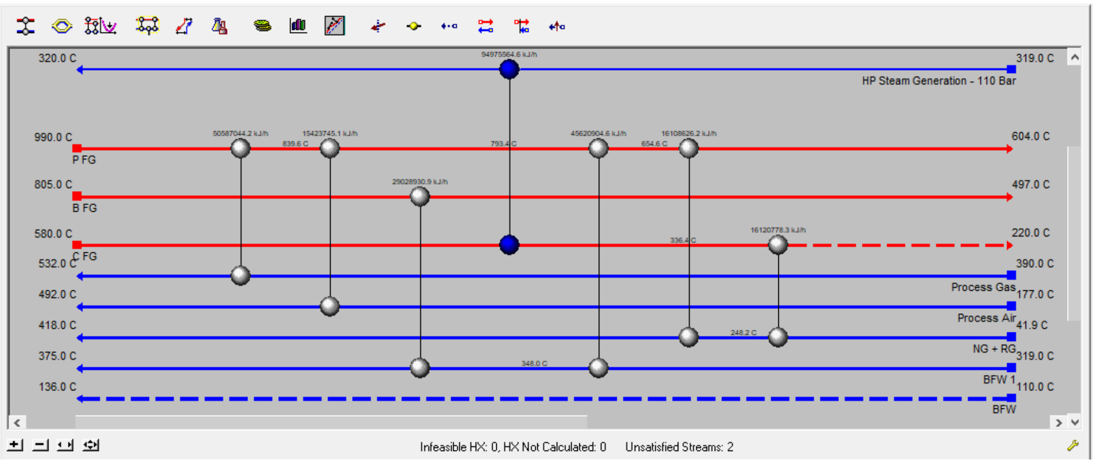
41.90 C

Cold Stream: NG + RG

● Counter Current ○ Shell and Tube

Calculations OK

Figure 4.16: Coil E 204 B

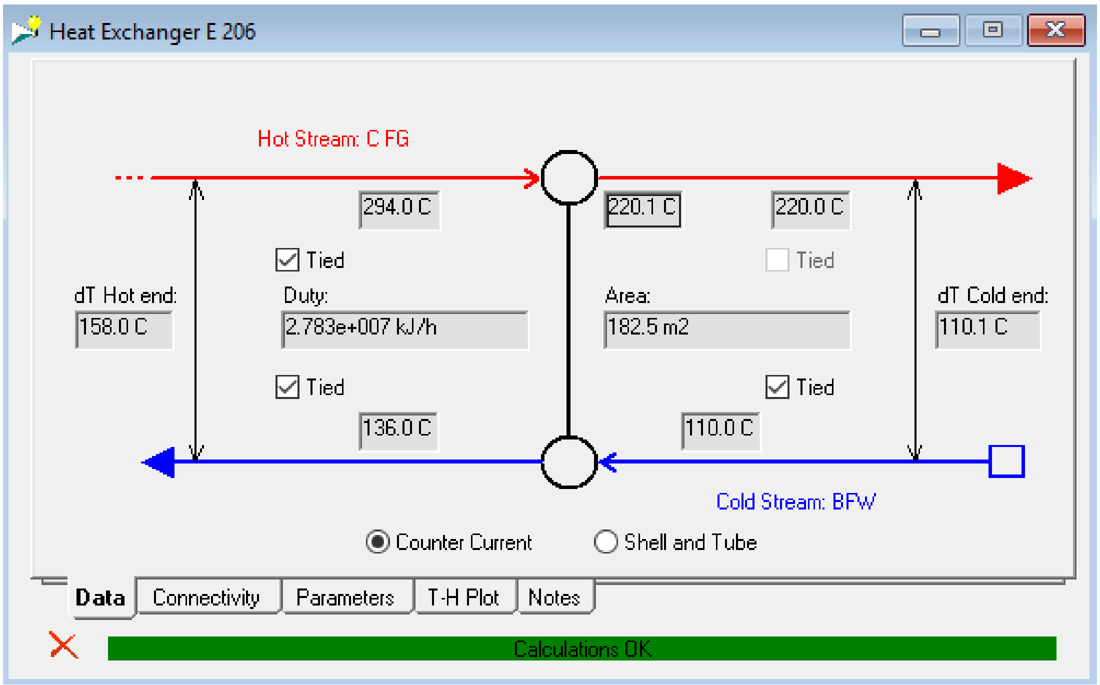


Infeasible HX: 0, HX Not Calculated: 0 Unsatisfied Streams: 2

Figure 4.17: Gird Diagram with E 204 B

4.2.9 Coil E 206

Now heat exchanger between Combine Flue Gas and Boiler Feed Water is placed. Heat exchanger is defined by providing supply or target temperatures of the streams exchanging heat.



Heat Exchanger E 206

Hot Stream: C FG

294.0 C

220.1 C

220.0 C

☑ Tied

☐ Tied

dT Hot end: 158.0 C

Duty: 2.783e+007 kJ/h

Area: 182.5 m2

dT Cold end: 110.1 C

☑ Tied

☑ Tied

136.0 C

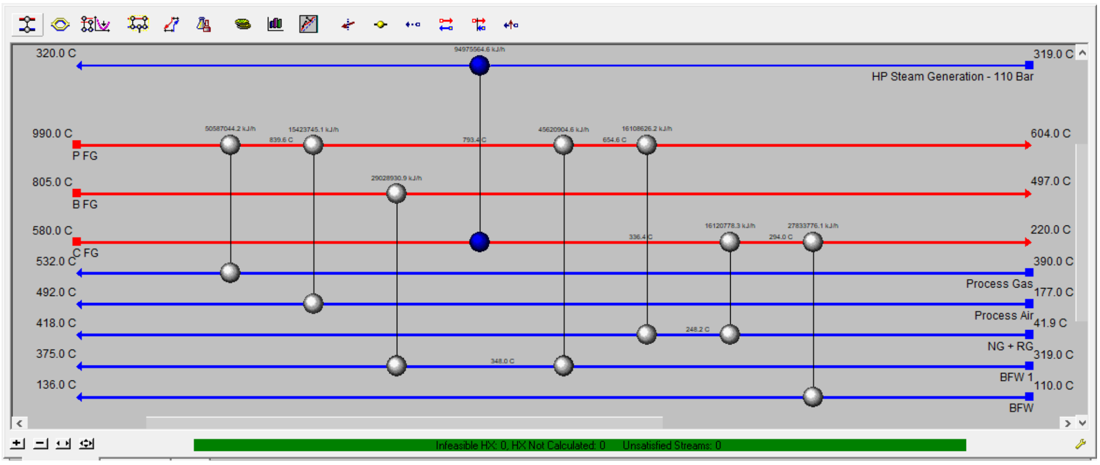
110.0 C

Cold Stream: BFW

● Counter Current ○ Shell and Tube

Calculations OK

Figure 4.18: Coil E 206



Infeasible HX: 0, HX Not Calculated: 0 Unsatisfied Streams: 0

Figure 4.19: Gird Diagram with E 206

This is the completely satisfied heat exchanger network diagram for existing process.

4.2.9 Existing Heat Exchanger Network Grid Diagram

Following is the completely satisfied Grid Diagram for existing heat exchanger network:

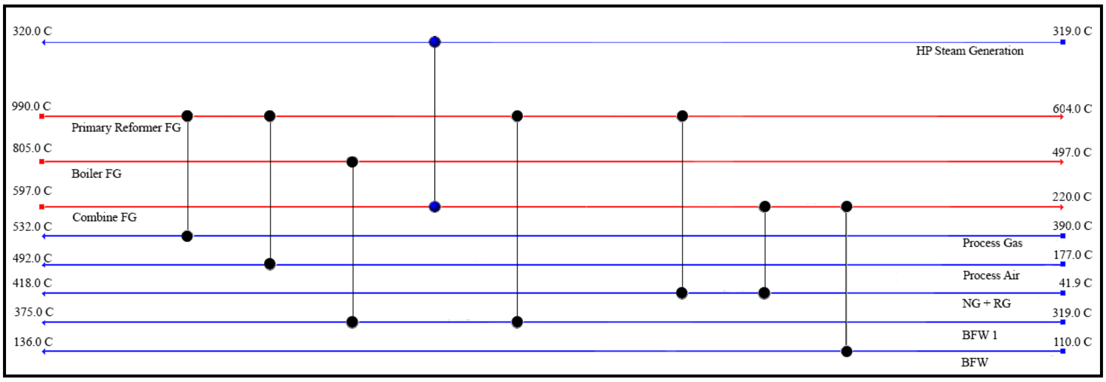
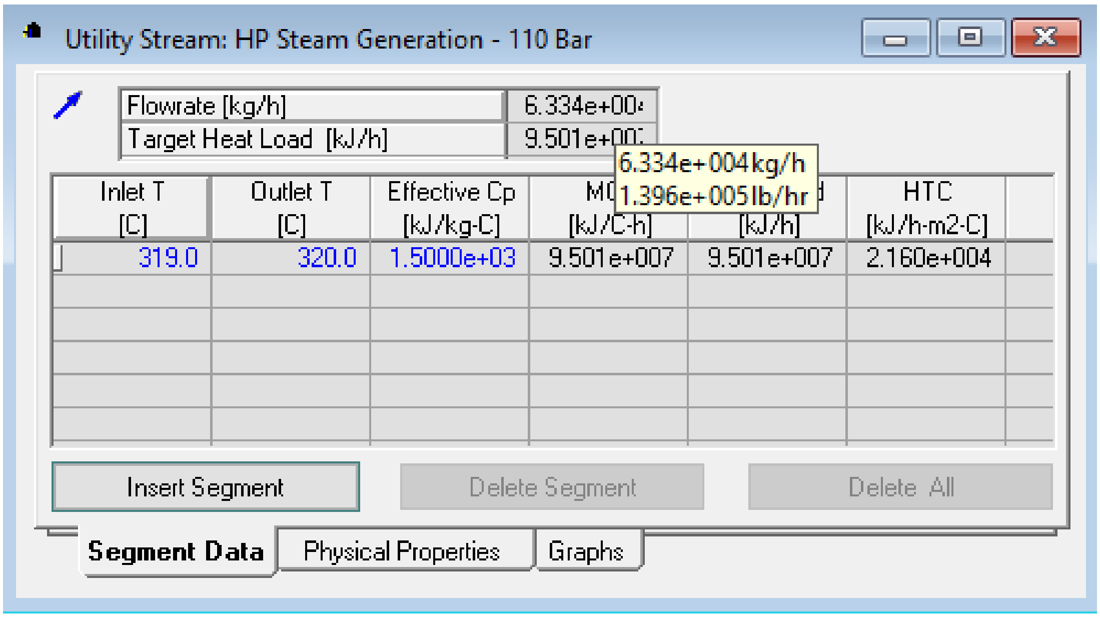


Figure 4.20: Existing Heat Exchanger Network Grid Diagram

4.2.10 High Pressure Steam Production in Coil E 205

Currently the production High Pressure Steam is 63340 kilograms per hour.



Flowrate [kg/h]	6.334e+00
Target Heat Load [kJ/h]	9.501e+00

6.334e+004 kg/h
1.396e+005 lb/hr

Inlet T [C]	Outlet T [C]	Effective Cp [kJ/kg-C]	M [kJ/C-h]	d [kJ/h]	HTC [kJ/h-m2-C]
319.0	320.0	1.5000e+03	9.501e+007	9.501e+007	2.160e+004

Figure 4.21: High Pressure Steam Mass Flowrate

4.3 Analysis of the Existing Process

Detail analysis of the existing process is performed to identify different possible sources of heat recovery. For this, soft and fix data has to be identified. In this process, outlet temperature of coil E 205 has to be kept constant that is 336 ^{o}C as it directly affect the pressure of steam being generated. Hence it is a fix data. On the other hand, stack temperature of the flue gas is soft data and can easily be changed by using Plus/Minus Principle. However, it cannot be reduce to less than sulphur dioxide dew point i.e. 120 ^{o}C. If the flue gas is released near sulphur dioxide dew point then it will not only be corrosive for the coils and shell as they are made up of carbon steel but flue gas will also cause acid rain. Hence it is harmful for both, the plant and the environment.

While reducing the temperature of the flue gas, following conditions must be considered.

- No temperature crossover occur between the streams.
- Approach temperature condition should not be violated in any heat exchangers.

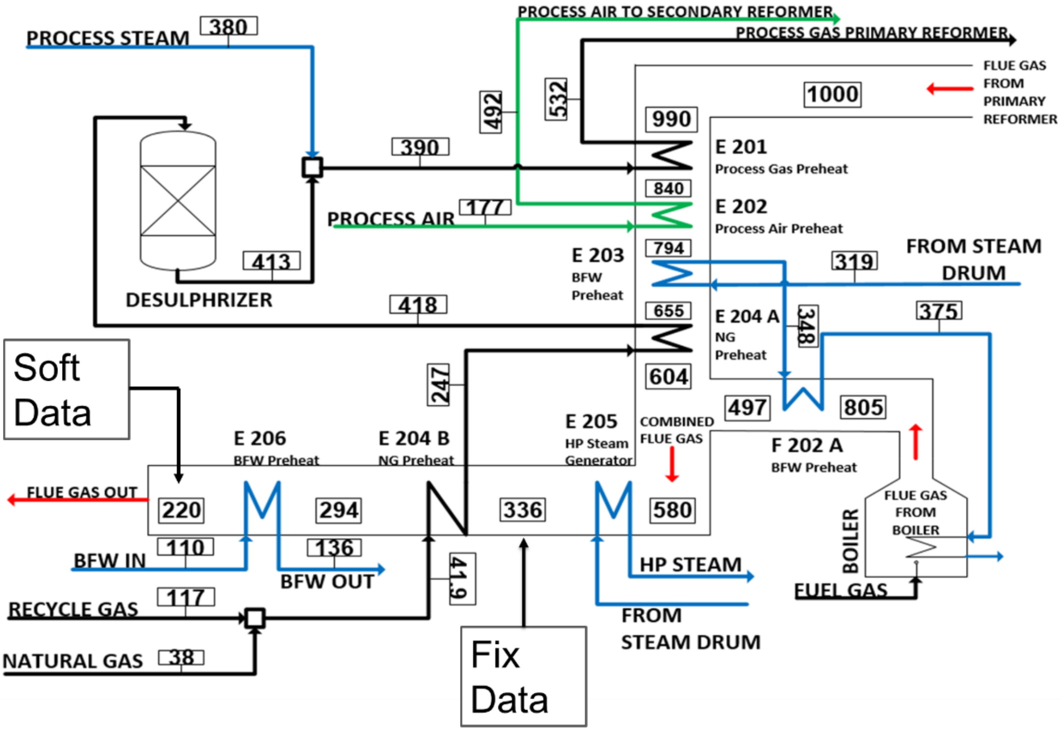
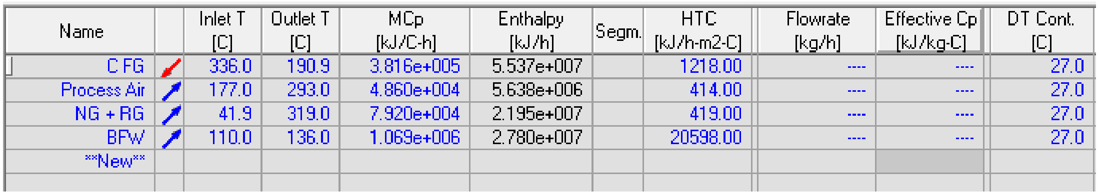


Figure 4.22: Process Flow Diagram for Existing Process

Coils being heated by Combine Flue Gas effects the coils heated by Primary Reformer Flue Gas. Number of new cases are designed for the Combine Flue Gas portion using Aspen Energy Analyzer and most suitable case is then selected.

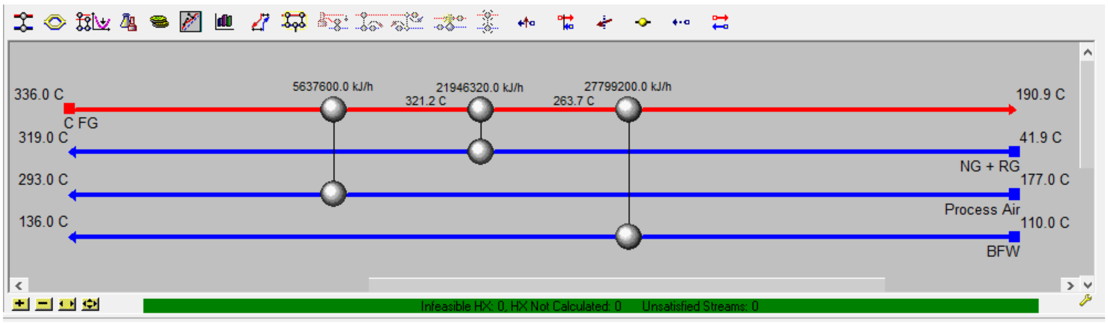
4.3.1 Case 1

In case 1, new pre heater is added to heat Process Air using Combine Flue Gas. Temperature of Combine Flue Gas was reduced from 336 ^{o}C to 190.7 ^{o}C while Process Air is heated from 177 ^{o}C to 293 ^{o}C and Natural and Recycle Gas is heated to 319 ^{o}C.



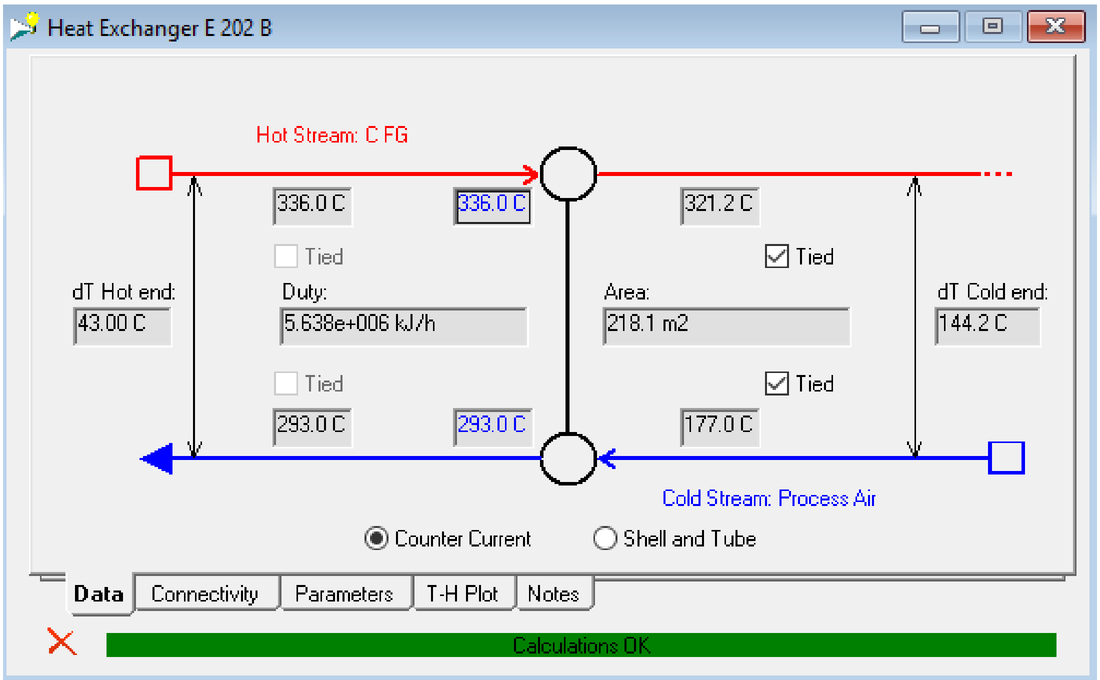
Name		Inlet T [C]	Outlet T [C]	MCp [kJ/C-h]	Enthalpy [kJ/h]	Segm.	HTC [kJ/h-m2-C]	Flowrate [kg/h]	Effective Cp [kJ/kg-C]	DT Cont. [C]
C FG		336.0	190.9	3.816e+005	5.537e+007		1218.00	----	----	27.0
Process Air		177.0	293.0	4.860e+004	5.638e+006		414.00	----	----	27.0
NG + RG		41.9	319.0	7.920e+004	2.195e+007		419.00	----	----	27.0
BFW		110.0	136.0	1.069e+006	2.780e+007		20598.00	----	----	27.0
New										

Figure 4.23: Case 1-Streams Data



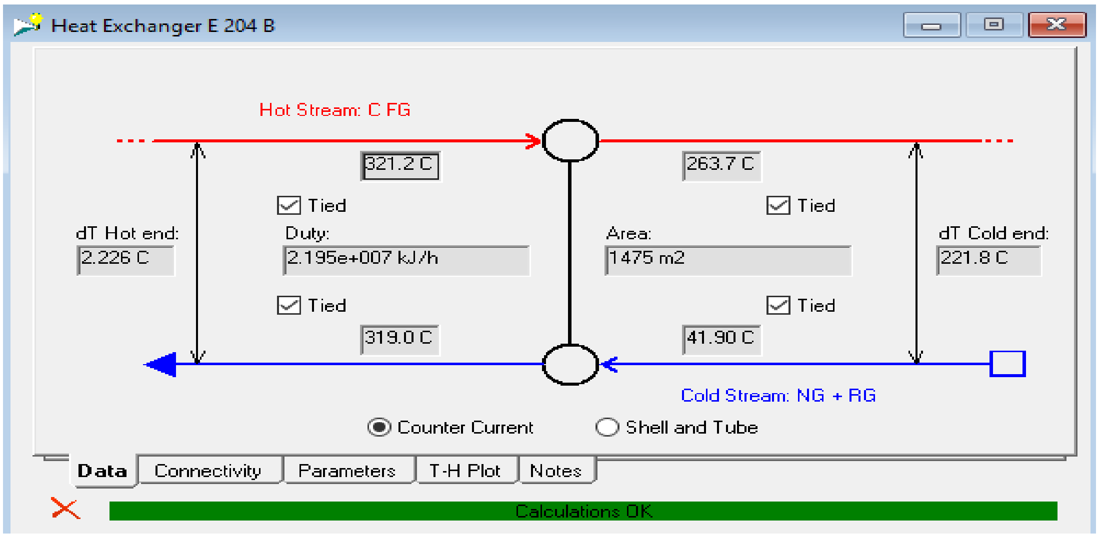
Infeasible HX: 0, HX Not Calculated: 0 Unsatisfied Streams: 0

Figure 4.24: Case 1-Grid Diagram



336.0 C
336.0 C
321.2 C
☐ Tied
☑ Tied
dT Hot end: 43.00 C
Duty: 5.638e+006 kJ/h
Area: 218.1 m2
dT Cold end: 144.2 C
☐ Tied
☑ Tied
293.0 C
293.0 C
177.0 C
● Counter Current ○ Shell and Tube
Calculations OK

Figure 4.25: Case 1-Coil E 202 B


Heat Exchanger E 204 B

Hot Stream: C FG

321.2 C

263.7 C

☑ Tied ☑ Tied

dT Hot end: 2.226 C

Duty: 2.195e+007 kJ/h

Area: 1475 m2

dT Cold end: 221.8 C

☑ Tied ☑ Tied

319.0 C

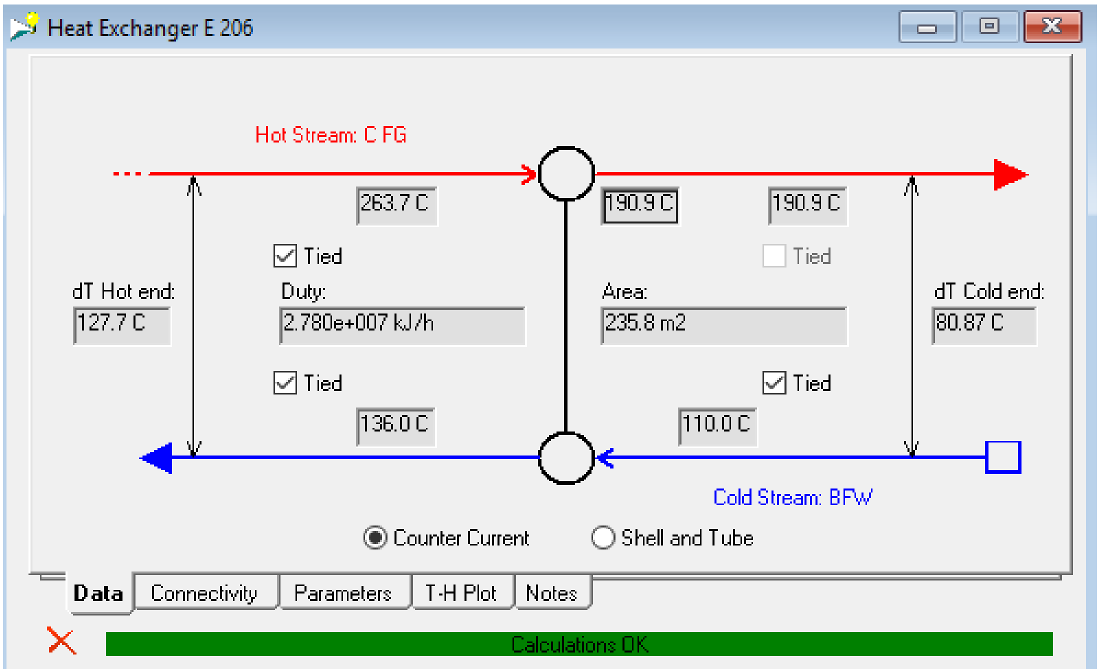
41.90 C

Cold Stream: NG + RG

● Counter Current ○ Shell and Tube

Calculations OK

Figure 4.26: Case 1-Coil E 204 B


Heat Exchanger E 206

Hot Stream: C FG

263.7 C

190.9 C

190.9 C

☑ Tied ☐ Tied

dT Hot end: 127.7 C

Duty: 2.780e+007 kJ/h

Area: 235.8 m2

dT Cold end: 80.87 C

☑ Tied ☑ Tied

136.0 C

110.0 C

Cold Stream: BFW

● Counter Current ○ Shell and Tube

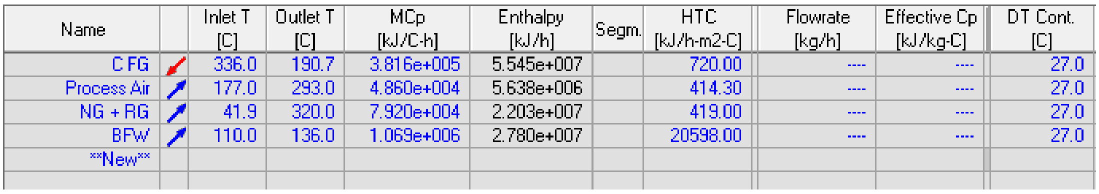
Calculations OK

Figure 4.27: Case 1-Coil E 206

Although temperature of the Flue Gas is significantly reduced, the areas of the exchangers came to be quite large making the case impractical to apply.

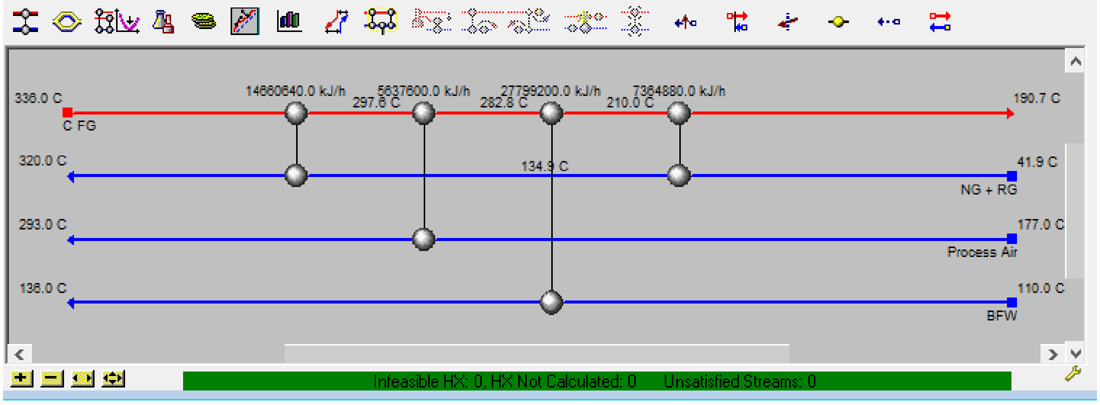
4.3.2 Case 2

In case 2, new pre heater to heat Process Air and heating Natural and Recycle Gas in two parts using Combine Flue Gas. Temperature of Combine Flue Gas was reduced from 336 °C to 190.7 °C while Process Air is heated from 177 °C to 293 °C and Natural and Recycle Gas is heated to 319 °C.



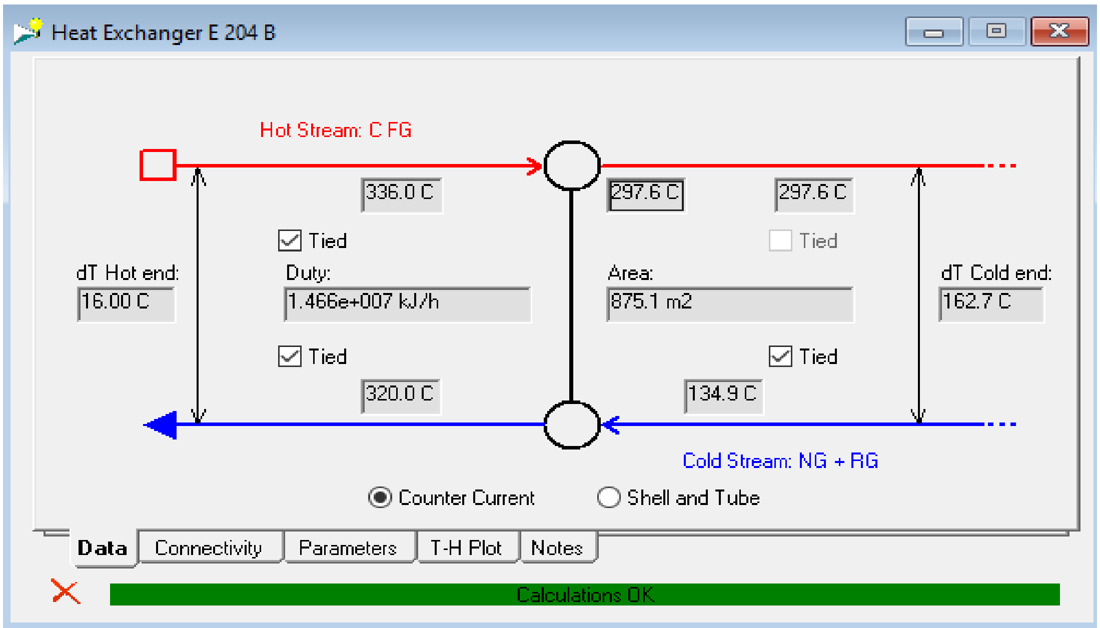
Name		Inlet T [C]	Outlet T [C]	MCp [kJ/C-h]	Enthalpy [kJ/h]	Segm.	HTC [kJ/h-m2-C]	Flowrate [kg/h]	Effective Cp [kJ/kg-C]	DT Cont. [C]
C FG		336.0	190.7	3.816e+005	5.545e+007		720.00	----	----	27.0
Process Air		177.0	293.0	4.860e+004	5.638e+006		414.30	----	----	27.0
NG + RG		41.9	320.0	7.920e+004	2.203e+007		419.00	----	----	27.0
BFW		110.0	136.0	1.069e+006	2.780e+007		20598.00	----	----	27.0
New										

Figure 4.28: Case 2-Streams Data



Infeasible HX: 0, HX Not Calculated: 0 Unsatisfied Streams: 0

Figure 4.29: Case 2-Grid Diagram



Heat Exchanger E 204 B

Hot Stream: C FG

336.0 C

297.6 C

297.6 C

☑ Tied

☐ Tied

dT Hot end: 16.00 C

Duty: 1.466e+007 kJ/h

Area: 875.1 m2

dT Cold end: 162.7 C

☑ Tied

☑ Tied

320.0 C

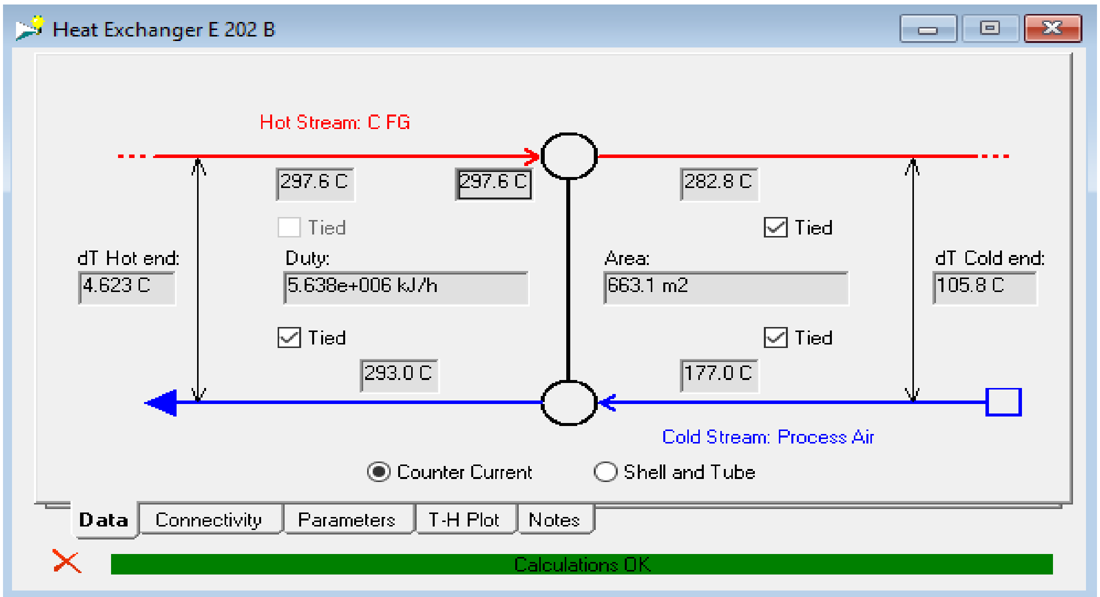
134.9 C

Cold Stream: NG + RG

● Counter Current ○ Shell and Tube

Calculations OK

Figure 4.30: Case 2-E 204 B



Hot Stream: C FG

297.6 C

297.6 C

282.8 C

☐ Tied

☑ Tied

dT Hot end: 4.623 C

Duty: 5.638e+006 kJ/h

Area: 663.1 m2

dT Cold end: 105.8 C

☑ Tied

☑ Tied

293.0 C

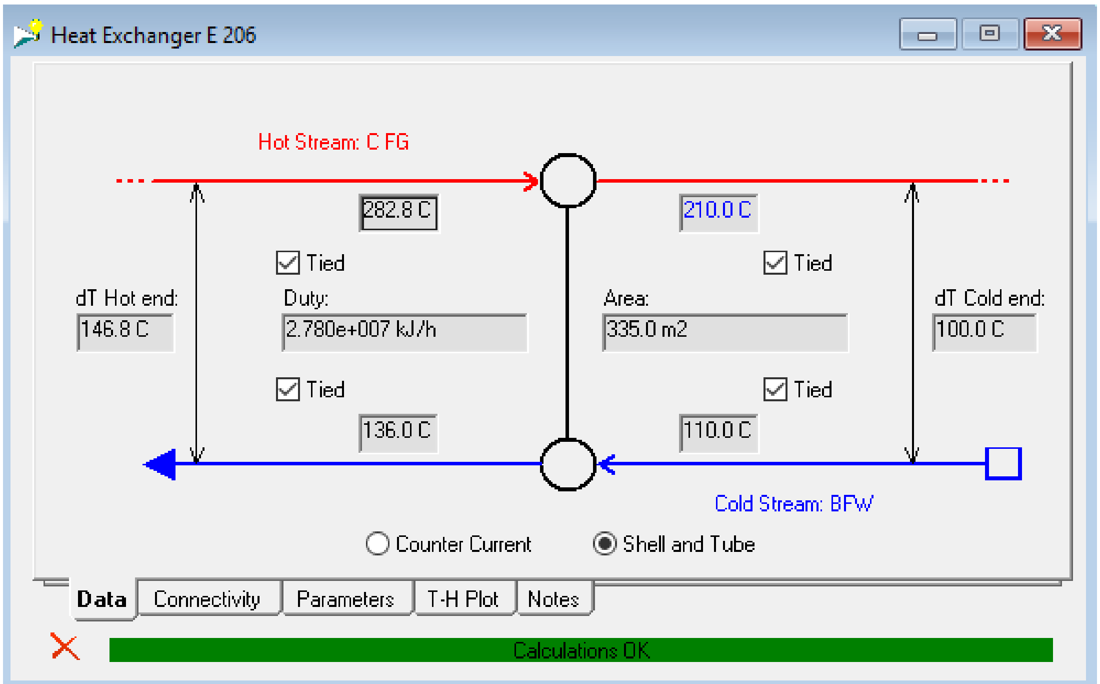
177.0 C

Cold Stream: Process Air

● Counter Current ○ Shell and Tube

Calculations OK

Figure 4.31: Case 2-E 202 B



Hot Stream: C FG

282.8 C

210.0 C

☑ Tied

☑ Tied

dT Hot end: 146.8 C

Duty: 2.780e+007 kJ/h

Area: 335.0 m2

dT Cold end: 100.0 C

☑ Tied

☑ Tied

136.0 C

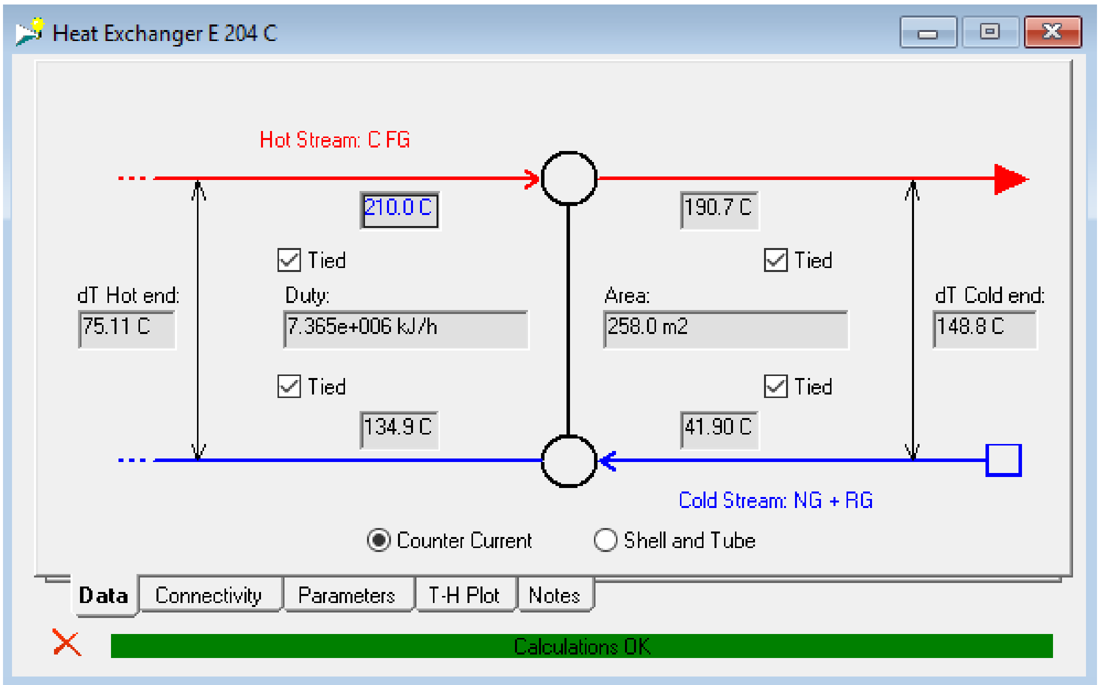
110.0 C

Cold Stream: BFW

○ Counter Current ● Shell and Tube

Calculations OK

Figure 4.32: Case 2-E 206



☑ Tied
☑ Tied
☑ Tied
☑ Tied

Hot Stream: C FG: 210.0 C → 190.7 C

Cold Stream: NG + RG: 41.90 C → 134.9 C

dT Hot end: 75.11 C

Duty: 7.365e+006 kJ/h

Area: 258.0 m2

dT Cold end: 148.8 C

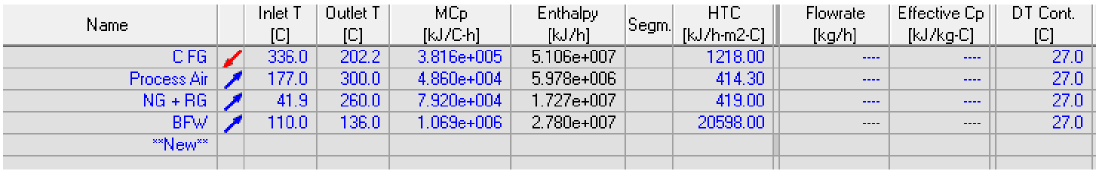
● Counter Current ○ Shell and Tube

Figure 4.33: Case 2-E 204 C

Similar issue is faced in this case as the area of exchangers are too large to use them in real system.

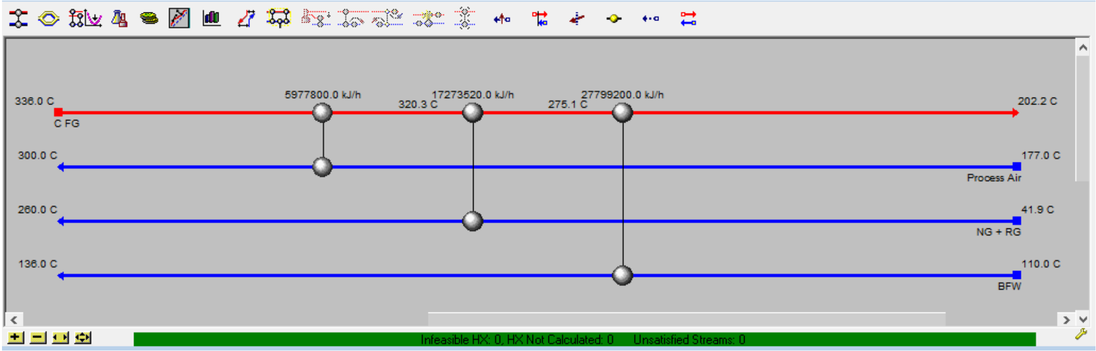
4.3.3 Case 3

In case 3, new pre heater to heat Process Air using Combine Flue Gas is added. However, temperature of Combine Flue Gas was reduced from 336 ºC to 202 ºC while Process Air is heated from 177 ºC to 300 ºC and Natural and Recycle Gas is heated to 260 ºC.



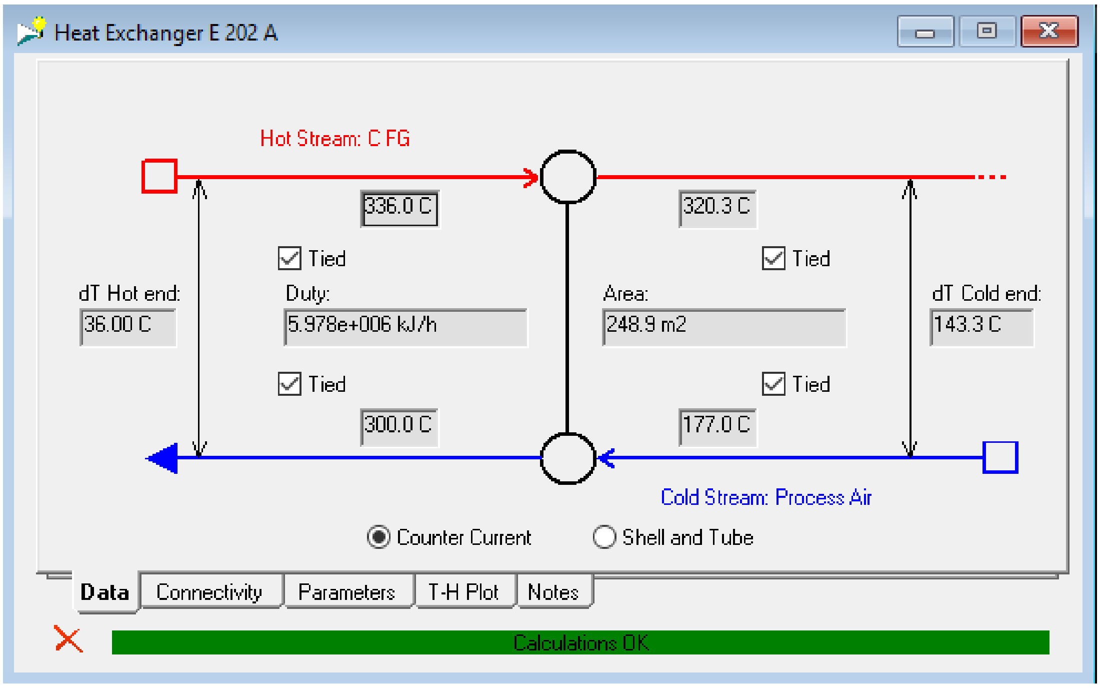
Name		Inlet T [C]	Outlet T [C]	MCp [kJ/C-h]	Enthalpy [kJ/h]	Segm.	HTC [kJ/h-m2-C]	Flowrate [kg/h]	Effective Cp [kJ/kg-C]	DT Cont. [C]
C FG	↗	336.0	202.2	3.816e+005	5.106e+007		1218.00	----	----	27.0
Process Air	↗	177.0	300.0	4.860e+004	5.978e+006		414.30	----	----	27.0
NG + RG	↗	41.9	260.0	7.920e+004	1.727e+007		419.00	----	----	27.0
BFW	↗	110.0	136.0	1.069e+006	2.780e+007		20598.00	----	----	27.0
New										

Figure 4.34: Case 3-Streams Data



Infeasible HX: 0, HX Not Calculated: 0 Unsatisfied Streams: 0

Figure 4.35: Case 3-Grid Diagram



Hot Stream: C FG

336.0 C

320.3 C

☑ Tied

☑ Tied

dT Hot end: 36.00 C

Duty: 5.978e+006 kJ/h

Area: 248.9 m2

dT Cold end: 143.3 C

☑ Tied

☑ Tied

300.0 C

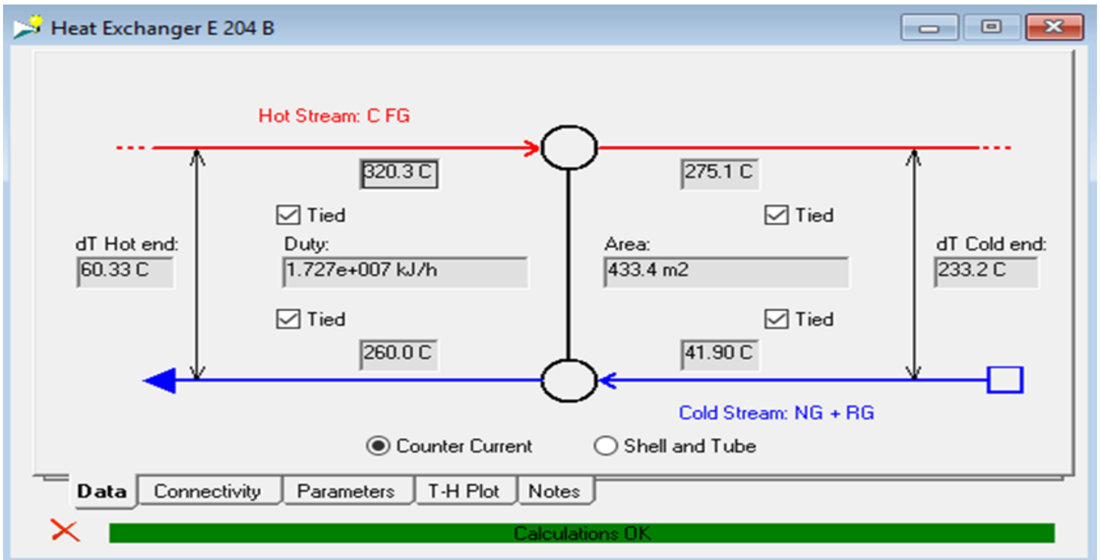
177.0 C

Cold Stream: Process Air

● Counter Current ○ Shell and Tube

Calculations OK

Figure 4.36: Case 3-E 202 B



Hot Stream: C FG

320.3 C

275.1 C

☑ Tied

☑ Tied

dT Hot end: 60.33 C

Duty: 1.727e+007 kJ/h

Area: 433.4 m2

dT Cold end: 233.2 C

☑ Tied

☑ Tied

260.0 C

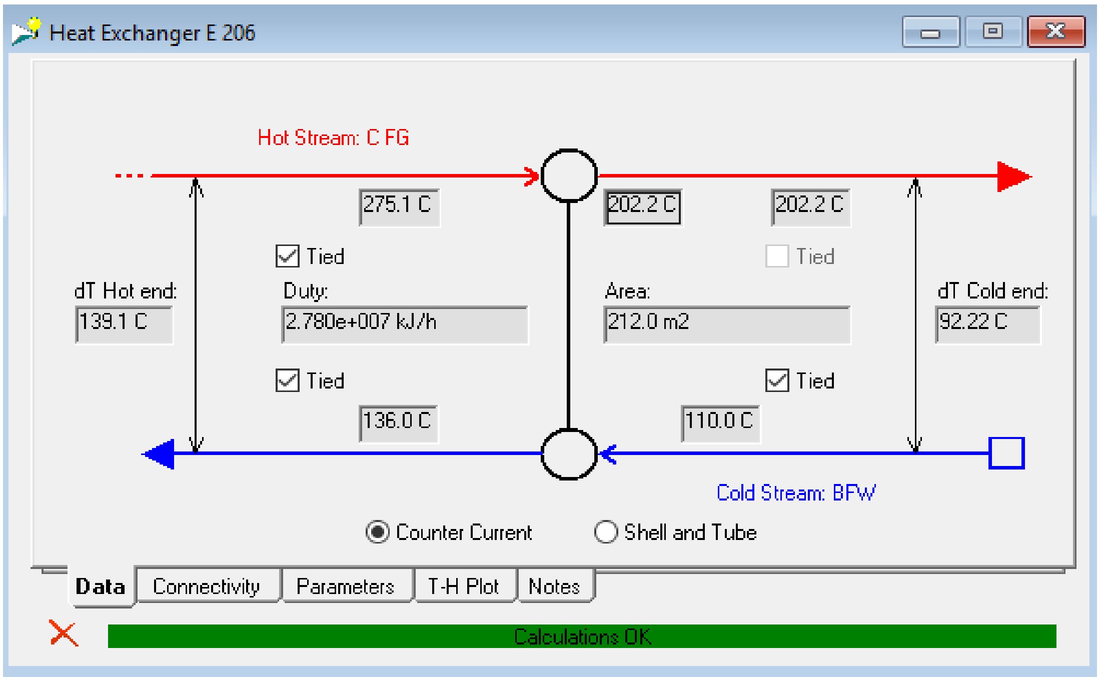
41.90 C

Cold Stream: NG + RG

● Counter Current ○ Shell and Tube

Calculations OK

Figure 4.37: Case 3-E 204 B



Hot Stream: C FG

275.1 C

202.2 C

202.2 C

☑ Tied

☐ Tied

dT Hot end: 139.1 C

Duty: 2.780e+007 kJ/h

Area: 212.0 m2

dT Cold end: 92.22 C

☑ Tied

☑ Tied

136.0 C

110.0 C

Cold Stream: BFW

● Counter Current ○ Shell and Tube

Calculations OK

Figure 4.38: Case 3-E 206

This is the post suitable case to apply new preheater for the Process Air as the areas estimated by Aspen Energy Analyzer of heat exchangers are also in practical range to design coils.

In the next chapter, the selected case will be applied to design new exchanger network and results will be calculated based on it.

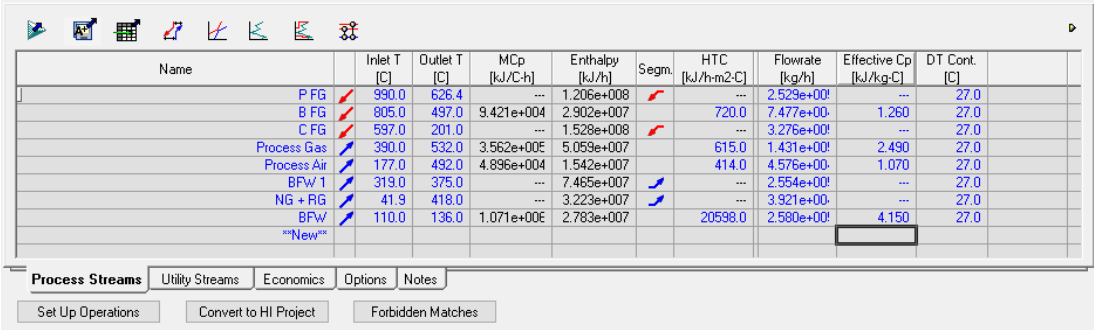
Chapter 5
New Heat Exchanger Network

Based on the solution suggested in chapter 4, proposed heat exchanger network is designed for all the process streams. Later new process flow diagram is drawn followed by comparing the results of both existing and proposed networks.

5.1 New Heat Exchanger Network Design

5.1.1 Defining Process Streams

Entire data for the process streams is same as the previous one except that outlet temperature of Combine Flue Gas is reduced to 201 ^{o}C.

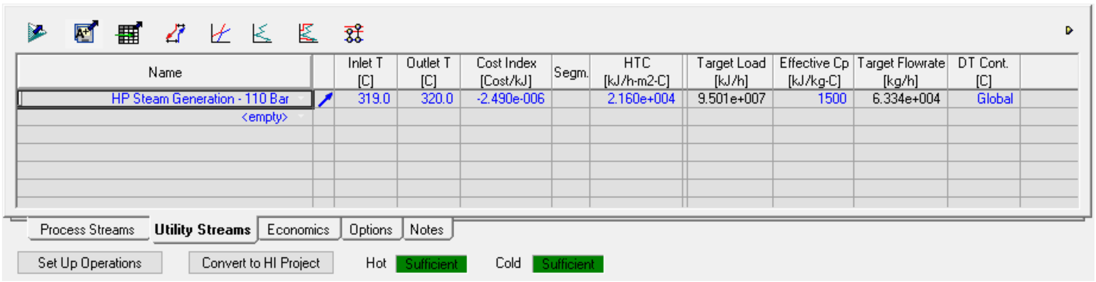


Name		Inlet T [C]	Outlet T [C]	MCp [kJ/C-h]	Enthalpy [kJ/h]	Segm.	HTC [kJ/h-m2-C]	Flowrate [kg/h]	Effective Cp [kJ/kg-C]	DT Cont. [C]
P FG	↙	990.0	626.4	---	1.206e+008	↙	---	2.529e+00!	---	27.0
B FG	↙	805.0	497.0	9.421e+004	2.902e+007		720.0	7.477e+00	1.260	27.0
C FG	↙	597.0	201.0	---	1.528e+008	↙	---	3.276e+00!	---	27.0
Process Gas	↗	390.0	532.0	3.562e+00E	5.059e+007		615.0	1.431e+00!	2.490	27.0
Process Air	↗	177.0	492.0	4.896e+004	1.542e+007		414.0	4.576e+00	1.070	27.0
BFW 1	↗	319.0	375.0	---	7.465e+007	↗	---	2.554e+00!	---	27.0
NG + RG	↗	41.9	418.0	---	3.223e+007	↗	---	3.921e+00	---	27.0
BFW	↗	110.0	136.0	1.071e+00E	2.783e+007		20598.0	2.580e+00!	4.150	27.0
New										

Figure 5.1: Modified Streams Data

5.1.2 Defining Utility Stream

Utility for the system same.



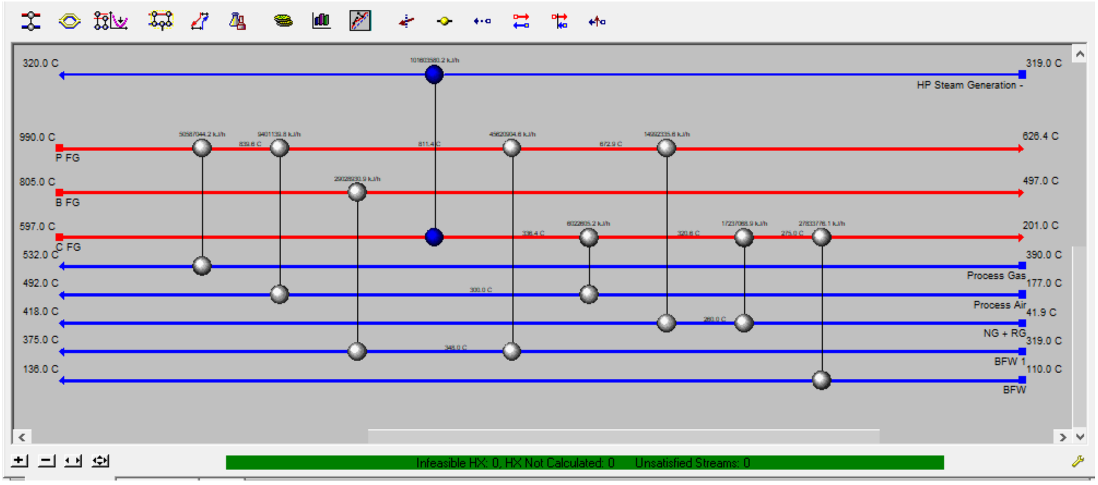
Name		Inlet T [C]	Outlet T [C]	Cost Index [Cost/kJ]	Segm.	HTC [kJ/h-m2-C]	Target Load [kJ/h]	Effective Cp [kJ/kg-C]	Target Flowrate [kg/h]	DT Cont. [C]
HP Steam Generation - 110 Bar	↗	319.0	320.0	-2.490e-006		2.160e+004	9.501e+007	1500	6.334e+004	Global
<empty>										

Hot: Sufficient
Cold: Sufficient

Figure 5.2: Utility Stream

5.1.3 New Heat Exchanger Network

In this network, a new preheater for Process Air is added after coil E 205. Total number of heat exchangers in the network are nine now.

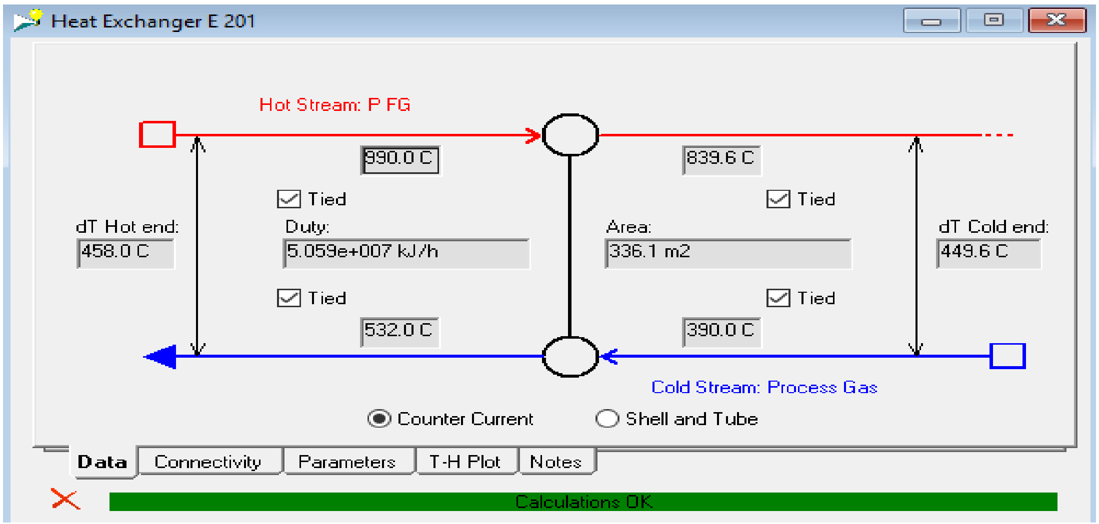


Infeasible HX: 0, HX Not Calculated: 0 Unsatisfied Streams: 0

Figure 5.3: Modified Heat Exchanger Network

5.1.3.1 Coil E 201

No change in the conditions of this exchanger.



Hot Stream: P FG

990.0 C ☑ Tied

839.6 C ☑ Tied

dT Hot end: 458.0 C

Duty: 5.059e+007 kJ/h

Area: 336.1 m2

dT Cold end: 449.6 C

532.0 C ☑ Tied

390.0 C ☑ Tied

Cold Stream: Process Gas

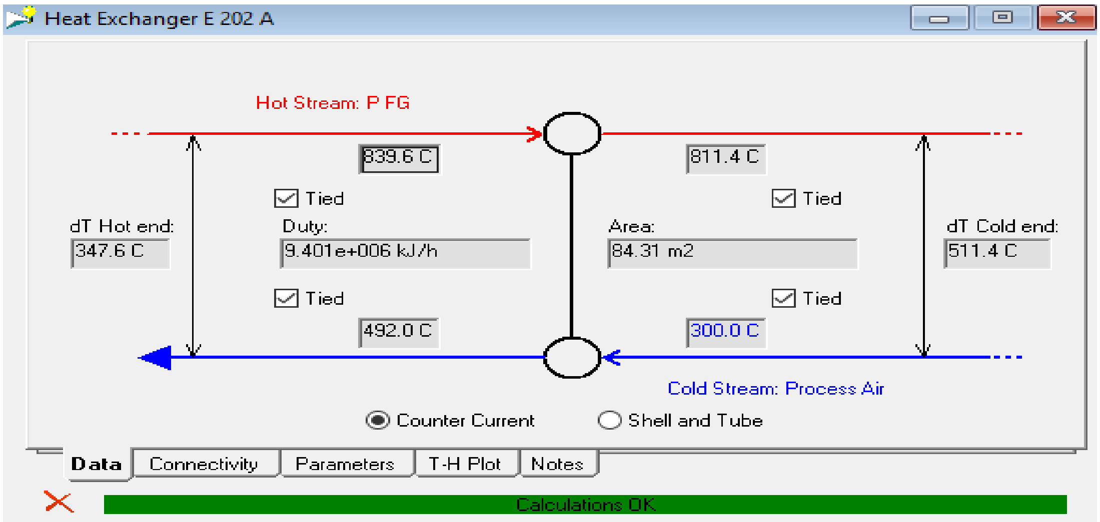
● Counter Current ○ Shell and Tube

Calculations OK

Figure 5.4: Coil E 201

5.1.3.2 Coil E 202 A

In this exchanger, inlet temperature of Process Air is now 300 ^{o}C instead of 177 ^{o}C while outlet temperature of Primary Reformer Flue Gas is dropped from 794 ^{o}C to 811 ^{o}C.



Heat Exchanger E 202 A

Hot Stream: P FG

839.6 C

811.4 C

dT Hot end: 347.6 C

Duty: 9.401e+006 kJ/h

Area: 84.31 m2

dT Cold end: 511.4 C

492.0 C

300.0 C

Cold Stream: Process Air

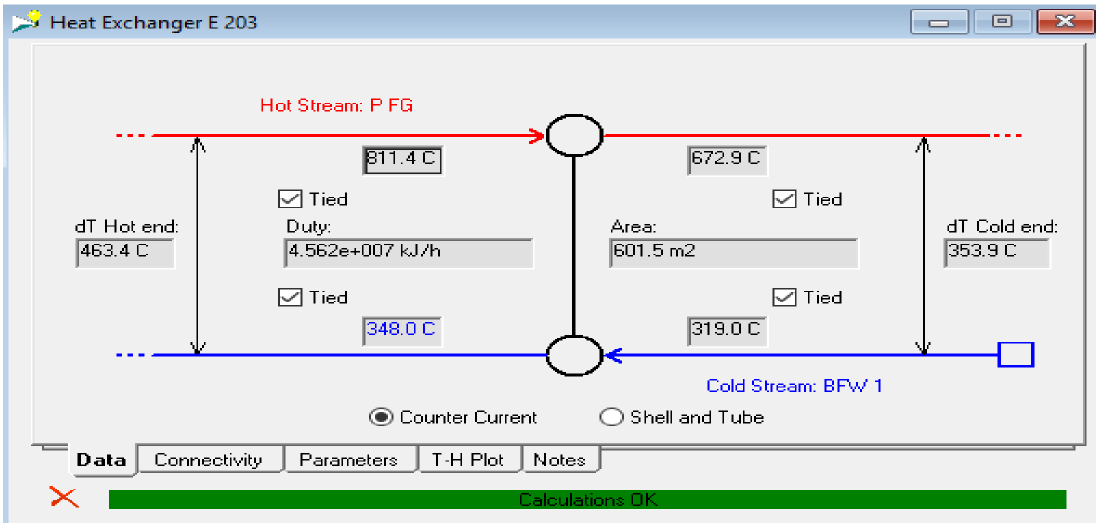
● Counter Current ○ Shell and Tube

Calculations OK

Figure 5.5: Coil E 202 A

5.1.3.3 Coil E 203

In this exchanger, conditions of the cold stream in same. However, inlet and outlet temperature of Primary Reformer Flue Gas are now 811 ^{o}C and 673 ^{o}C instead of 794 ^{o}C and 655 ^{o}C, respectively.



Heat Exchanger E 203

Hot Stream: P FG

811.4 C

672.9 C

dT Hot end: 463.4 C

Duty: 4.562e+007 kJ/h

Area: 601.5 m2

dT Cold end: 353.9 C

348.0 C

319.0 C

Cold Stream: BFW 1

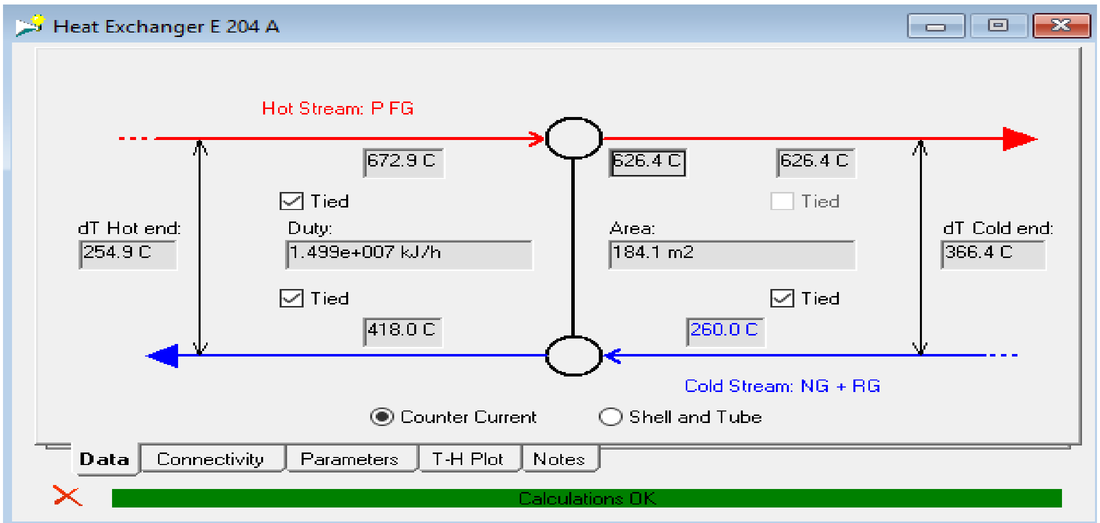
● Counter Current ○ Shell and Tube

Calculations OK

Figure 5.6: Coil E 203

5.1.3.4 Coil E 204 A

In this exchanger, inlet temperature of Natural and Recycle Gas is now 260 °C instead of 247 °C. Inlet and outlet temperature of Primary Reformer Flue Gas are changed from 655 °C and 604 °C to 673 °C and 626 °C, respectively.



Heat Exchanger E 204 A

Hot Stream: P FG

672.9 C

626.4 C

626.4 C

☑ Tied

☐ Tied

dT Hot end: 254.9 C

Duty: 1.499e+007 kJ/h

Area: 184.1 m2

dT Cold end: 366.4 C

☑ Tied

☑ Tied

418.0 C

260.0 C

Cold Stream: NG + RG

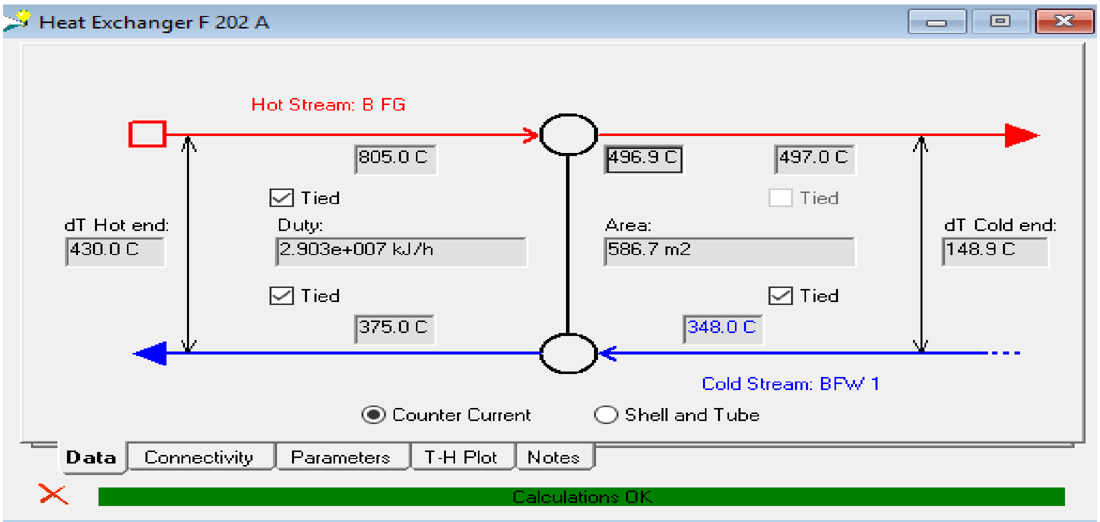
● Counter Current

○ Shell and Tube

Figure 5.7: Coil E 204 A

5.1.3.5 Coil F 202 A

No change in the conditions of this exchanger.



Heat Exchanger F 202 A

Hot Stream: B FG

805.0 C

496.9 C

497.0 C

☑ Tied

☐ Tied

dT Hot end: 430.0 C

Duty: 2.903e+007 kJ/h

Area: 586.7 m2

dT Cold end: 148.9 C

☑ Tied

☑ Tied

375.0 C

348.0 C

Cold Stream: BFW 1

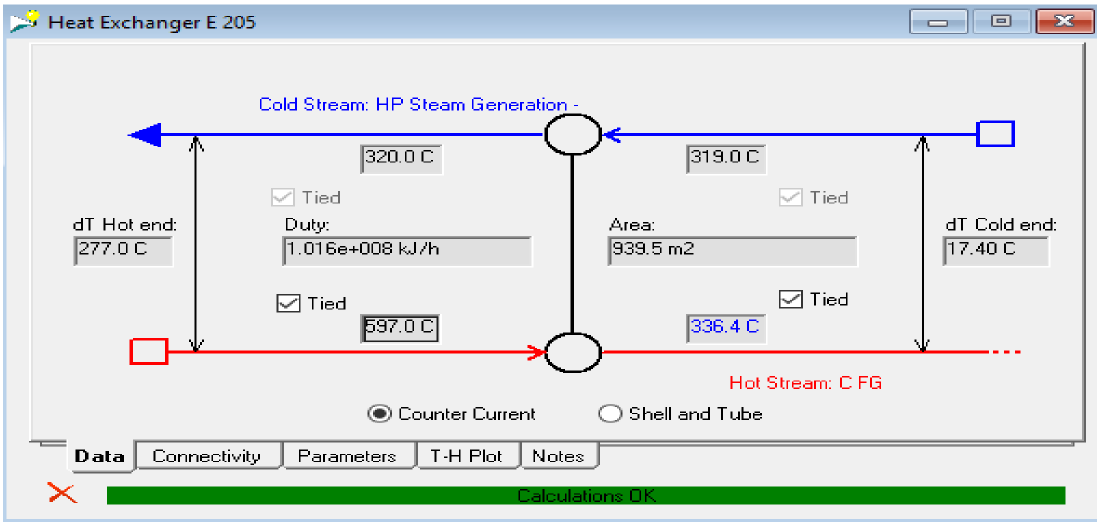
● Counter Current

○ Shell and Tube

Figure 5.8: Coil F 202 A

5.1.3.6 Coil E 205

In this exchanger, due to increase in the temperature of Primary Reformer Flue Gas, Combine flue Gas temperature is increased from 580 ^{o}C to 597 ^{o}C. However outlet temperature of Combine Flue Gas is kept constant to 336 ^{o}C as it will affect the pressure of steam being generated.



Heat Exchanger E 205

Cold Stream: HP Steam Generation -

320.0 C

319.0 C

☑ Tied

☑ Tied

dT Hot end: 277.0 C

Duty: 1.016e+008 kJ/h

Area: 939.5 m2

dT Cold end: 17.40 C

☑ Tied 597.0 C

☑ Tied 336.4 C

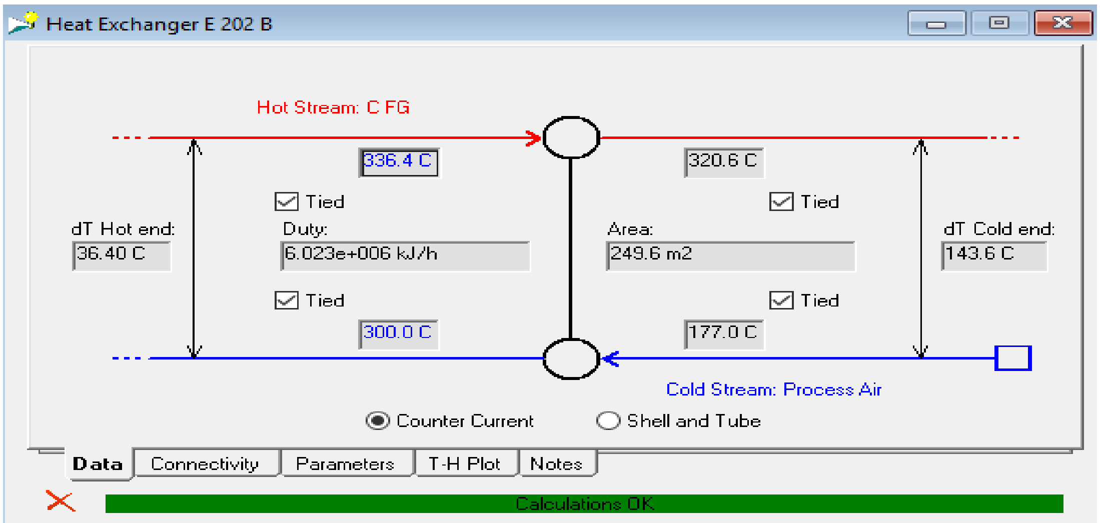
Hot Stream: C FG

● Counter Current ○ Shell and Tube

Figure 5.9: Coil E 205

5.1.3.7 Coil E 202 B

This is the new exchanger in the system. The inlet and outlet temperature of Process Air are estimated in previous chapter that are 177 ^{o}C and 300 ^{o}C, respectively.



Heat Exchanger E 202 B

Hot Stream: C FG

336.4 C

320.6 C

☑ Tied

☑ Tied

dT Hot end: 36.40 C

Duty: 6.023e+006 kJ/h

Area: 249.6 m2

dT Cold end: 143.6 C

☑ Tied 300.0 C

☑ Tied 177.0 C

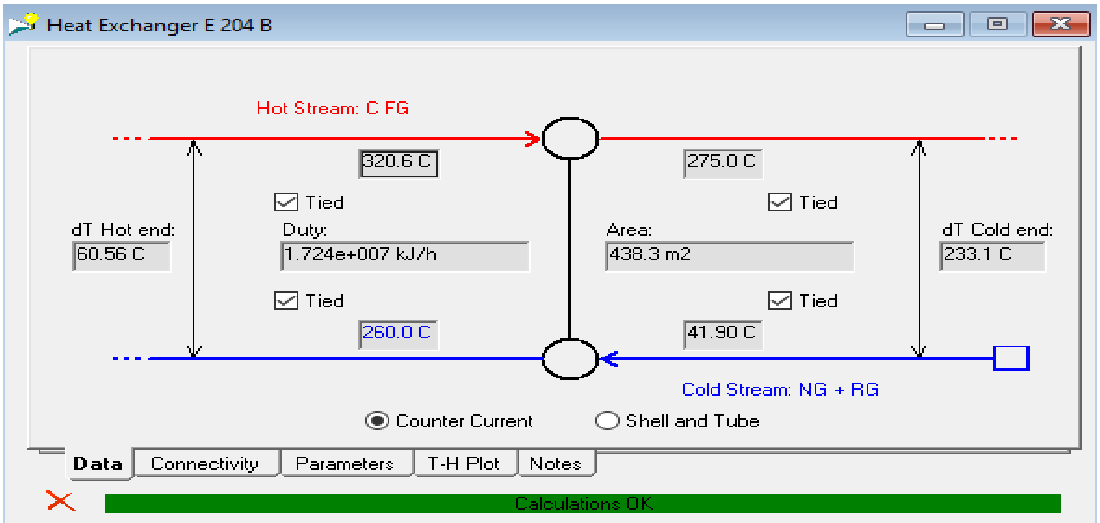
Cold Stream: Process Air

● Counter Current ○ Shell and Tube

Figure 5.10: Coil E 202 B

5.1.3.8 Coil E 204 B

In this exchanger, outlet temperature of Natural and Recycle Gas is increased to 260 °C from 247 °C. Inlet and outlet temperature of Combine Flue Gas are now 320 °C and 275 °C instead of 336 °C and 294 °C, respectively.

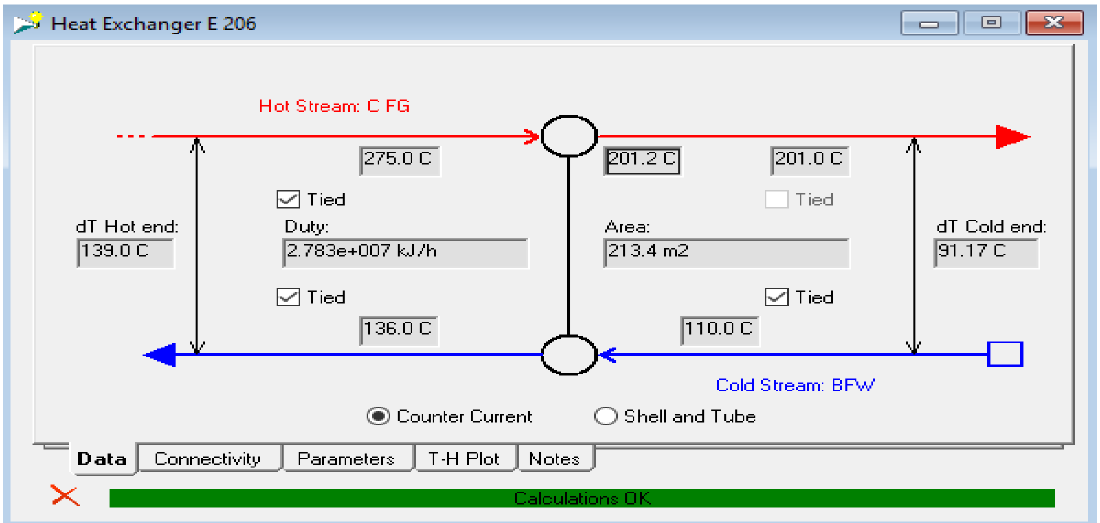


Hot Stream: C FG
320.6 C
275.0 C
☑ Tied
☑ Tied
dT Hot end: 60.56 C
Duty: 1.724e+007 kJ/h
Area: 438.3 m2
dT Cold end: 233.1 C
☑ Tied
☑ Tied
260.0 C
41.90 C
Cold Stream: NG + RG
● Counter Current ○ Shell and Tube
Calculations OK

Figure 5.11: Coil E 204 B

5.1.3.9 Coil E 206

In this exchanger, conditions of the cold stream in same. However, inlet and outlet temperature of Combine Flue Gas are now 275 °C and 201 °C instead of 294 °C and 220 °C, respectively.

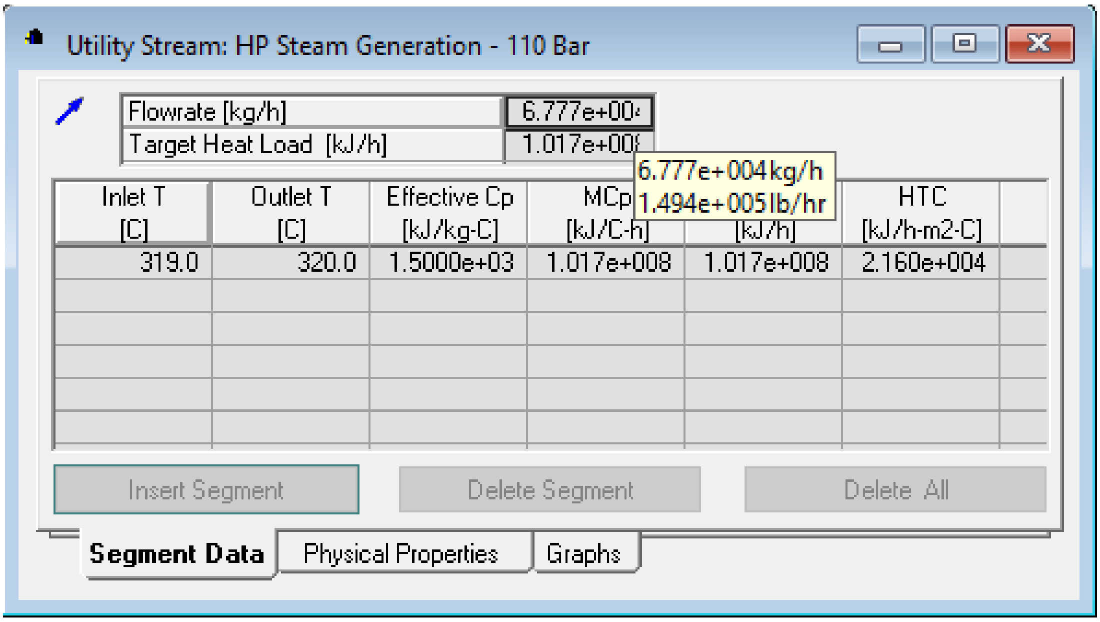


Hot Stream: C FG
275.0 C
201.2 C
201.0 C
☑ Tied
☐ Tied
dT Hot end: 139.0 C
Duty: 2.783e+007 kJ/h
Area: 213.4 m2
dT Cold end: 91.17 C
☑ Tied
☑ Tied
136.0 C
110.0 C
Cold Stream: BFW
● Counter Current ○ Shell and Tube
Calculations OK

Figure 5.12: Coil E 206

5.1.3.10 High Pressure Steam Production in Coil E 205

As a result of increased inlet temperature of Combine Flue Gas at coil E 205, production of High Pressure Steam is increased from 63340 kilograms per hour to 67770 kilograms per hour.



Utility Stream: HP Steam Generation - 110 Bar

Flowrate [kg/h]	6.777e+00
Target Heat Load [kJ/h]	1.017e+00

6.777e+004 kg/h
1.494e+005 lb/hr

Inlet T [C]	Outlet T [C]	Effective Cp [kJ/kg-C]	MCp [kJ/C-h]	[kJ/h]	HTC [kJ/h-m2-C]
319.0	320.0	1.5000e+03	1.017e+008	1.017e+008	2.160e+004

Figure 5.13: New High Pressure Steam Mass Flowrate

5.1.4 New Heat Exchanger Network Grid Diagram

Following is the completely satisfied Grid Diagram for new heat exchanger network:

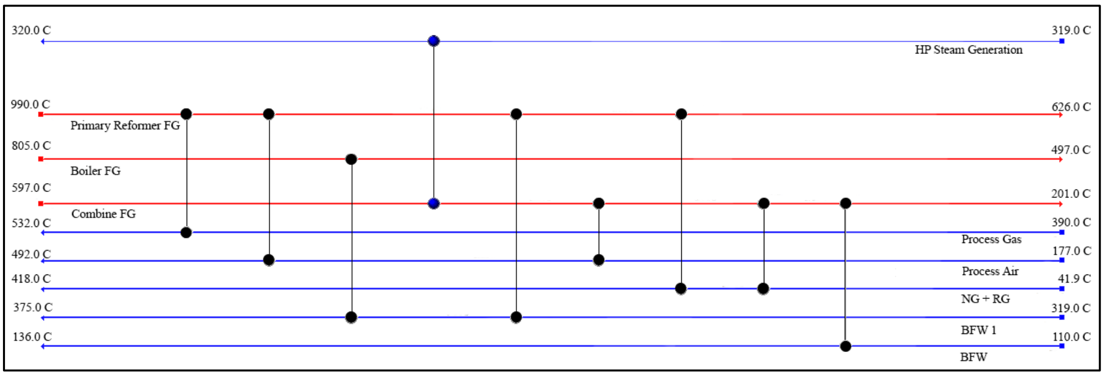


Figure 5.14 New Heat Exchanger Grid Network Design

5.2 Comparison of Networks

Following tables shows comparison between existing and proposed heat exchanger network.

Parameters	Existing Heat Exchanger Network	New Heat Exchanger Network
Stack Gas Temperature (^oC)	220	201
HP Steam Production (kg/hr)	63340	67770

Table 5-1: Comparison of New and Existing Heat Exchange Network

5.2 New Process Flow Diagram

Based on the new heat exchanger network, modified process flow diagram is drawn. In this diagram, coil E 202 B is added after coil E 205 to pre heat Process Air.

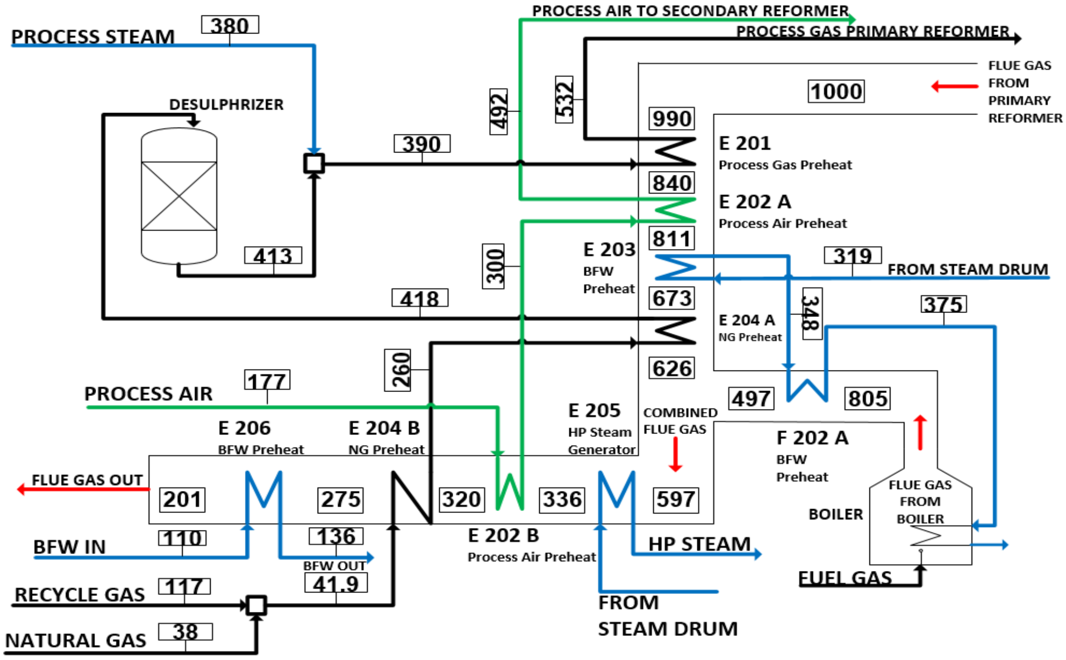


Figure 5.15: New Process Flow Diagram

5.3 Material and Energy Balance of New Process

After the new system is designed, material and energy balance on the modified process is performed.

5.3.1 Material Balance

As no reaction or mixing is take place, material balance of the modified process will remain same to that performed in chapter 3. However, due to addition of new coil and changing the temperatures of some of the streams, energy balance of the system will change and is shown in the next section.

5.3.2 Energy Balance

Reference Temperature = 25 °C

Balance on Natural and Recycle Gas Mixer

Stream	M (kg/hr)	c_p (kJ/kg °C)	T_S (°C)	H_{In} (kJ/hr)
Natural Gas	38240	1.92	38	954470.4
Recycle Gas	996	3.87	117	354615.84
Natural & Recycle Gas	39210	1.99	41.9	1318671.51

$$\Delta H_{In} = \Delta H_{Natural\ Gas} + \Delta H_{Recycle\ Gas}$$

$$\Delta H_{In} = 954470.4 + 354615.84$$

$$\Delta H_{In} = 1309086.24\ kJ/hr$$

$$\Delta H_{Out} = 1318671.51\ kJ/hr$$

$$\text{So, } \Delta H_{In} \cong \Delta H_{Out}$$

Balance on Desulphurized Gas and Process Steam Mixer

Stream	**M (kg/hr)**	**c_p (kJ/kg °C)**	**T_S (°C)**	**H_{In} (kJ/hr)**
Process Steam	103950	2.52	380	92993670
Desulphurized Gas	39210	2.42	413	36816621.6
Process Steam & Desulphurized Gas	143071	2.49	390	130030078.4

$$\Delta H_{In} = \Delta H_{Process\ steam} + \Delta H_{Desulphurized\ Gas}$$

$$\Delta H_{In} = 92993670 + 36816621.6$$

$$\Delta H_{In} = 129810291.6\ kJ/hr$$

$$\Delta H_{Out} = 130030078\ kJ/hr$$

$$\text{So, } \Delta H_{In} \cong \Delta H_{Out}$$

Balance on Primary Reformer Flue Gas and Boiler Flue Gas Mixer

Stream	**M (kg/hr)**	**c_p (kJ/kg °C)**	**T_S (°C)**	**H_{In} (kJ/hr)**
Primary Reformer Flue Gas	252861	1.3	626	197560325
Boiler Flue Gas	74772	1.26	497	44468452.52
Combined Flue Gas	327633	1.19	597	223013230.4

$$\Delta H_{In} = \Delta H_{Primary\ Reformer\ Flue\ Gas} + \Delta H_{Boiler\ Flue\ Gas}$$

$$\Delta H_{In} = 197560325 + 444.68452.52$$

$$\Delta H_{In} = 271046327.1\ kJ/hr$$

$$\Delta H_{Out} = 223013230.4\ kJ/hr$$

$$\text{So, } \Delta H_{In} \cong \Delta H_{Out}$$

Balance on Coil E 201

Stream	**M (kg/hr)**	**c_p (kJ/kg °C)**	**T_S (°C)**	**T_T (°C)**	**H_{In} (kJ/hr)**	**H_{Out} (kJ/hr)**
Process Gas	143071	2.49	390	532	130030078.4	180617122.5
Primary Reformer Flue Gas	252861	1.33	990	840	324534492.7	274088716.6

Duty E 201 = $Q_{E\ 201}$ = 12.06 Gcal/hr = 50410800 kJ/hr

For Process Gas,

$$Q = \Delta H_{Out} - \Delta H_{In}$$

$$Q = 180617122.5 - 130030078.4$$

$$Q = 50587044.18\ kJ/hr$$

$$\text{So, } Q_{E\ 201} \cong Q$$

For Primary Reformer Flue Gas,

$$Q = \Delta H_{In} - \Delta H_{Out}$$

$$Q = 324534492.7 - 274088716.6$$

$$Q = 50445776.07\ kJ/hr$$

$$\text{So, } Q_{E\ 201} \cong Q$$

Balance on Coil E 202 A

Stream	**M (kg/hr)**	**c_p (kJ/kg °C)**	**T_S (°C)**	**T_T (°C)**	**H_{In} (kJ/hr)**	**H_{Out} (kJ/hr)**
Process Air	45761	1.07	300	492	13465174.25	22866314.1
Primary Reformer Flue Gas	252861	1.33	840	811	274088716.6	262348378.9

Duty E 202 A = $Q_{E\ 202\ A}$ = ?

For Process Air,

$$Q = \Delta H_{Out} - \Delta H_{In}$$

$$Q = 22866314.09 - 13465174.25$$

$$Q = 9401139.84\ kJ/hr$$

$$\text{So, } Q_{E\ 202\ A} = Q = 9401139.84\ kJ/hr$$

For Primary Reformer Flue Gas,

$$Q = \Delta H_{In} - \Delta H_{Out}$$

$$Q = 274088716.6 - 26234873.9$$

$$Q = 11740337.76 kJ/hr$$

$$\text{So, } Q_{E\ 202\ A} \cong Q$$

$$Q_{E\ 202\ A} = 2.25\ Gcal/hr$$

Balance on Coil E 203

Stream	**M (kg/hr)**	**c_p (kJ/kg °C)**	**T_S (°C)**	**T_T (°C)**	**H_{In} (kJ/hr)**	**H_{Out} (kJ/hr)**
Steam from Steam Drum	143071	2.49	319	348	316092803.5	345622277.2
Primary Reformer Flue Gas	252861	1.3	811	673	262348378.9	213010134.0

Duty E 203 = $Q_{E\ 203}$ = 10.9 Gcal/hr = 45562000 kJ/hr

For Steam from Steam Drum,

$$Q = \Delta H_{Out} - \Delta H_{In}$$

$$Q = 345622277.2 - 316092803.5$$

$$Q = 29529473.77\ kJ/hr$$

$$\text{So, } Q_{E\ 203} \cong Q$$

For Primary Reformer Flue Gas,

$$Q = \Delta H_{In} - \Delta H_{Out}$$

$$Q = 262348378.9 - 213010134$$

$$Q = 49338244.75\ kJ/hr$$

$$\text{So, } Q_{E\ 203} \cong Q$$

Balance on Coil E 204 A

Stream	**M (kg/hr)**	**c_p (kJ/kg °C)**	**T_S (°C)**	**T_T (°C)**	**H_{In} (kJ/hr)**	**H_{Out} (kJ/hr)**
Natural and Recycle Gas	39210	2.2	260	418	18336556	33900966
Primary Reformer Flue Gas	252861	1.33	673	626	213010134	197560325

Duty E 204 A = $Q_{E\ 204\ A}$ = ?

For Natural and Recycle Gas,

$$Q = \Delta H_{Out} - \Delta H_{In}$$

$$Q = 33900966 - 18336556$$

$$Q = 15564409.5\ kJ/hr$$

$$\text{So, } Q_{E\ 204\ A} = Q = 15564409.5\ kJ/hr$$

For Primary Reformer Flue Gas,

$$Q = \Delta H_{In} - \Delta H_{Out}$$

$$Q = 213010134.1 - 197560325$$

$$Q = 15449809.11\ kJ/hr$$

$$\text{So, } Q_{E\ 204\ A} \cong Q$$

$$Q_{E\ 204\ A} = 3.72\ Gcal/hr$$

Balance on Coil F 202 A

Stream	**M** **(kg/hr)**	**c_p** **(kJ/kg °C)**	**T_S** **(°C)**	**T_T** **(°C)**	**H_{In}** **(kJ/hr)**	**H_{Out}** **(kJ/hr)**
Steam from Steam Drum	255379	4.19	348	375	345622277	374513303
Boiler Flue Gas	74772	1.26	805	497	73486002	44468452

Duty F 202 A = $Q_{F\ 202\ A}$ = 6.93 Gcal/hr = 28967400 kJ/hr

For Steam from Steam Drum,

$$Q = \Delta H_{Out} - \Delta H_{In}$$

$$Q = 374513303.5 - 345622277.2$$

$$Q = 28891026.27 \ kJ/hr$$

$$\text{So, } Q_{F\ 202\ A} \cong Q$$

For Boiler Flue Gas,

$$Q = \Delta H_{In} - \Delta H_{Out}$$

$$Q = 73486002.5 - 44468452.52$$

$$Q = 29017549.53 \ kJ/hr$$

$$\text{So, } Q_{F\ 202\ A} \cong Q$$

Balance on Coil E 205

Stream	M (kg/hr)	c_p (kJ/kg °C)	T_S (°C)	T_T (°C)	H_{In} (kJ/hr)	H_{Out} (kJ/hr)
Combined Flue Gas	327633	1.19	597	336	223013230.4	118196881.1

Duty E 205 = $Q_{E\ 205}$ = 22.81 Gcal/hr = 95345800 kJ/hr

For Combined Flue Gas,

$$Q = \Delta H_{In} - \Delta H_{Out}$$

$$Q = 216385214.9 - 118196881.1$$

$$Q = 104816349.4 kJ/hr$$

$$\text{So, } Q_{E\ 205} \cong Q$$

Balance on Heat Exchanger E 202 B

Stream	M (kg/hr)	c_p (kJ/kg °C)	T_S (°C)	T_T (°C)	H_{In} (kJ/hr)	H_{Out} (kJ/hr)
Process Air	45761	1.07	177	300	7442569.04	13465174.25
Combined Flue Gas	327633	1.16	336	320	118196881.1	112116013

Duty E 202 B = $Q_{E\ 202\ B}$ = ?

For Process Air,

$$Q = \Delta H_{Out} - \Delta H_{In}$$

$$Q = 33900966 - 18336556$$

$$Q = 15564409.5\ kJ/hr$$

$$\text{So, } Q_{E\ 202\ B} = Q = 15564409.5\ kJ/hr$$

For Combine Flue Gas,

$$Q = \Delta H_{In} - \Delta H_{Out}$$

$$Q = 213010134.1 - 197560325$$

$$Q = 15449809.11 \; kJ/hr$$

$$\text{So, } Q_{E\ 202\ B} \cong Q$$

$$Q_{E\ 202\ B} = 3.72 \; Gcal/hr$$

Balance on Coil E 204 B

Stream	**M (kg/hr)**	**c_p (kJ/kg °C)**	**T_S (°C)**	**T_T (°C)**	**H_{In} (kJ/hr)**	**H_{Out} (kJ/hr)**
Natural and Recycle Gas	39210	2.49	41.9	260	1318671.51	18336556.5
Combined Flue Gas	327633	1.16	320	275	112116013	94194487.5

Duty E 204 B = $Q_{E\ 204\ B}$ = ?

For Natural and Recycle Gas,

$$Q = \Delta H_{Out} - \Delta H_{In}$$

$$Q = 33900966 - 18336556$$

$$Q = 17017884.99 \; kJ/hr$$

$$\text{So, } Q_{E\ 204\ B} = Q = 17017884.99 \; kJ/hr$$

For Combine Flue Gas,

$$Q = \Delta H_{In} - \Delta H_{Out}$$

$$Q = 213010134.1 - 197560325$$

$$Q = 17921525.1 \; kJ/hr$$

$$\text{So, } Q_{E\ 204\ B} \cong Q$$

$$Q_{E\ 204\ B} = 4.07\ Gcal/hr$$

Balance on Coil E 206

Stream	**M (kg/hr)**	**c_p (kJ/kg °C)**	**T_S (°C)**	**T_T (°C)**	**H_{In} (kJ/hr)**	**H_{Out} (kJ/hr)**
Boiler Feed Water	257959	4.2	110	136	92091363	119974151.3
Combined Flue Gas	327633	1.16	275	201	94194487.5	66312919.2

Duty E 206 = $Q_{E\ 206}$ = ?

For Natural and Recycle Gas,

$$Q = \Delta H_{Out} - \Delta H_{In}$$

$$Q = 33900966 - 18336556$$

$$Q = 27882788.31\ kJ/hr$$

$$\text{So, } Q_{E\ 206} = Q = 27882788.31\ kJ/hr$$

For Primary Reformer Flue Gas,

$$Q = \Delta H_{In} - \Delta H_{Out}$$

$$Q = 213010134.1 - 197560325$$

$$Q = 27881568.3\ kJ/hr$$

$$\text{So, } Q_{E\ 206} \cong Q$$

$$Q_{E\ 206} = 6.67\ Gcal/hr$$

After performing energy balance on the new system, it was determined that duty of some of the coils has changed. Design of these coils are given in details in the next chapter.

Chapter 6
Coils Designing

In the previous chapter, duty of the following coils were determined.

- E 202 A
- E 202 B
- E 204 A
- E 204 B
- E 206

In this chapter detail designing of these coils is performed.

6.1 Design Parameters

These coils are designed as the helical coil heat exchangers. For this purpose following parameters for each coil are calculated. Equations used to design these parameters are given along with them.

- **Heat Load (Q)**

$$\Delta H = Q = M \times c_p (T_S - T_T) = C_P \left(T_S - T_T\right)$$

- **Overall Heat Transfer Coefficient (U)**

$$\frac{1}{U} = \frac{1}{h_{Hot}} + \frac{1}{h_{Cold}}$$

- **Log Mean Temperature Difference (ΔT_{LM})**

$$\Delta T_{LM} = \frac{(T_{Hot\ S} - T_{Cold\ T}) - (T_{Hot\ T} - T_{Cold\ S})}{\ln \frac{(T_{Hot\ S} - T_{Cold\ T})}{(T_{Hot\ T} - T_{Cold\ S})}}$$

- **Area (A)**

$$A = \frac{Q}{U \times \Delta T_{LM}}$$

- **Length (L)**

$$L = \frac{A}{\pi \times d_o}$$

- **Pitch (p)**

$$p = 1.5 \times d_o$$

- **Number of Turns (N)**

$$N = \frac{L}{\sqrt{(2\pi \times R)^2 + p^2}}$$

Where following factors are known

- Outside Diameter of Coil (d_o)
- Helix Radius (R)
- Mass flowrate of streams passing through coils (M)
- Specific Heat Capacity of streams passing through coils (c_p)
- Streams Supply and Target Temperatures
- Heat transfer coefficient of Hot Fluid (h_{hot})
- Heat transfer coefficient of Hot Fluid (h_{cold})

6.2 Designing of Coils

6.2.1 Coil E 202 A

In coil E 202 A, Process Air is heated by Primary Reformer Flue Gas. Following information is known.

Outside Diameter of Coil (d_o) = 88.9 mm

Helix Radius (R) = 1.15 m

Stream	**M (kg/hr)**	**c_p (kJ/kg °C)**	**T_S (°C)**	**T_T (°C)**	**h (W/m² °C)**
Process Air	45761	1.07	300	492	115
Primary Reformer Flue Gas	252861	1.31	840	811	200

$$C_{P\,Cold} = 45761 \times 1.07/3600$$

$$C_{P\,Cold} = 13.6\ kW/℃$$

$$Q = 13.6 \times (492 - 300)$$

$$Q = 2611.4\ kW$$

$$\frac{1}{U} = \frac{1}{200} + \frac{1}{115}$$

$$U = 73.02\ W/m^2℃$$

$$\Delta T_{LM} = \frac{(840 - 492) - (811 - 300)}{\ln\frac{(840 - 492)}{(811 - 300)}}$$

$$\Delta T_{LM} = 424.3℃$$

$$A = \frac{2611.4 \times 1000}{73.02 \times 424.3}$$

$$A = 84.3\ m^2$$

$$L = \frac{84.3}{\pi \times (88.9/1000)}$$

$$L = 302\ \mathrm{m}$$

$$p = 1.5 \times (88.9/1000)$$

$$p = 0.133$$

$$N = \frac{302}{\sqrt{(2\pi \times 1.15)^2 + 0.133^2}}$$

$$N = 42$$

6.2.2 Coil E 202 B

In coil E 202 B, Process Air is heated by Combine Flue Gas. Following information is known.

Outside Diameter of Coil (d_o) = 114.3 mm

Helix Radius (R) = 1.48 m

Stream	**M** **(kg/hr)**	**c_p** **(kJ/kg °C)**	**T_S** **(°C)**	**T_T** **(°C)**	**h** **(W/m² °C)**
Process Air	45761	1.07	177	300	115
Combine Flue Gas	327633	1.18	336	320	338

$$C_{P\ Cold} = 45761 \times 1.07/3600$$

$$C_{P\ Cold} = 13.6\ kW/℃$$

$$Q = 13.6 \times (300 - 177)$$

$$Q = 1673\ kW$$

$$\frac{1}{U} = \frac{1}{338} + \frac{1}{115}$$

$$U = 85.8\ W/m^2℃$$

$$\Delta T_{LM} = \frac{(336 - 300) - (320 - 177)}{\ln\frac{(336 - 300)}{(320 - 177)}}$$

$$\Delta T_{LM} = 77.57\ ℃$$

$$A = \frac{1673 \times 1000}{77.57 \times 85.81}$$

$$A = 251.34\ m^2$$

$$L = \frac{251.34}{\pi \times (114.3/1000)}$$

$$L = 700\ \mathrm{m}$$

$$p = 1.5 \times (114.3/1000)$$

$$p = 0.171$$

$$N = \frac{700}{\sqrt{(2\pi \times 1.48)^2 + 0.171^2}}$$

$$N = 75$$

6.2.3 Coil E 204 A

In coil E 204 A, Natural and Recycle Gas is heated by Primary Reformer Flue Gas. Following information is known.

Outside Diameter of Coil (d_o) = 88.9 mm

Helix Radius (R) = 1.15 m

Stream	**M (kg/hr)**	**c_p (kJ/kg °C)**	**T_S (°C)**	**T_T (°C)**	**h (W/m^2 °C)**
Natural and Recycle Gas	39210	2.18	260	418	116.4
Primary Reformer Flue Gas	252861	1.33	673	626	200

$$C_{P\,Cold} = 39210 \times 2.18/3600$$

$$C_{P\,Cold} = 23.72\ kW/°\text{C}$$

$$Q = 23.72 \times (418 - 260)$$

$$Q = 4164\ kW$$

$$\frac{1}{U} = \frac{1}{200} + \frac{1}{116.4}$$

$$U = 73.58\ W/m^2°\text{C}$$

$$\Delta T_{LM} = \frac{(673 - 418) - (626 - 260)}{\ln\frac{(673 - 418)}{(626 - 260)}}$$

$$\Delta T_{LM} = 307.16\ °\text{C}$$

$$A = \frac{4164 \times 1000}{307.2 \times 73.58}$$

$$A = 184.2\ m^2$$

$$L = \frac{184.2}{\pi \times (88.9/1000)}$$

$$L = 660\ \text{m}$$

$$p = 1.5 \times (88.9/1000)$$

$$p = 0.133$$

$$N = \frac{660}{\sqrt{(2\pi \times 1.15)^2 + 0.133^2}}$$

$$N = 91$$

6.2.4 Coil E 204 B

In coil E 204 B, Natural and Recycle Gas is heated by Combine Flue Gas. Following information is known.

Outside Diameter of Coil (d_o) = 114.3 mm

Helix Radius (R) = 1.48 m

Stream	**M (kg/hr)**	**c_p (kJ/kg °C)**	**T_S (°C)**	**T_T (°C)**	**h (W/m² °C)**
Natural and Recycle Gas	39210	2.18	41.9	260	116.4
Combine Flue Gas	327633	1.18	320	275	338

$$C_{P\ Cold} = 39210 \times 2.18/3600$$

$$C_{P\ Cold} = 23.72\ kW/°\text{C}$$

$$Q = 23.72 \times (260 - 41.9)$$

$$Q = 4789\ kW$$

$$\frac{1}{U} = \frac{1}{338} + \frac{1}{116.4}$$

$$U = 86.6\ W/m^2°\text{C}$$

$$\Delta T_{LM} = \frac{(320 - 260) - (275 - 41.9)}{\ln\frac{(320 - 260)}{(275 - 41.9)}}$$

$$\Delta T_{LM} = 127.55\ °\text{C}$$

$$A = \frac{4789 \times 1000}{86.6 \times 127.55}$$

$$A = 433.63\ m^2$$

$$L = \frac{433.63}{\pi \times (114.3/1000)}$$

$$L = 1208 \text{ m}$$

$$p = 1.5 \times (114.3/1000)$$

$$p = 0.171$$

$$N = \frac{1208}{\sqrt{(2\pi \times 1.48)^2 + 0.171^2}}$$

$$N = 130$$

6.2.5 Coil E 206

In coil E 206, Boiler Feed Water is heated by Combine Flue Gas. Following information is known.

Outside Diameter of Coil (d_o) = 114.3 mm

Helix Radius (R) = 1.48 m

Stream	**M (kg/hr)**	**c_p (kJ/kg °C)**	**T_S (°C)**	**T_T (°C)**	**h (W/m^2 °C)**
Boiler Feed Water	252861	4.15	110	136	5722
Combine Flue Gas	327633	1.18	320	201	338

$$C_{P\ Cold} = 252861 \times 12.8/3600$$

$$C_{P\ Cold} = 297.01\ kW/°\text{C}$$

$$Q = 297.01 \times (136 - 110)$$

$$Q = 7730\ kW$$

$$\frac{1}{U} = \frac{1}{338} + \frac{1}{5722}$$

$$U = 319.15\ W/m^2 °\text{C}$$

$$\Delta T_{LM} = \frac{(275 - 136) - (201 - 110)}{\ln \frac{(275 - 136)}{(201 - 110)}}$$

$$\Delta T_{LM} = 113.3\ ℃$$

$$A = \frac{7730 \times 1000}{319.15 \times 113.3}$$

$$A = 213.76\ m^2$$

$$L = \frac{213.76}{\pi \times (114.3/1000)}$$

$$L = 595.7\ \mathrm{m}$$

$$p = 1.5 \times (114.3/1000)$$

$$p = 0.171$$

$$N = \frac{595.7}{\sqrt{(2\pi \times 1.48)^2 + 0.171^2}}$$

$$N = 64$$

6.3 Coil Design Summary

Following table shows the summary of coil design:

Parameters	**E 202 A**	**E 202 B**	**E 204 A**	**E 204 B**	**E 206**
Heat Load (kW)	2611.4	1673	4163.8	4788.8	7730
Hot Fluid T_S (^oC)	840	336	673	320	275
Hot Fluid T_T (^oC)	811	320	626	275	201
Cold Fluid T_S (^oC)	300	177	260	41.9	110
Cold Fluid T_T (^oC)	492	300	418	260	136
Hot Fluid h ($W/m^2\ ^oC$)	200	338	200	338	338
Cold Fluid h ($W/m^2\ ^oC$)	115	115	116.4	116.4	5722
Overall Heat Transfer Coefficient U ($W/m^2\ ^oC$)	73.02	85.81	73.58	86.58	319
LMTD (^oC)	424.29	77.57	307.16	127.55	113
Area (m^2)	84.29	251.34	184.24	433.63	213.8
Coil Outer Diameter d_o (mm)	88.9	114.3	88.9	114.3	114.3
Helix Radius R (m)	1.15	1.48	1.15	1.48	1.48
Length L (m)	301.97	700.30	660.00	1208.2	595.6
Pitch p (m)	0.133	0.171	0.133	0.171	0.171
Number of Turns (N)	42	75	91	130	64

Table 6-1: Coils Design Summary

6.4 Area Targeting

To calculate the minimum area requirement for the heat exchanger network, balanced combine composite curve is used. Balanced combine composite curve is same as combine composite curve, in addition to it also includes utility streams. By including utility streams, both hot and cold composite curve are in energy balance with one another. First minimum cold utility requirement is calculated using problem table algorithm for the new stream data.

6.4.1 Stream Data

Modified streams data is as followed:

Process Streams	**M (kg/hr)**	**c_p (kJ/kg °C)**	**C_P (kW/ °C)**	**T_S (°C)**	**T_T (°C)**	**ΔH (kJ/hr)**	**h (W/m² °C)**
Natural & Recycle Gas	39210	2.18	23.7	41	418	8943	116.4
Process Air	45761	1.07	13.6	177	492	4280	115.0
Process Gas	143071	2.67	106	400	532	14026	170.8
Steam from Steam Drum	255379	5.22	370	319	375	20736	70.0
BFW	257959	4.15	297	110	136	7722	5721
Primary Reformer Flue Gas	252861	1.31	91.9	990	626	35483	200.0
Boiler Flue Gas	74772	1.26	26.1	805	497	8060	200.0
Combine Flue Gas	327633	1.18	107	597	201	38693	338.3

Table 6-2: Modified Streams Data

Streams shifted temperatures are as followed:

Process Streams	**T_S (^oC)**	**T_T (^oC)**	**Stream Type**	**Shift T_S (^oC)**	**Shift T_T (^oC)**
Natural & Recycle Gas	41	418	Cold	54.5	431.5
Process Air	177	492	Cold	190.5	505.5
Process Gas	400	532	Cold	413.5	545.5
Steam	319	375	Cold	332.5	388.5
BFW	110	136	Cold	123.5	149.5
Primary Reformer Flue Gas	990	629	Hot	976.5	612.5
Boiler Flue Gas	805	497	Hot	791.5	483.5
Combine Flue Gas	597	201	Hot	583.5	187.5

Table 6-3: Modified Streams Shifted Temperatures

$$\Delta \boldsymbol{T_{min}} = \boldsymbol{27}^{\circ}\mathrm{C}$$

6.4.2 Problem Table Algorithm

To determine minimum cold utility requirement, problem table algorithm is developed for modified streams data.

Shift Temperature (^oC)	Interval	$T_{(i+1)}$-T_i (^oC)	$C_{P\ net}$ (kW/^oC)	dH (kW)		Infeasible/ Feasible Cascade
976.5					Pinch	0
	1	185	91.9	17001.5	S	
791.5						17002
	2	179	118.1	21139.9	S	
612.5						38141
	3	29	26.2	759.8	S	
583.5						38901
	4	38	133.7	5080.6	S	
545.5						43982
	5	40	27.4	1096.0	S	
505.5						45078
	6	22	13.8	303.6	S	
483.5						45381
	7	52	-12.4	-644.8	D	
431.5						44737
	8	28	-36.12	-1011.4	D	
403.5						43725
	9	15	70.18	1052.7	S	
388.5						44778
	10	56	-300.1	-16807	D	
332.5						27971
	11	142	70.18	9965.56	S	
190.5						37937
	12	3	83.78	251.34	S	
187.5						38188
	13	38	-23.72	-901.36	D	
149.5						37287
	14	26	-131.2	-3411.7	D	
123.5						33875
	15	68.1	-23.72	-1615.3	D	
55.4						32260

Table 6-4: Problem Table Algorithm for Modified Data

Hence, minimum cold utility requirement = 32260 kW

6.4.3 Balanced Combine Composite Curve

For hot balance composite curve, following tables are used.

Process Stream	**T_S (°C)**	**T_T (°C)**	**C_P (kW/°C)**	**h (W/m^2 °C)**
Primary Reformer Flue Gas	990	626	91.9	200
Boiler Flue Gas	805	497	26.2	200
Combine Flue Gas	597	201	107.5	338.3

Table 6-5: Modified Hot Streams Data

Actual Temperature (°C)	**Interval**	**$T_{(i+1)}$-T_i (°C)**	**$\Sigma C_{P\ Hot}$ (kW/°C)**	**Heat Load (kW)**	**Total Heat Flow (kW)**
201					0
497	1	296	107.5	31820	31820
597	2	100	133.7	13370	45190
626	3	29	26.2	759.8	45949.8
805	4	179	118.1	21140	67089.8
990	5	185	91.9	170002	84091.2

Table 6-6: Data for Hot Balance Composite Curve

For cold balance composite curve,

Process Stream	**T_S (°C)**	**T_T (°C)**	**C_P (kW/ °C)**	**h (W/m² °C)**
Natural & Recycle Gas	41.9	418	23.72	116.4
Process Air	177	492	13.6	115
Process Gas	390	532	106.3	170.8
Steam from Steam Drum	319	375	370.3	70.0
Boiler Feed Water	110	136	107.5	5721
HP Steam Generation	319	320	32259.7	6000

Table 6-7: Modified Cold Streams Data

Actual Temperature (°C)	**Interval**	**$T_{(i+1)}$-T_i (°C)**	**$\Sigma C_{P\ Cold}$ (kW/°C)**	**Heat Load (kW)**	**Total Heat Flow (kW)**
41.9					0
110	1	68.1	23.72	1615.332	1615.3
136	2	26	131.22	3411.72	5027.052
177	3	41	23.72	972.52	5999.572
319	4	142	37.32	5299.44	11299.01
320	5	1	26935.6	32667.32	43966.33
375	6	55	407.6	22419.1	66385.43
390	7	15	37.32	559.8	66945.23
418	8	28	143.6	4021.36	70966.59
492	9	74	119.9	8872.6	79839.19
532	10	40	106.3	4252	84091.19

Table 6-8: Data for Cold Balance Composite Curve

Now plot temperature against heat flow to obtain balance combine composite curve.

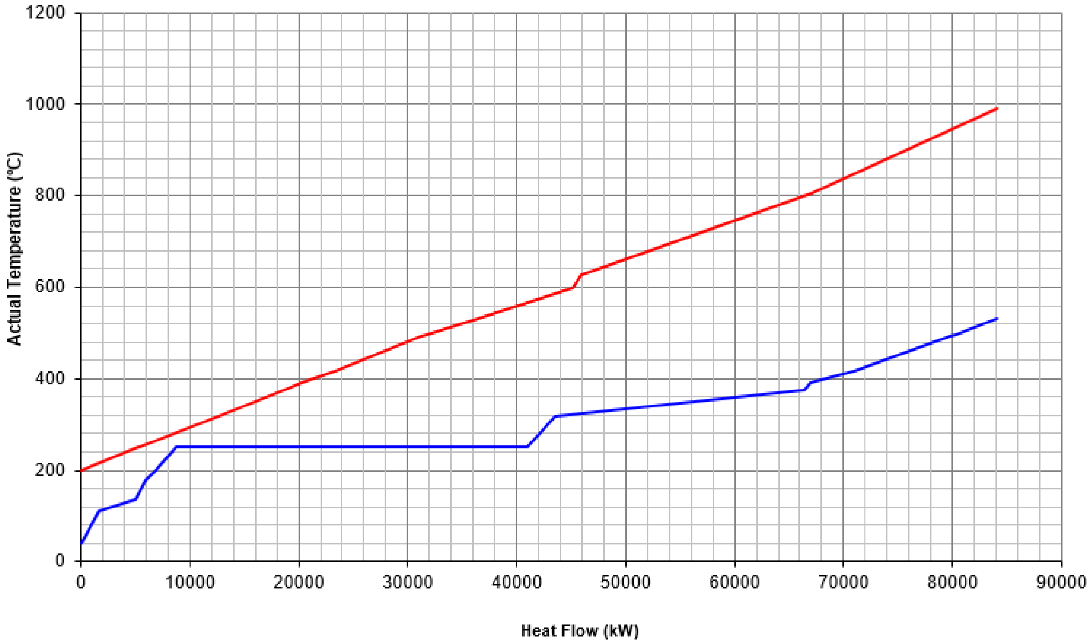


Figure 6.1: Balance Combine Composite Curve

Bath algorithm is used to determine minimum heat transfer area for the heat exchanger network.

6.4.4 Bath Algorithm

In bath algorithm, first the values of total heat flow for both hot and cold streams are arrange in ascending order and any repeated value is omitted. Corresponding value of temperatures for hot and cold composite curve are calculated by using following relation if it is unknown for the enthalpy interval.

For hot curve,

$$T_{H,row\ q} = T_{H,row\ r} - (\Delta H_{row\ r} - \Delta H_{row\ q})/\Sigma C_{P\ H,row\ r}$$

Where

$T_{H,row\ q}$ = Unknown value of hot curve temperature in row q

$T_{H,\ row}$ r = Known value of hot curve temperature in row r

$\Delta H_{row\ r}$ = Heat load in row r

$\Delta H_{row\ q}$ = Heat load in row q

$\Sigma C_{P\ H,\ row\ r}$ = Summation of hot streams heat capacity in row r

For cold curve,

$$T_{C,row\ q} = T_{C,row\ r} - (\Delta H_{row\ r} - \Delta H_{row\ q})/\Sigma C_{P\ C,row\ r}$$

Where

$T_{C,row\ q}$ = Unknown value of cold curve temperature in row q

$T_{C,\ row}$ r = Known value of cold curve temperature in row r

$\Delta H_{row\ r}$ = Heat load in row r

$\Delta H_{row\ q}$ = Heat load in row q

$\Sigma C_{P\ C,\ row\ r}$ = Summation of cold streams heat capacity in row r

Then compute Σ $(C_P/h)_h$ and Σ $(C_P/h)_c$ for each enthalpy interval. After that calculate Σ $(Q/h)_n$ for each interval using:

$$\Sigma(Q/h)_n = \Delta T_{H,n} \times \Sigma(C_P/h)_h + \Delta T_{C,n} \times \Sigma(C_P/h)_C$$

Calculate log mean temperature different (ΔT_{LM}) for each interval and divide Σ $(Q/h)_n$ with it to calculate the area for these intervals. Sum of the areas of all these intervals will give the total area required for heat exchange.

In	**Q_n (kW)**	**T_{Hn} (°C)**	**T_{Cn} (°C)**	**$\Sigma C_{P\ H/C}$**	**$\Sigma(C_P/h)$ Hot**	**$\Sigma(C_P/h)$ Cold**	**$\Sigma(Q/h)_n$**	**ΔT_{lm} (°C)**	**A_n (m^2)**
	0	201	41.9	0	0	0	0	0	0
1	1615	216	110	23.72	317.7	203.7	18652	130	142
2	5027	247	136	131.2	317.7	222.5	15871	108	145
3	5999	303	177	23.72	317.7	203.7	26186	119	219
4	11299	343.5	319	37.32	317.7	322	58325	62	936
5	31820	497	319.5	107.5	317.7	10988	54805	77	709.3
6	43966	587.8	320	26935	448.7	10988	45724	219	208.2
7	45190	597	323	133.7	448.7	5612	20868	270	77
8	45949	626	3245	26.2	131	5612	14262	287	49.6
9	66385	800	375	407.5	590.5	5612	384240	359	1068
10	66945	804	390	37.32	590.5	322	7132	419	17
11	67089	805	394	143.6	590.5	944	4249	412	10.3
12	70966	866	418	118.1	459.5	944	51061	429	118.8
13	79839	950	492	119.9	459.5	740	93159	453	205.5
14	84091	990	532	106.3	459.5	622	43274	458	94.5

Table 6-9: Area Targeting

Total area is 4004 m^2.

6.3.4 Area Comparison

Area of current installed coils and newly designed coils is as followed:

Coils	**Area (m^2)**
E 201	336
E 202 A	84.3
E 203	629
E 204 A	184.2
F 202 A	587
E 205	940
E 202 B	251.3
E204 B	433.63
E 206	213.8
Total Area	**3659.23**

Table 6-10: Area of Coils

$$Area\ Target\ Achieved = \frac{3660}{4004} \times 100\% = 91.5\%$$

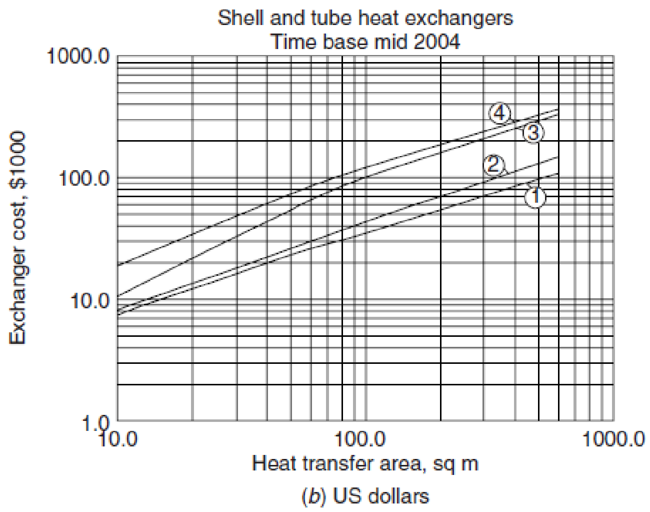
Chapter 7
Economic Analysis

7.1 Cost of Coils

Coils are made up of carbon steel. Cost of coils is determine by using following graphs and lag factors. Calculation for fixed capital is shown below:

Equipment	**Area (m^2)**	**Equipment Cost ($)**
E 202 A	84.29	30000
E 202 B	251.34	61000
E 204 A	184.24	49000
E 204 B	433.63	85000
E 206	213.76	52000
Purchas Cost of Equipment	**PCE ($)**	**277,000**
f_1 Equipment erection	0.4	
f_2 Piping	0.7	
f_3 Instrumentation	0.2	
f_4 Electrical	0.1	
f_5 Buildings	none required	
f_6 Utilities	not applicable	
f_7 Storages	not applicable	
f_8 Site development	not applicable	
f_9 Ancillary buildings	none required	
Total Physical Plant Cost	**PPC ($)**	**664,800**
f_{10} Design and Engineering	0.3	
f_{11} Contractor’s Fee	none	
f_{12} Contingencies	0.1	
	Fixed Capital ($)	**930,720**

Table 7-1: Fixed Capital ($)



Materials	
Shell	Tubes
① Carbon steel	Carbon steel
② C.S.	Brass
③ C.S.	Stainless steel
④ S.S.	S.S.

Pressure factors	
1–10 bar	× 1.0
10–20	× 1.1
20–30	× 1.25
30–50	× 1.3
50–70	× 1.5

Type factors	
Floating head	× 1.0
Fixed tube sheet	× 0.8
U tube	× 0.85
Kettle	× 1.3

Figure 6.3*a*, *b*. Shell and tube heat exchangers. Time base mid-2004
Purchased cost = (bare cost from figure) × Type factor × Pressure factor

Figure 7.1: Cost to Heat Transfer Area Graph

7.2 Economic Analysis

Following tables shows the cost saving due to extra production of steam and payback period of project by introducing these new coils.

Total Area of Coils	1167.26 m^2
Capital Cost of Coils	$ 930,720
Surplus HP Steam Generated	4430 kg/hr
Cost Saved Per Annum by Surplus HP Steam	$ 4.51×10^5 per year
Pay Back Period	2.06 years

Table 7-2: Economic Analysis

*Cost of steam is approximately $12 per ton.

So, the payback period of this project is 2.06 years so it is feasible to apply these modifications.

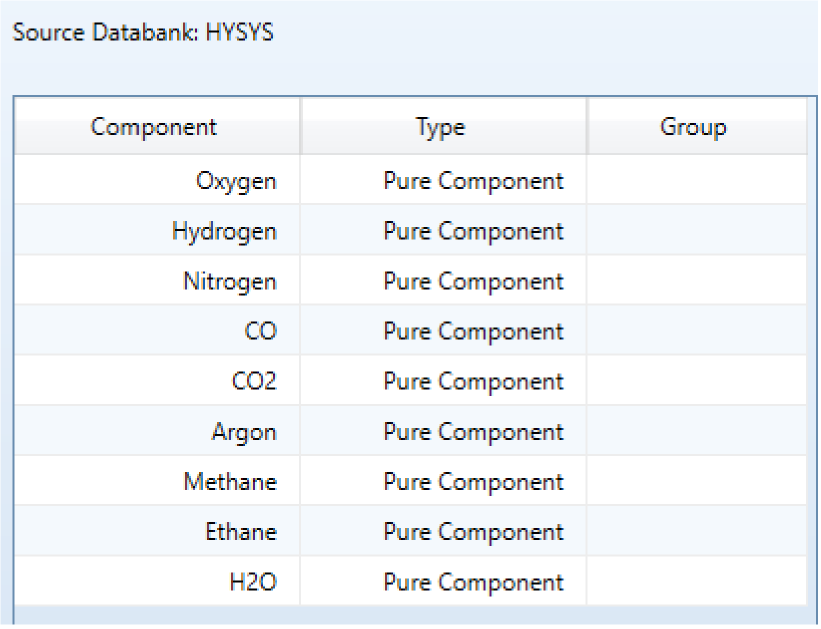
Chapter 8
Aspen HYSYS Simulation

For the verification of proposed design, it was simulated on Aspen HYSYS. Detailed simulation of the process is as followed.

8.1 HYSYS Simulation

8.1.1 Components

Following list of components was selected from HYSYS databank.



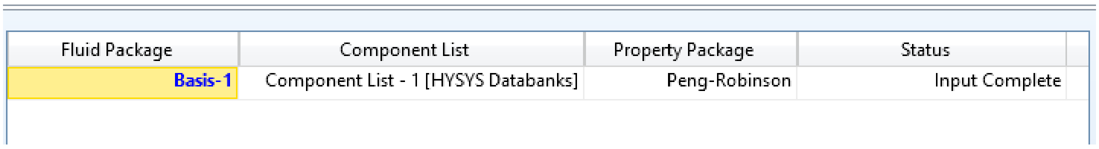
Source Databank: HYSYS

Component	Type	Group
Oxygen	Pure Component	
Hydrogen	Pure Component	
Nitrogen	Pure Component	
CO	Pure Component	
CO2	Pure Component	
Argon	Pure Component	
Methane	Pure Component	
Ethane	Pure Component	
H2O	Pure Component	

Figure 8.1: Components

8.1.2 Fluid Package

Peng Robinson is selected as fluid package for simulation as only non-polar gases area involve.



Fluid Package	Component List	Property Package	Status
Basis-1	Component List - 1 [HYSYS Databanks]	Peng-Robinson	Input Complete

Figure 8.2: Fluid Package

8.1.3 Model

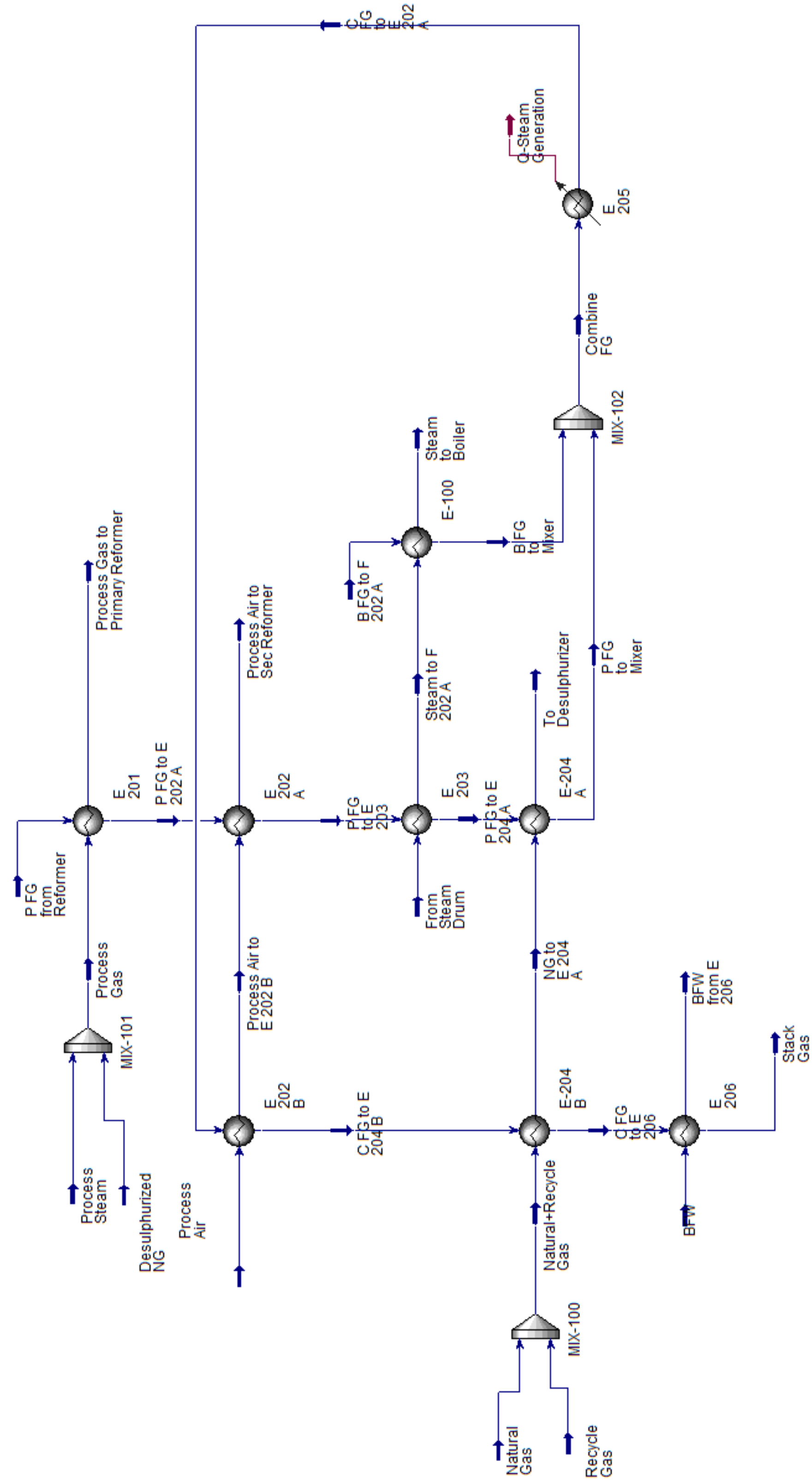
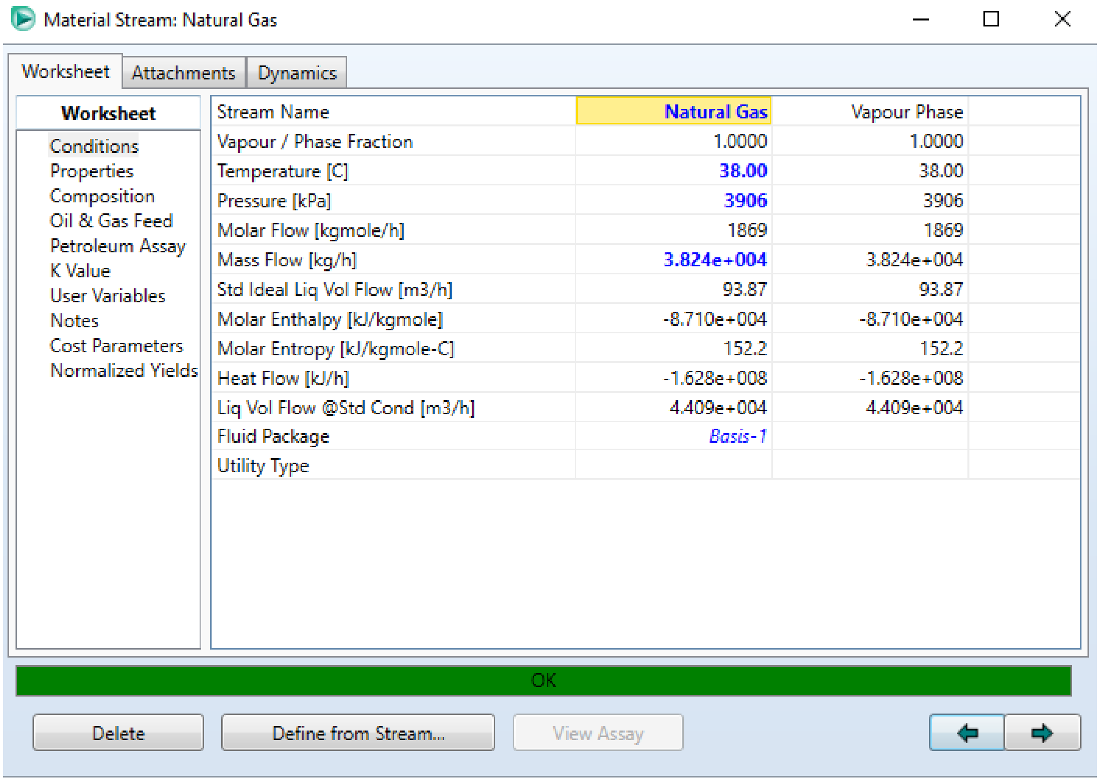


Figure 8.3: Simulation Model

8.1.4 Defined Streams

Following streams were defined on HYSYS.

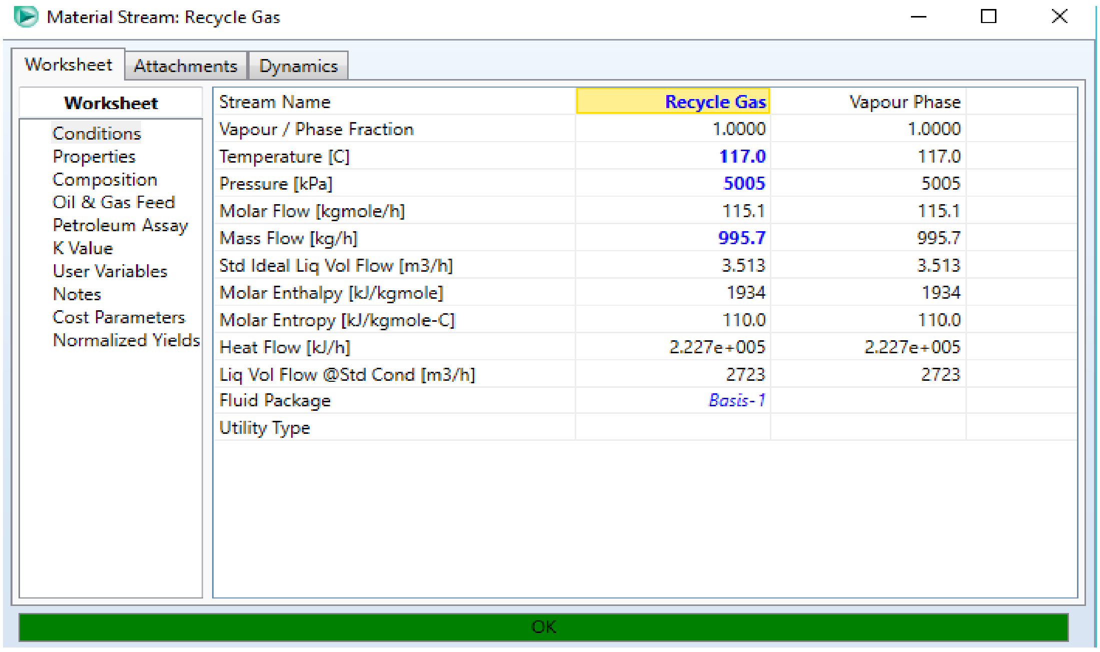
Natural Gas



Stream Name	Natural Gas	Vapour Phase
Vapour / Phase Fraction	1.0000	1.0000
Temperature [C]	38.00	38.00
Pressure [kPa]	3906	3906
Molar Flow [kgmole/h]	1869	1869
Mass Flow [kg/h]	3.824e+004	3.824e+004
Std Ideal Liq Vol Flow [m3/h]	93.87	93.87
Molar Enthalpy [kJ/kgmole]	-8.710e+004	-8.710e+004
Molar Entropy [kJ/kgmole-C]	152.2	152.2
Heat Flow [kJ/h]	-1.628e+008	-1.628e+008
Liq Vol Flow @Std Cond [m3/h]	4.409e+004	4.409e+004
Fluid Package	Basis-1	
Utility Type		

Figure 8.4: Natural Gas

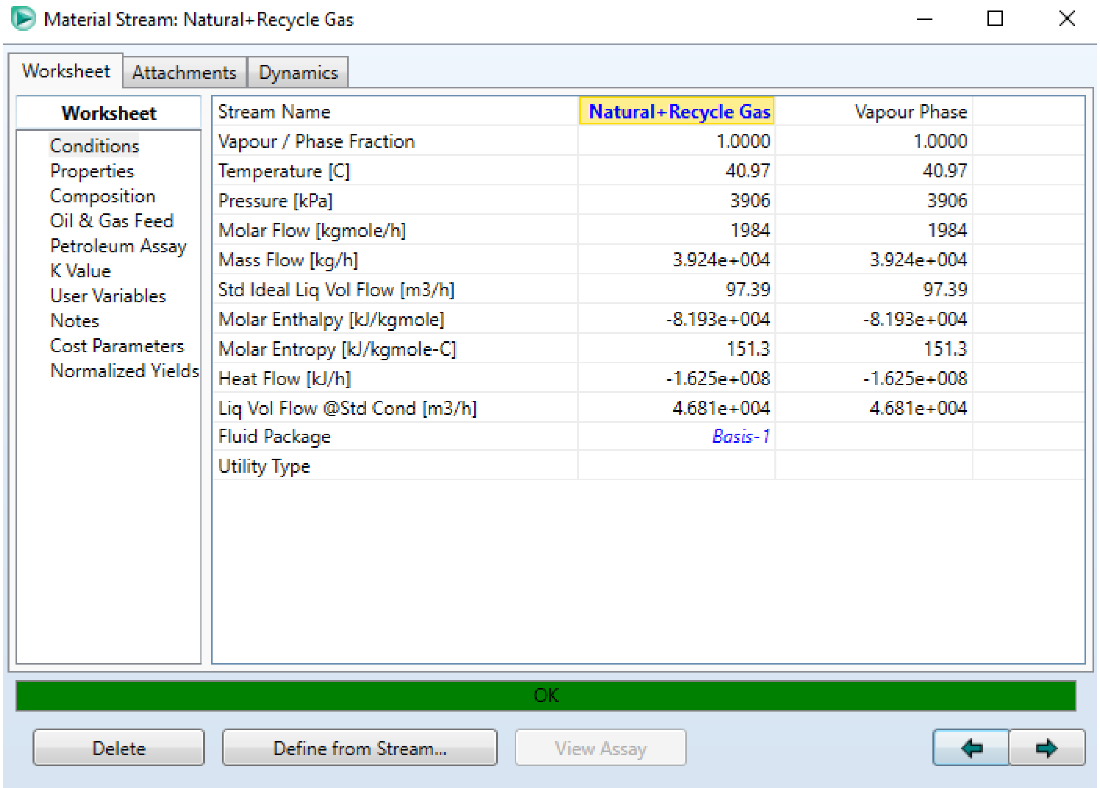
Recycle Gas



Stream Name	Recycle Gas	Vapour Phase
Vapour / Phase Fraction	1.0000	1.0000
Temperature [C]	117.0	117.0
Pressure [kPa]	5005	5005
Molar Flow [kgmole/h]	115.1	115.1
Mass Flow [kg/h]	995.7	995.7
Std Ideal Liq Vol Flow [m3/h]	3.513	3.513
Molar Enthalpy [kJ/kgmole]	1934	1934
Molar Entropy [kJ/kgmole-C]	110.0	110.0
Heat Flow [kJ/h]	2.227e+005	2.227e+005
Liq Vol Flow @Std Cond [m3/h]	2723	2723
Fluid Package	Basis-1	
Utility Type		

Figure 8.5: Recycle Gas

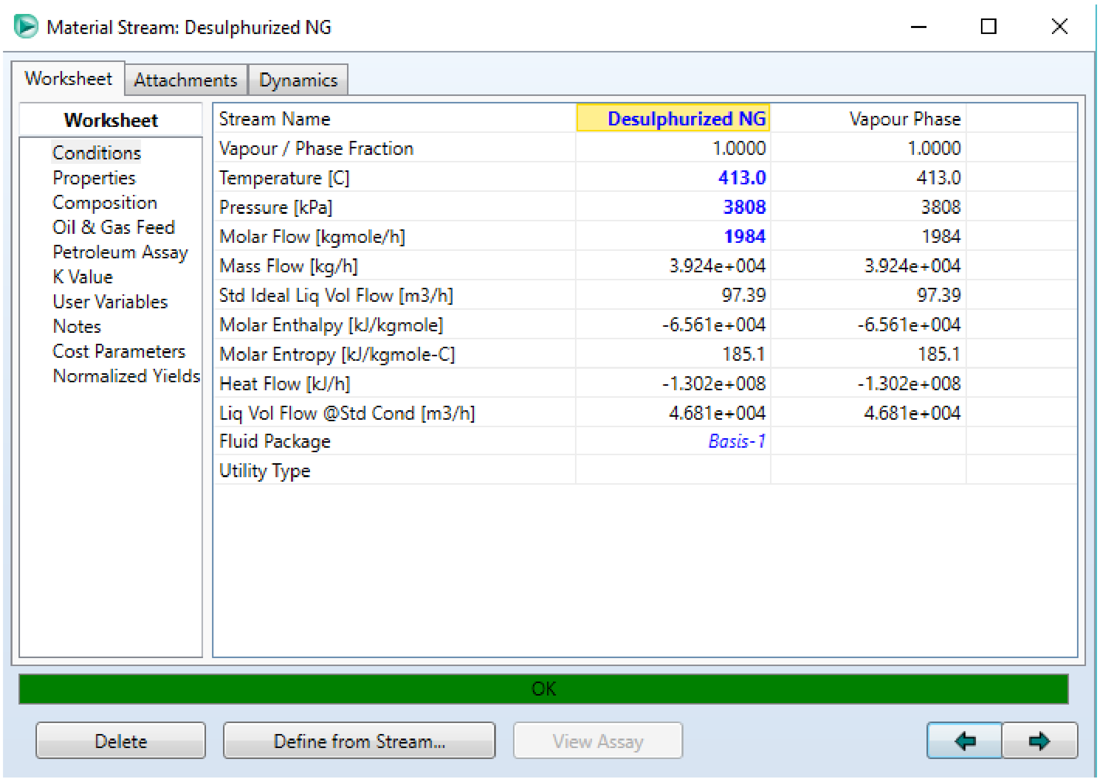
Natural & Recycle Gas



Stream Name	Natural+Recycle Gas	Vapour Phase
Vapour / Phase Fraction	1.0000	1.0000
Temperature [C]	40.97	40.97
Pressure [kPa]	3906	3906
Molar Flow [kgmole/h]	1984	1984
Mass Flow [kg/h]	3.924e+004	3.924e+004
Std Ideal Liq Vol Flow [m3/h]	97.39	97.39
Molar Enthalpy [kJ/kgmole]	-8.193e+004	-8.193e+004
Molar Entropy [kJ/kgmole-C]	151.3	151.3
Heat Flow [kJ/h]	-1.625e+008	-1.625e+008
Liq Vol Flow @Std Cond [m3/h]	4.681e+004	4.681e+004
Fluid Package	*Basis-1*	
Utility Type		

Figure 8.6: Natural & Recycle Gas

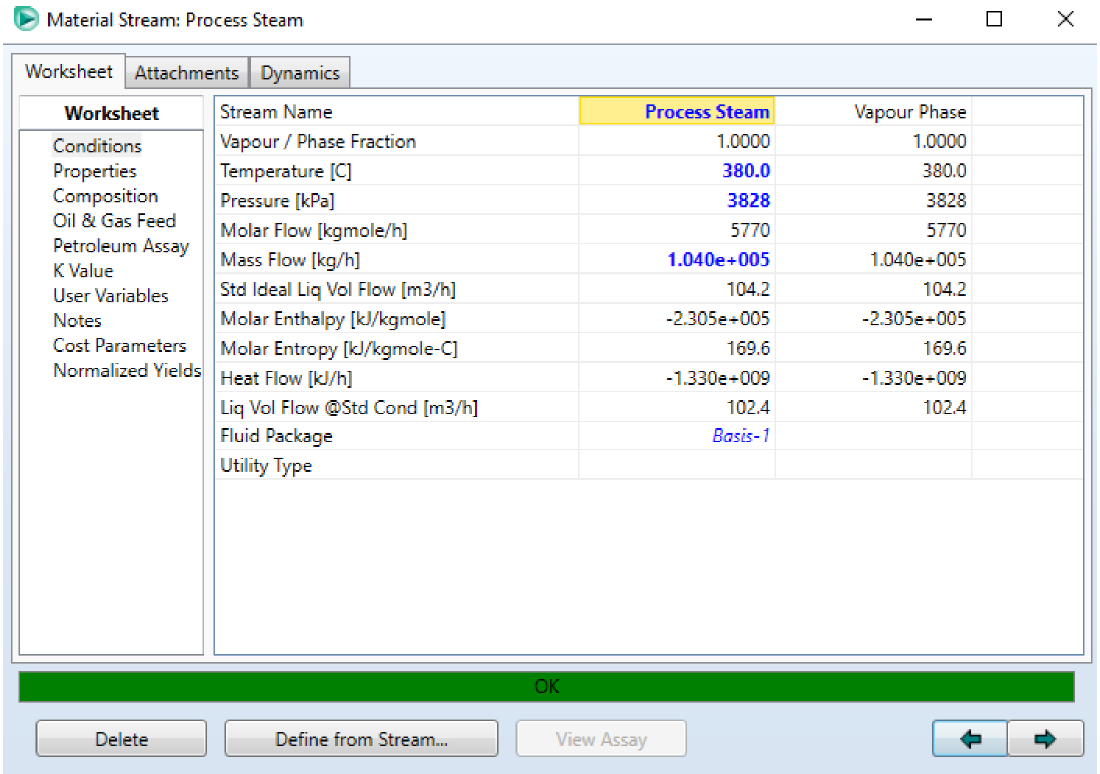
Desulphurized Natural Gas



Stream Name	Desulphurized NG	Vapour Phase
Vapour / Phase Fraction	1.0000	1.0000
Temperature [C]	**413.0**	413.0
Pressure [kPa]	**3808**	3808
Molar Flow [kgmole/h]	**1984**	1984
Mass Flow [kg/h]	3.924e+004	3.924e+004
Std Ideal Liq Vol Flow [m3/h]	97.39	97.39
Molar Enthalpy [kJ/kgmole]	-6.561e+004	-6.561e+004
Molar Entropy [kJ/kgmole-C]	185.1	185.1
Heat Flow [kJ/h]	-1.302e+008	-1.302e+008
Liq Vol Flow @Std Cond [m3/h]	4.681e+004	4.681e+004
Fluid Package	*Basis-1*	
Utility Type		

Figure 8.7: Desulphurized Natural Gas

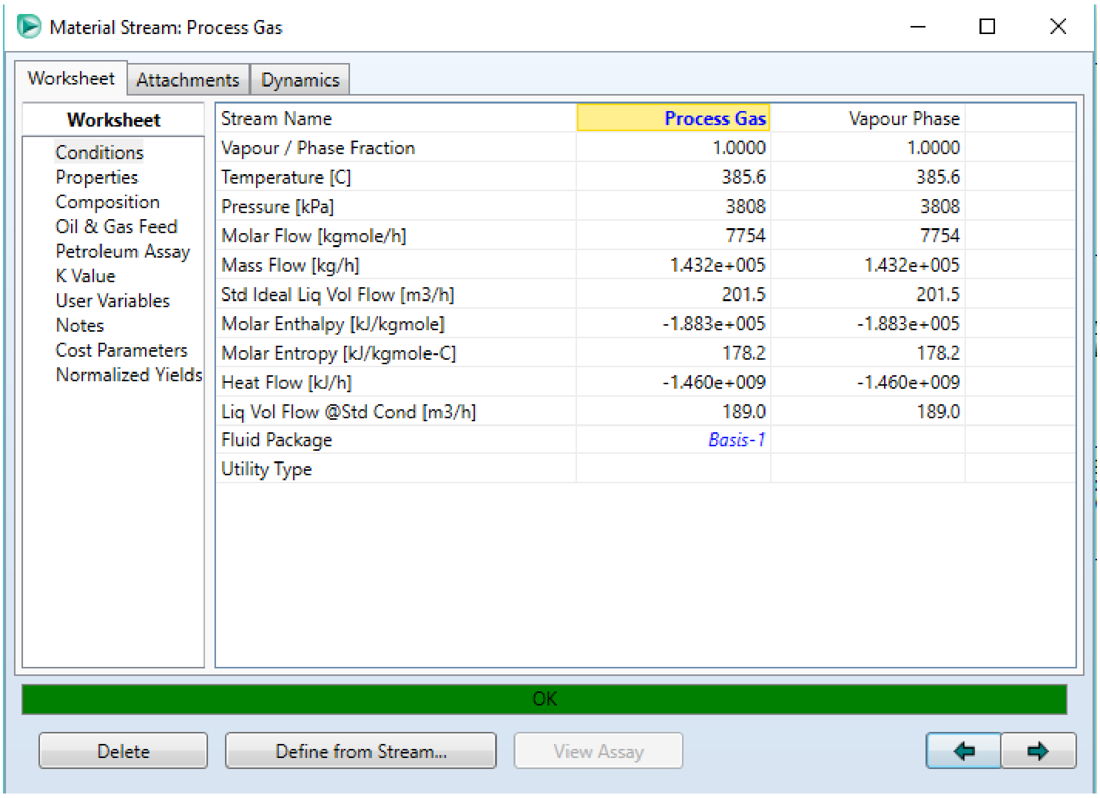
Process Steam



Stream Name	Process Steam	Vapour Phase
Vapour / Phase Fraction	1.0000	1.0000
Temperature [C]	380.0	380.0
Pressure [kPa]	3828	3828
Molar Flow [kgmole/h]	5770	5770
Mass Flow [kg/h]	1.040e+005	1.040e+005
Std Ideal Liq Vol Flow [m3/h]	104.2	104.2
Molar Enthalpy [kJ/kgmole]	-2.305e+005	-2.305e+005
Molar Entropy [kJ/kgmole-C]	169.6	169.6
Heat Flow [kJ/h]	-1.330e+009	-1.330e+009
Liq Vol Flow @Std Cond [m3/h]	102.4	102.4
Fluid Package	Basis-1	
Utility Type		

Figure 8.8: Process Steam

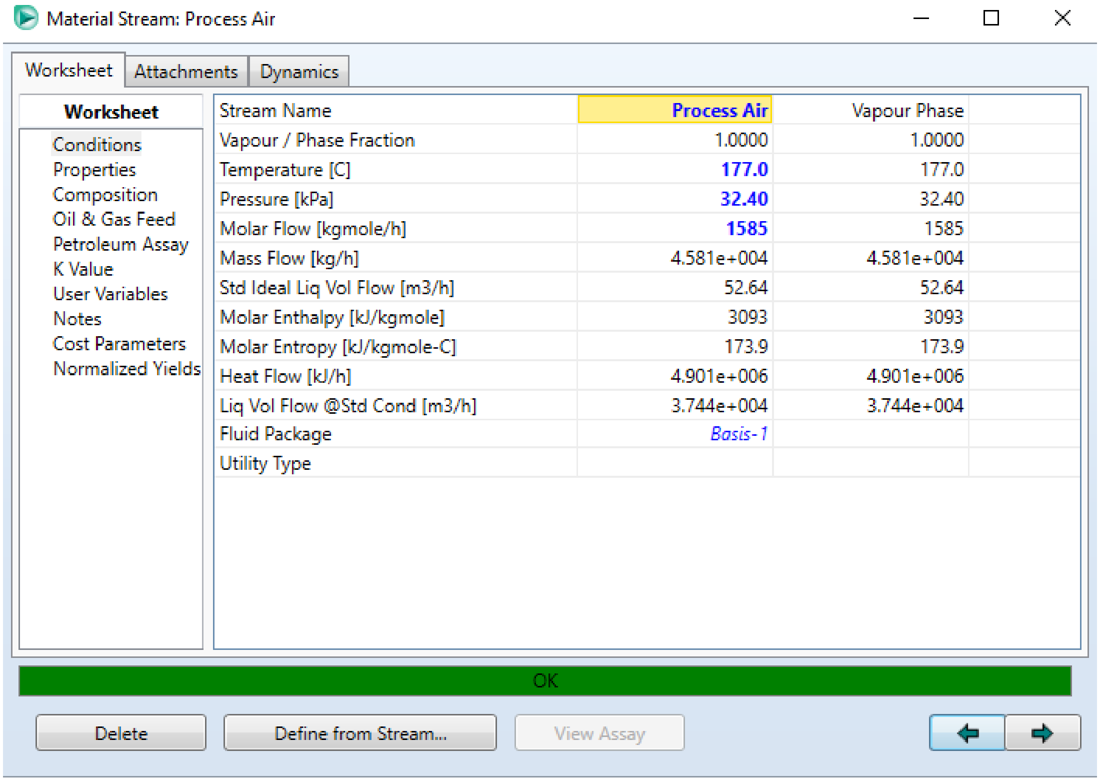
Process Gas



Stream Name	Process Gas	Vapour Phase
Vapour / Phase Fraction	1.0000	1.0000
Temperature [C]	385.6	385.6
Pressure [kPa]	3808	3808
Molar Flow [kgmole/h]	7754	7754
Mass Flow [kg/h]	1.432e+005	1.432e+005
Std Ideal Liq Vol Flow [m3/h]	201.5	201.5
Molar Enthalpy [kJ/kgmole]	-1.883e+005	-1.883e+005
Molar Entropy [kJ/kgmole-C]	178.2	178.2
Heat Flow [kJ/h]	-1.460e+009	-1.460e+009
Liq Vol Flow @Std Cond [m3/h]	189.0	189.0
Fluid Package	Basis-1	
Utility Type		

Figure 8.9: Process Gas

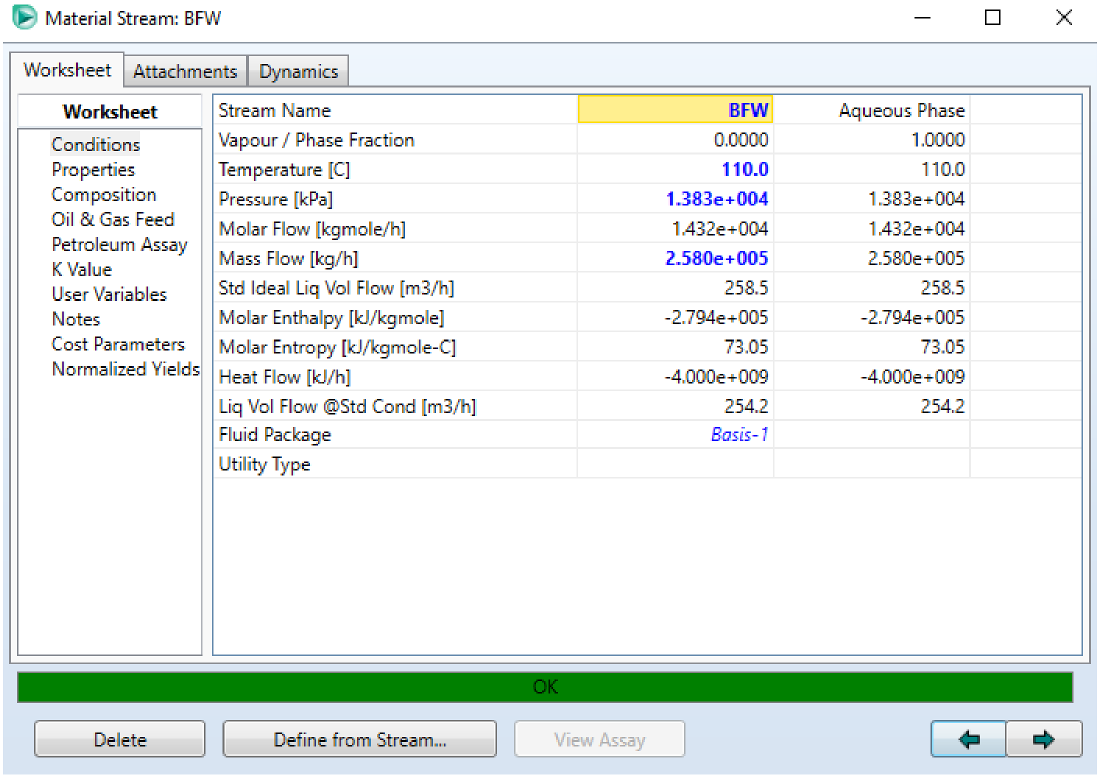
Process Air



Stream Name	Process Air	Vapour Phase
Vapour / Phase Fraction	1.0000	1.0000
Temperature [C]	177.0	177.0
Pressure [kPa]	32.40	32.40
Molar Flow [kgmole/h]	1585	1585
Mass Flow [kg/h]	4.581e+004	4.581e+004
Std Ideal Liq Vol Flow [m3/h]	52.64	52.64
Molar Enthalpy [kJ/kgmole]	3093	3093
Molar Entropy [kJ/kgmole-C]	173.9	173.9
Heat Flow [kJ/h]	4.901e+006	4.901e+006
Liq Vol Flow @Std Cond [m3/h]	3.744e+004	3.744e+004
Fluid Package	*Basis-1*	
Utility Type		

Figure 8.10: Process Air

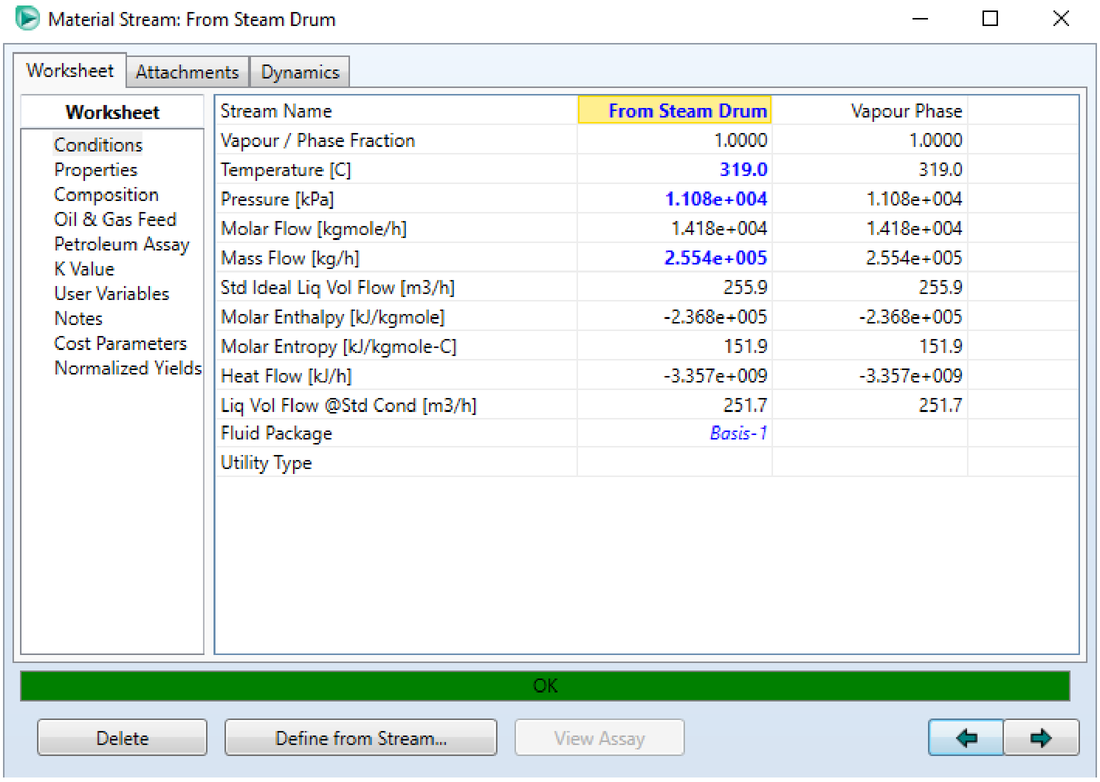
Boiler Feed Water



Stream Name	BFW	Aqueous Phase
Vapour / Phase Fraction	0.0000	1.0000
Temperature [C]	110.0	110.0
Pressure [kPa]	1.383e+004	1.383e+004
Molar Flow [kgmole/h]	1.432e+004	1.432e+004
Mass Flow [kg/h]	2.580e+005	2.580e+005
Std Ideal Liq Vol Flow [m3/h]	258.5	258.5
Molar Enthalpy [kJ/kgmole]	-2.794e+005	-2.794e+005
Molar Entropy [kJ/kgmole-C]	73.05	73.05
Heat Flow [kJ/h]	-4.000e+009	-4.000e+009
Liq Vol Flow @Std Cond [m3/h]	254.2	254.2
Fluid Package	*Basis-1*	
Utility Type		

Figure 8.11: Boiler Feed Water

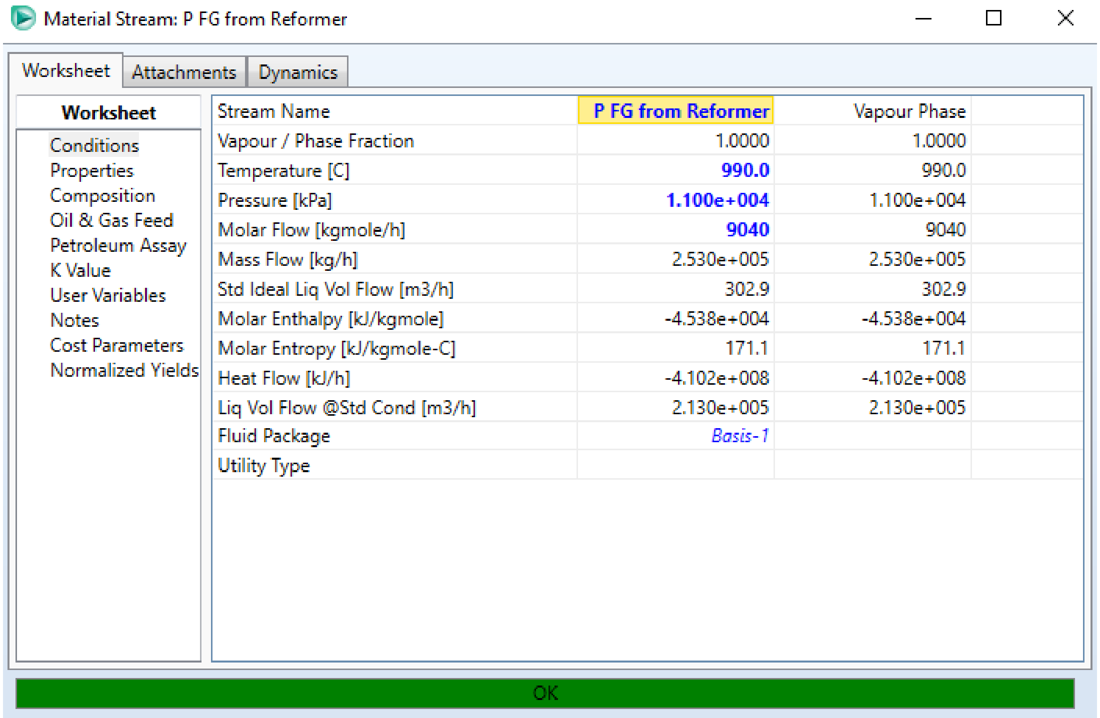
Steam from Steam Drum



Stream Name	From Steam Drum	Vapour Phase
Vapour / Phase Fraction	1.0000	1.0000
Temperature [C]	319.0	319.0
Pressure [kPa]	1.108e+004	1.108e+004
Molar Flow [kgmole/h]	1.418e+004	1.418e+004
Mass Flow [kg/h]	2.554e+005	2.554e+005
Std Ideal Liq Vol Flow [m3/h]	255.9	255.9
Molar Enthalpy [kJ/kgmole]	-2.368e+005	-2.368e+005
Molar Entropy [kJ/kgmole-C]	151.9	151.9
Heat Flow [kJ/h]	-3.357e+009	-3.357e+009
Liq Vol Flow @Std Cond [m3/h]	251.7	251.7
Fluid Package	*Basis-1*	
Utility Type		

Figure 8.12: Steam from Steam Drum

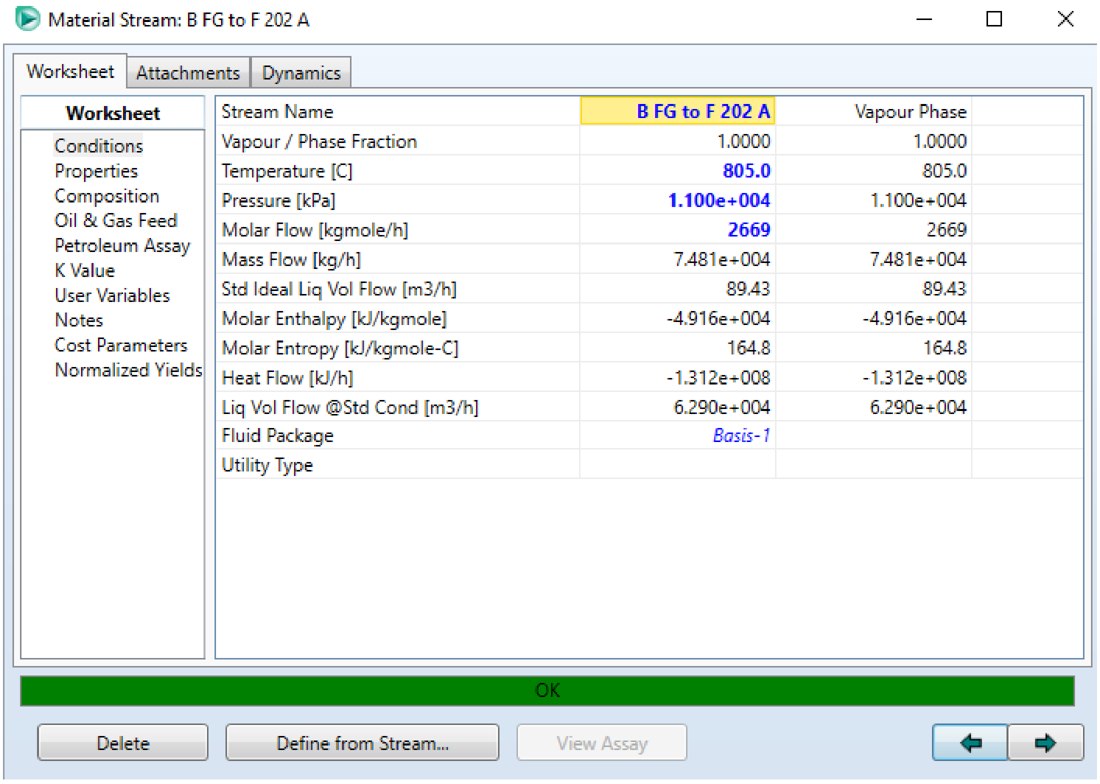
Primary Reformer Flue Gas



Stream Name	P FG from Reformer	Vapour Phase
Vapour / Phase Fraction	1.0000	1.0000
Temperature [C]	990.0	990.0
Pressure [kPa]	1.100e+004	1.100e+004
Molar Flow [kgmole/h]	9040	9040
Mass Flow [kg/h]	2.530e+005	2.530e+005
Std Ideal Liq Vol Flow [m3/h]	302.9	302.9
Molar Enthalpy [kJ/kgmole]	-4.538e+004	-4.538e+004
Molar Entropy [kJ/kgmole-C]	171.1	171.1
Heat Flow [kJ/h]	-4.102e+008	-4.102e+008
Liq Vol Flow @Std Cond [m3/h]	2.130e+005	2.130e+005
Fluid Package	*Basis-1*	
Utility Type		

Figure 8.13: Primary Reformer Flue Gas

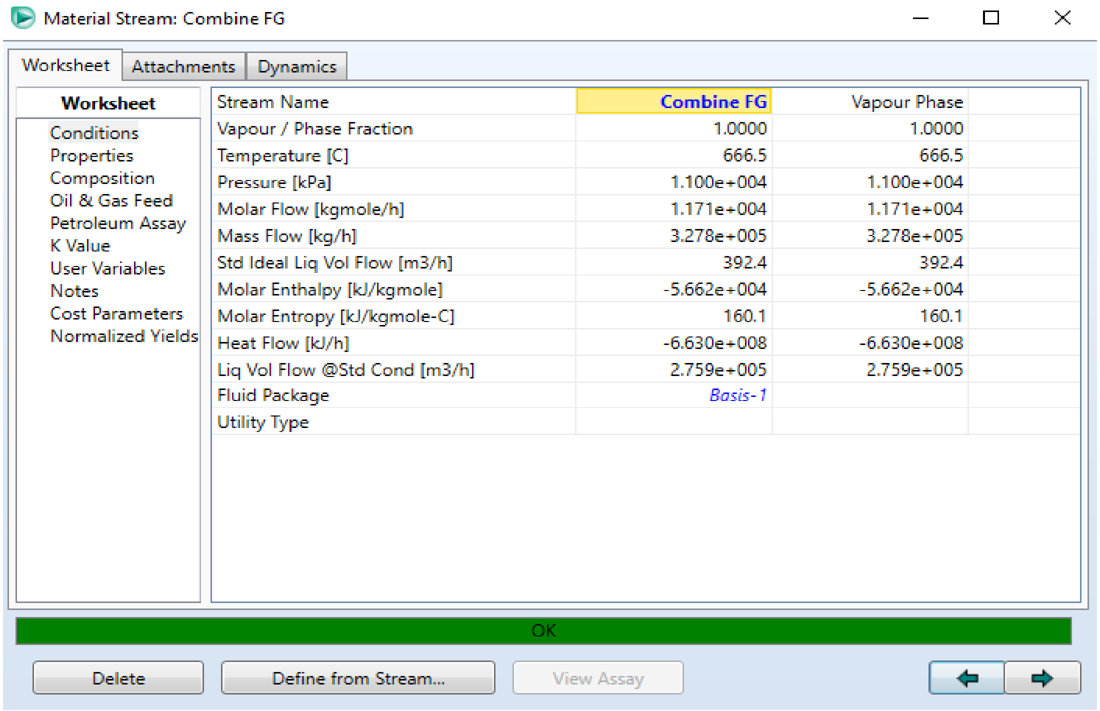
Boiler Flue Gas



Stream Name	B FG to F 202 A	Vapour Phase
Vapour / Phase Fraction	1.0000	1.0000
Temperature [C]	805.0	805.0
Pressure [kPa]	1.100e+004	1.100e+004
Molar Flow [kgmole/h]	2669	2669
Mass Flow [kg/h]	7.481e+004	7.481e+004
Std Ideal Liq Vol Flow [m3/h]	89.43	89.43
Molar Enthalpy [kJ/kgmole]	-4.916e+004	-4.916e+004
Molar Entropy [kJ/kgmole-C]	164.8	164.8
Heat Flow [kJ/h]	-1.312e+008	-1.312e+008
Liq Vol Flow @Std Cond [m3/h]	6.290e+004	6.290e+004
Fluid Package	*Basis-1*	
Utility Type		

Figure 8.14: Boiler Flue Gas

Combine Flue Gas



Stream Name	Combine FG	Vapour Phase
Vapour / Phase Fraction	1.0000	1.0000
Temperature [C]	666.5	666.5
Pressure [kPa]	1.100e+004	1.100e+004
Molar Flow [kgmole/h]	1.171e+004	1.171e+004
Mass Flow [kg/h]	3.278e+005	3.278e+005
Std Ideal Liq Vol Flow [m3/h]	392.4	392.4
Molar Enthalpy [kJ/kgmole]	-5.662e+004	-5.662e+004
Molar Entropy [kJ/kgmole-C]	160.1	160.1
Heat Flow [kJ/h]	-6.630e+008	-6.630e+008
Liq Vol Flow @Std Cond [m3/h]	2.759e+005	2.759e+005
Fluid Package	*Basis-1*	
Utility Type		

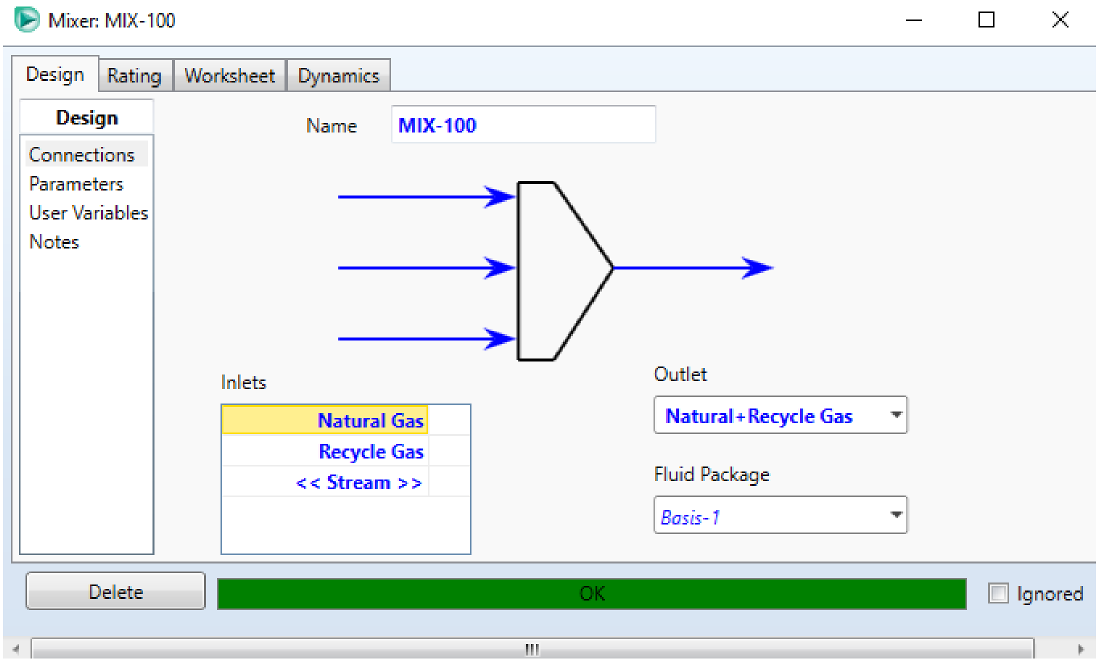
Figure 8.15: Combine Flue Gas

8.1.5 Mixers

Now mixers are simulated in HYSYS.

MIX-100

The Natural Gas and Recycle Gas are mixed together in MIX-100 to produce mixture of Natural and Recycle Gas.



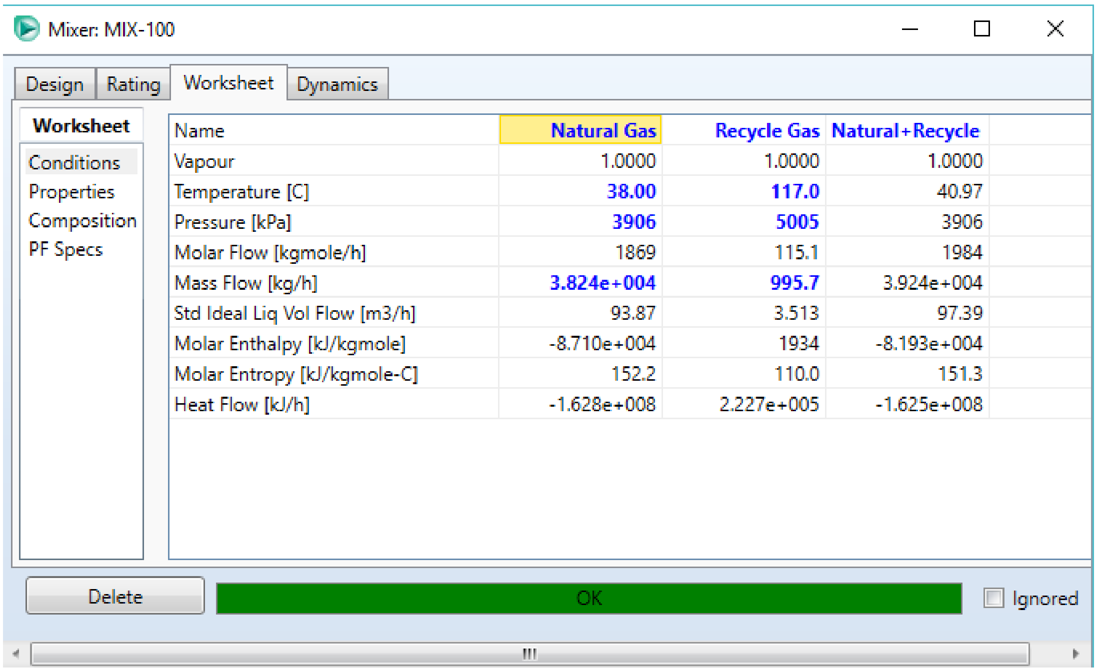
Name: MIX-100

Inlets:
- Natural Gas
- Recycle Gas
- << Stream >>

Outlet: Natural+Recycle Gas

Fluid Package: Basis-1

Figure 8.16: MIX-100 Design

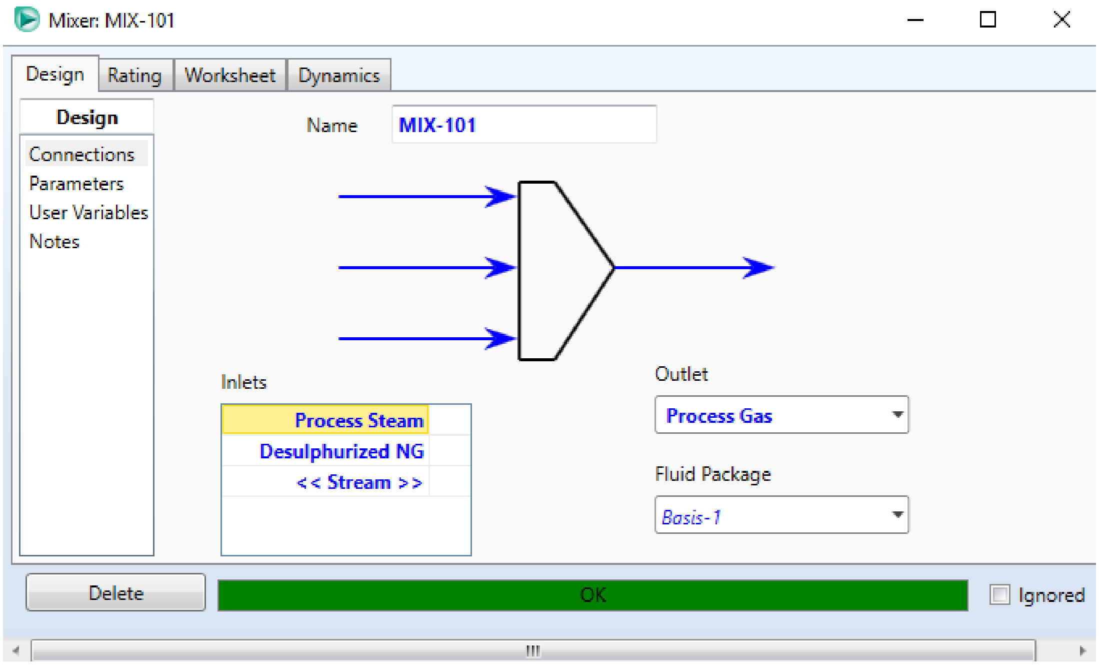


Name	Natural Gas	Recycle Gas	Natural+Recycle
Vapour	1.0000	1.0000	1.0000
Temperature [C]	38.00	117.0	40.97
Pressure [kPa]	3906	5005	3906
Molar Flow [kgmole/h]	1869	115.1	1984
Mass Flow [kg/h]	3.824e+004	995.7	3.924e+004
Std Ideal Liq Vol Flow [m3/h]	93.87	3.513	97.39
Molar Enthalpy [kJ/kgmole]	-8.710e+004	1934	-8.193e+004
Molar Entropy [kJ/kgmole-C]	152.2	110.0	151.3
Heat Flow [kJ/h]	-1.628e+008	2.227e+005	-1.625e+008

Figure 8.17: MIX-100 Worksheet

MIX-101

The streams; Process Steam and Desulphurized Natural Gas, are mixed together in MIX-101 to produce Process Gas.



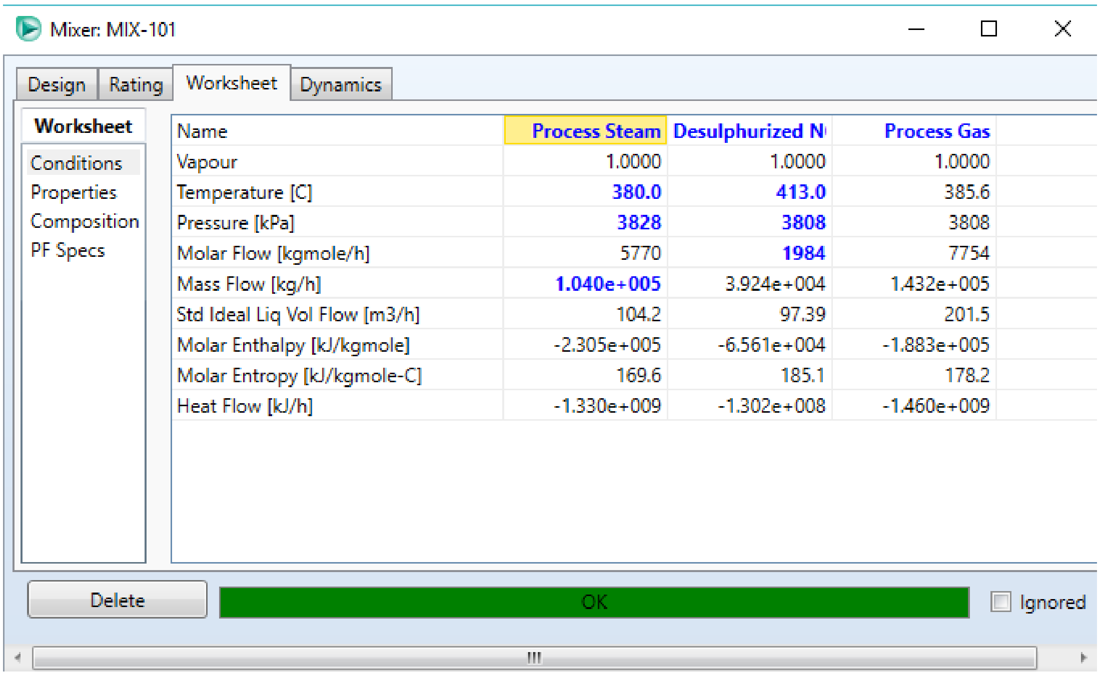
Name: MIX-101

Inlets:
- Process Steam
- Desulphurized NG
- << Stream >>

Outlet: Process Gas

Fluid Package: Basis-1

Figure 8.18 MIX-101 Design

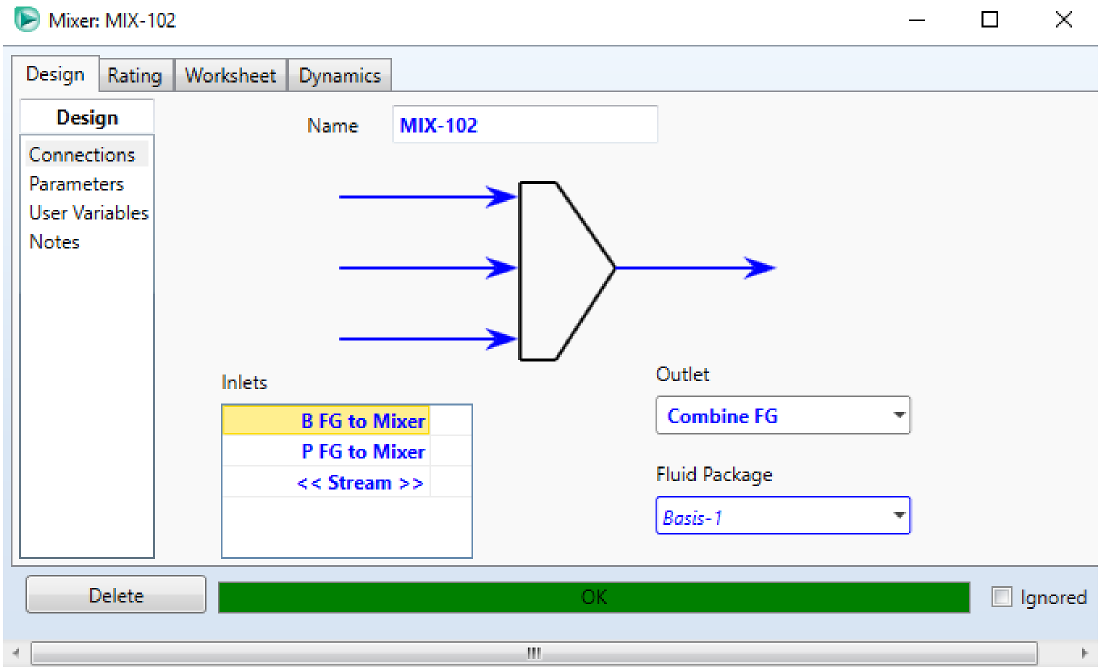


Name	Process Steam	Desulphurized N	Process Gas
Vapour	1.0000	1.0000	1.0000
Temperature [C]	380.0	413.0	385.6
Pressure [kPa]	3828	3808	3808
Molar Flow [kgmole/h]	5770	1984	7754
Mass Flow [kg/h]	1.040e+005	3.924e+004	1.432e+005
Std Ideal Liq Vol Flow [m3/h]	104.2	97.39	201.5
Molar Enthalpy [kJ/kgmole]	-2.305e+005	-6.561e+004	-1.883e+005
Molar Entropy [kJ/kgmole-C]	169.6	185.1	178.2
Heat Flow [kJ/h]	-1.330e+009	-1.302e+008	-1.460e+009

Figure 8.19: MIX-101 Worksheet

MIX-102

Boiler Flue Gas and Primary Flue Gas are mixed together in MIX-102 to produce Combine Flue Gas.



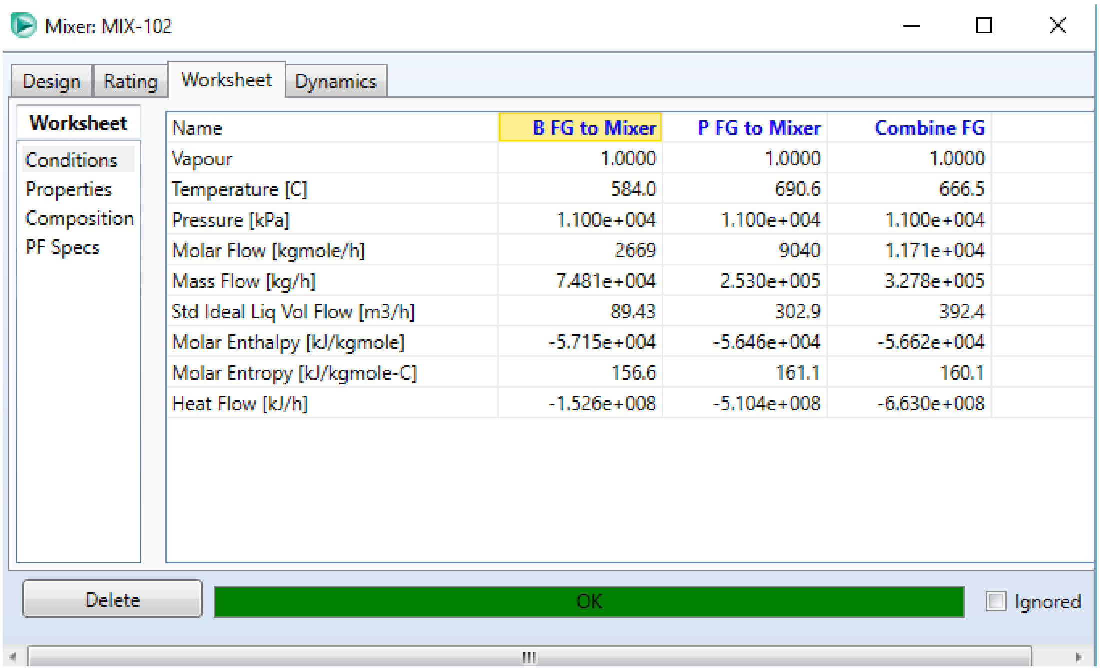
Name: MIX-102

Inlets:
- B FG to Mixer
- P FG to Mixer
- << Stream >>

Outlet: Combine FG

Fluid Package: Basis-1

Figure 8.20: MIX-102 Design



Name	B FG to Mixer	P FG to Mixer	Combine FG
Vapour	1.0000	1.0000	1.0000
Temperature [C]	584.0	690.6	666.5
Pressure [kPa]	1.100e+004	1.100e+004	1.100e+004
Molar Flow [kgmole/h]	2669	9040	1.171e+004
Mass Flow [kg/h]	7.481e+004	2.530e+005	3.278e+005
Std Ideal Liq Vol Flow [m3/h]	89.43	302.9	392.4
Molar Enthalpy [kJ/kgmole]	-5.715e+004	-5.646e+004	-5.662e+004
Molar Entropy [kJ/kgmole-C]	156.6	161.1	160.1
Heat Flow [kJ/h]	-1.526e+008	-5.104e+008	-6.630e+008

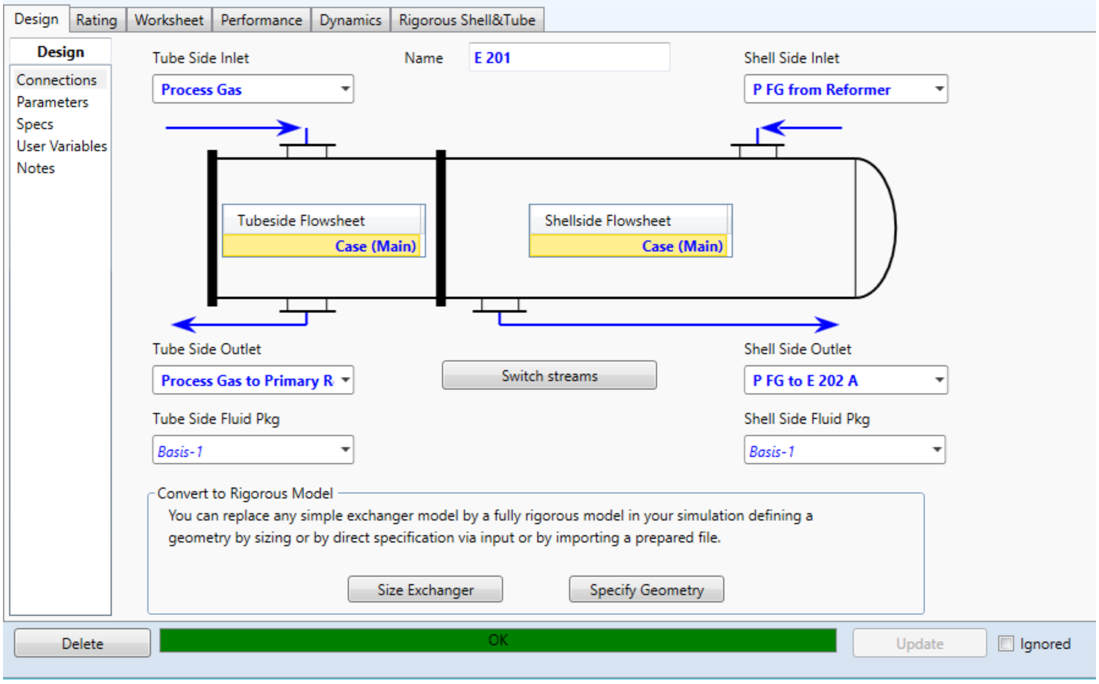
Figure 8.21: MIX-102 Worksheet

8.1.6 Heat Exchangers

Now exchangers are simulated in HYSYS.

Exchanger E 201

In this exchanger, Process Gas is heated using Primary Reformer Flue Gas.



Tube Side Inlet: Process Gas

Name: E 201

Shell Side Inlet: P FG from Reformer

Tube Side Outlet: Process Gas to Primary R

Shell Side Outlet: P FG to E 202 A

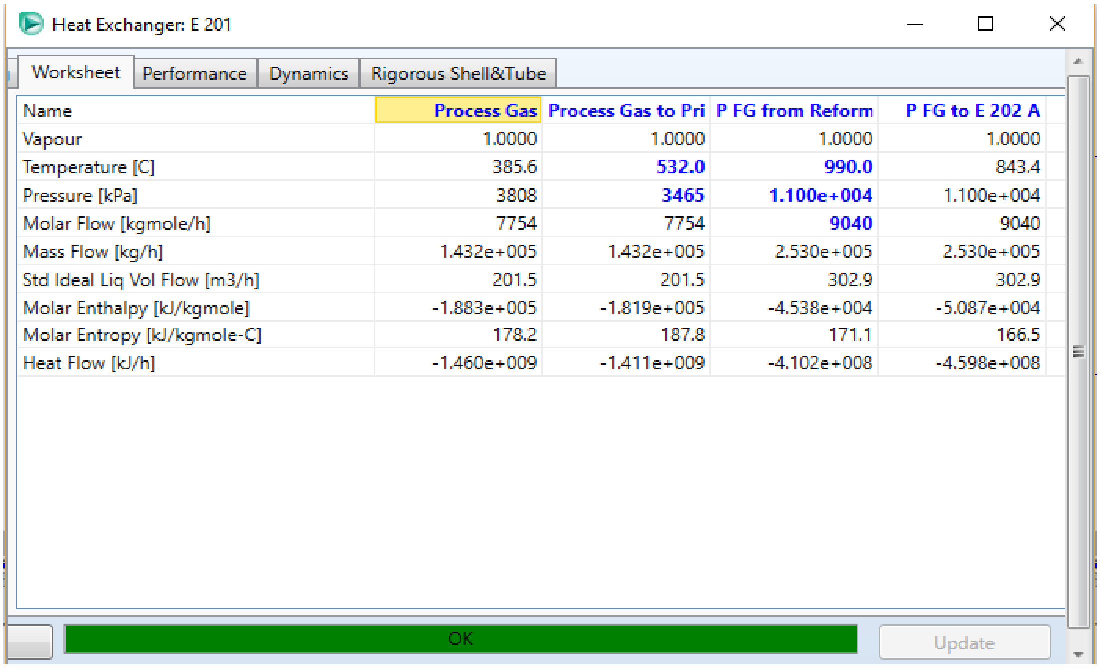
Tube Side Fluid Pkg: Basis-1

Shell Side Fluid Pkg: Basis-1

Convert to Rigorous Model

You can replace any simple exchanger model by a fully rigorous model in your simulation defining a geometry by sizing or by direct specification via input or by importing a prepared file.

Figure 8.22: Exchanger E 201 Design

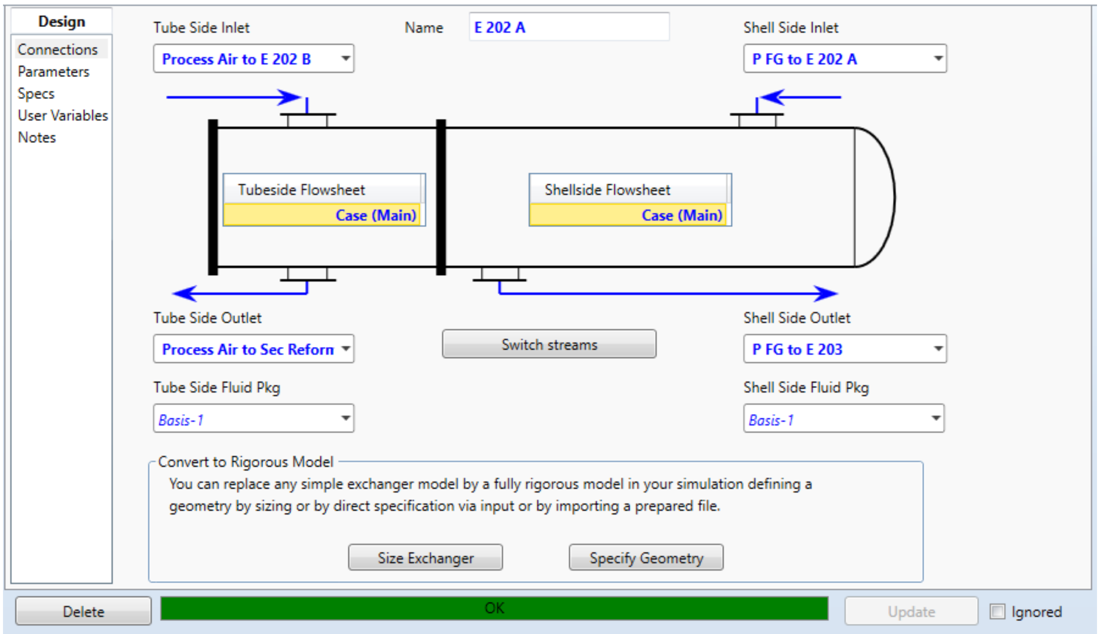


Name	Process Gas	Process Gas to Pri	P FG from Reform	P FG to E 202 A
Vapour	1.0000	1.0000	1.0000	1.0000
Temperature [C]	385.6	532.0	990.0	843.4
Pressure [kPa]	3808	3465	1.100e+004	1.100e+004
Molar Flow [kgmole/h]	7754	7754	9040	9040
Mass Flow [kg/h]	1.432e+005	1.432e+005	2.530e+005	2.530e+005
Std Ideal Liq Vol Flow [m3/h]	201.5	201.5	302.9	302.9
Molar Enthalpy [kJ/kgmole]	-1.883e+005	-1.819e+005	-4.538e+004	-5.087e+004
Molar Entropy [kJ/kgmole-C]	178.2	187.8	171.1	166.5
Heat Flow [kJ/h]	-1.460e+009	-1.411e+009	-4.102e+008	-4.598e+008

Figure 8.23: Exchanger E 201 Worksheet

Exchanger E 202 A

In this exchanger, Process Air is heated using Primary Reformer Flue Gas.



Tube Side Inlet: Process Air to E 202 B

Name: E 202 A

Shell Side Inlet: P FG to E 202 A

Tube Side Outlet: Process Air to Sec Reforn

Shell Side Outlet: P FG to E 203

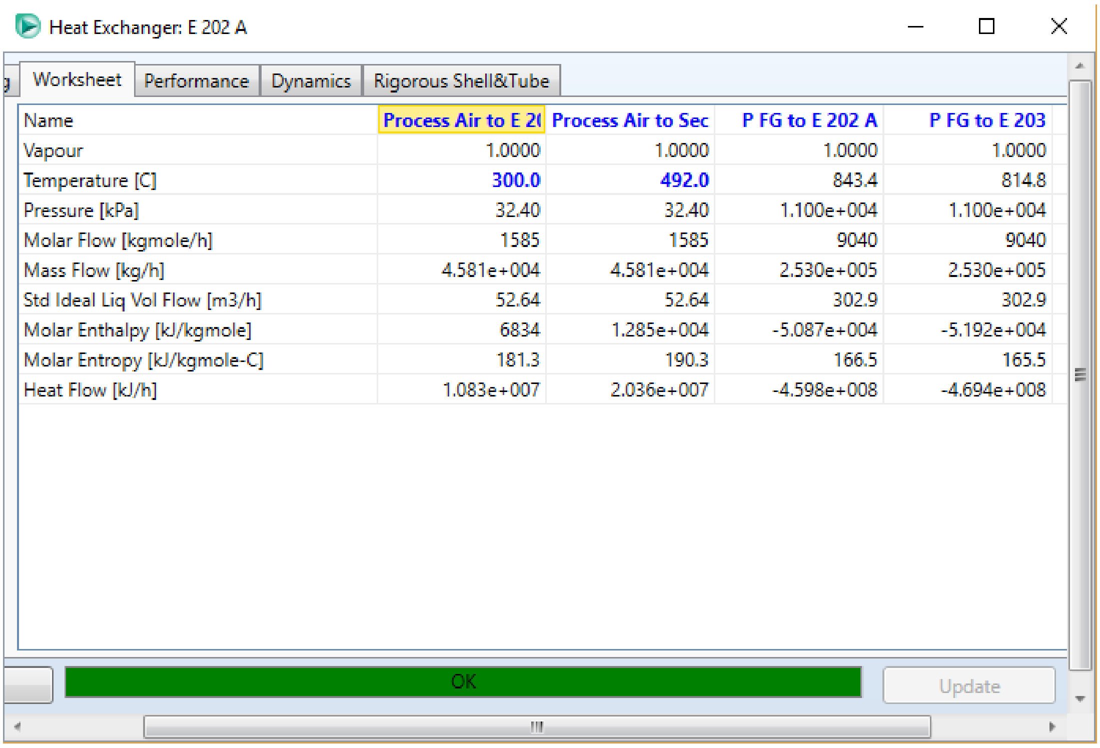
Tube Side Fluid Pkg: Basis-1

Shell Side Fluid Pkg: Basis-1

Convert to Rigorous Model

You can replace any simple exchanger model by a fully rigorous model in your simulation defining a geometry by sizing or by direct specification via input or by importing a prepared file.

Figure 8.24: Exchanger E 202 A Design

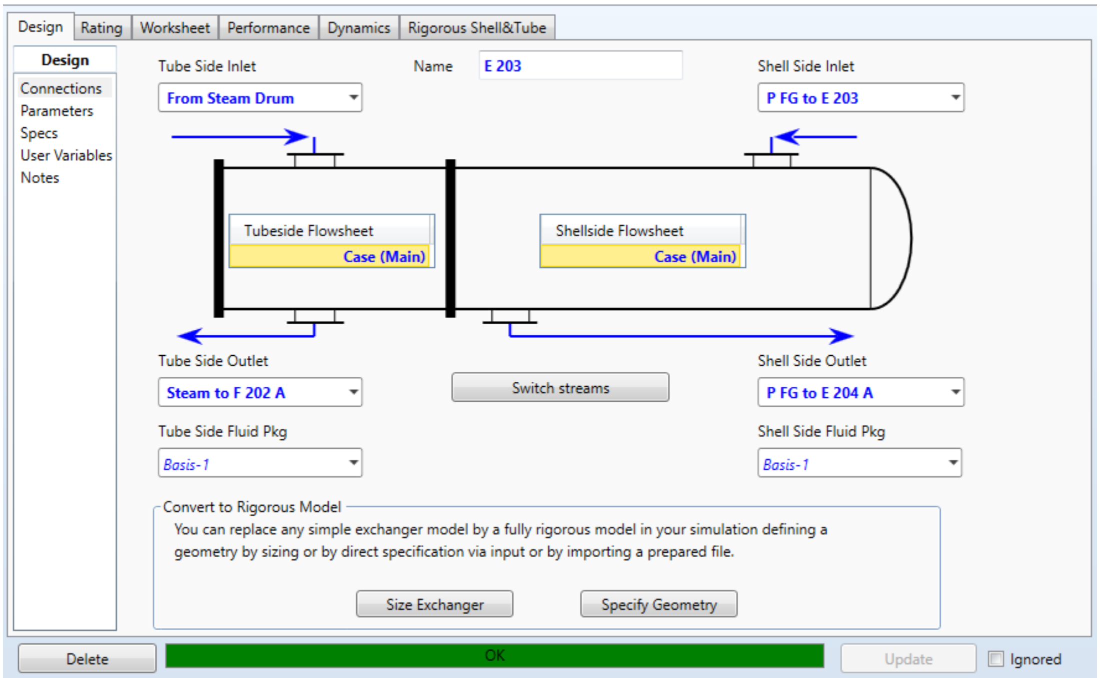


Name	Process Air to E 2(	Process Air to Sec	P FG to E 202 A	P FG to E 203
Vapour	1.0000	1.0000	1.0000	1.0000
Temperature [C]	300.0	492.0	843.4	814.8
Pressure [kPa]	32.40	32.40	1.100e+004	1.100e+004
Molar Flow [kgmole/h]	1585	1585	9040	9040
Mass Flow [kg/h]	4.581e+004	4.581e+004	2.530e+005	2.530e+005
Std Ideal Liq Vol Flow [m3/h]	52.64	52.64	302.9	302.9
Molar Enthalpy [kJ/kgmole]	6834	1.285e+004	-5.087e+004	-5.192e+004
Molar Entropy [kJ/kgmole-C]	181.3	190.3	166.5	165.5
Heat Flow [kJ/h]	1.083e+007	2.036e+007	-4.598e+008	-4.694e+008

Figure 8.25: Exchanger E 202 A Worksheet

Exchanger E 203

In this exchanger, Steam from steam drum is heated using Primary Reformer Flue Gas.



Tube Side Inlet: From Steam Drum

Name: E 203

Shell Side Inlet: P FG to E 203

Tube Side Outlet: Steam to F 202 A

Shell Side Outlet: P FG to E 204 A

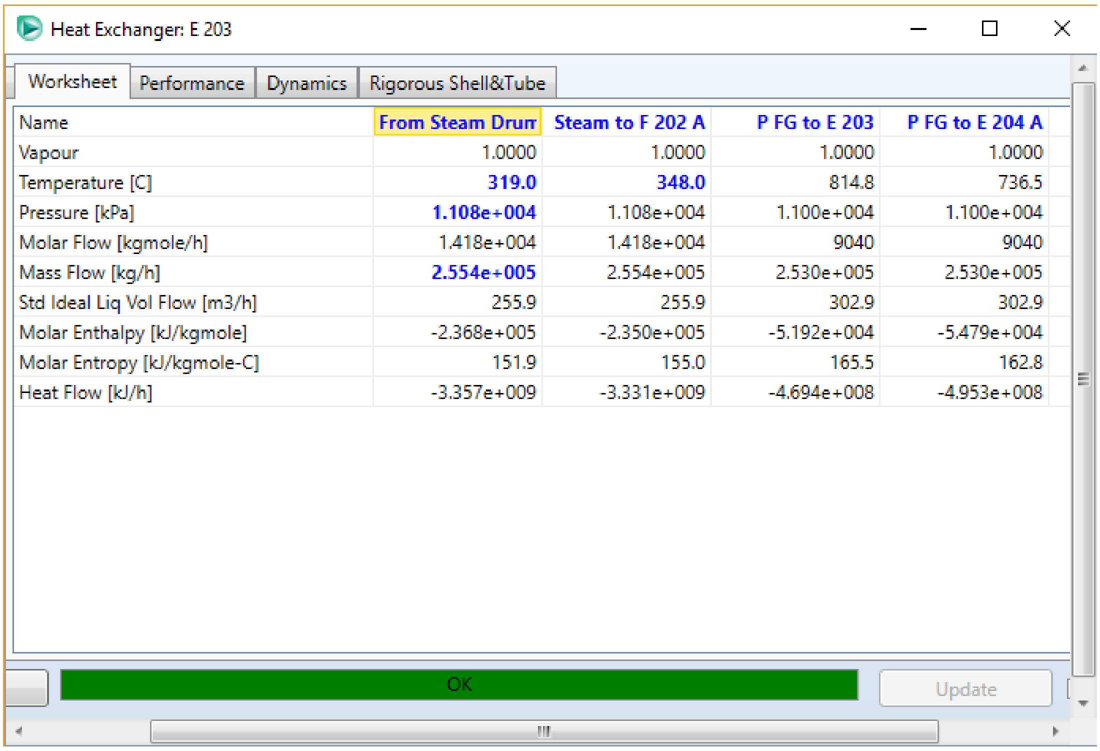
Tube Side Fluid Pkg: Basis-1

Shell Side Fluid Pkg: Basis-1

Convert to Rigorous Model

You can replace any simple exchanger model by a fully rigorous model in your simulation defining a geometry by sizing or by direct specification via input or by importing a prepared file.

Figure 8.26: Exchanger E 203 Design

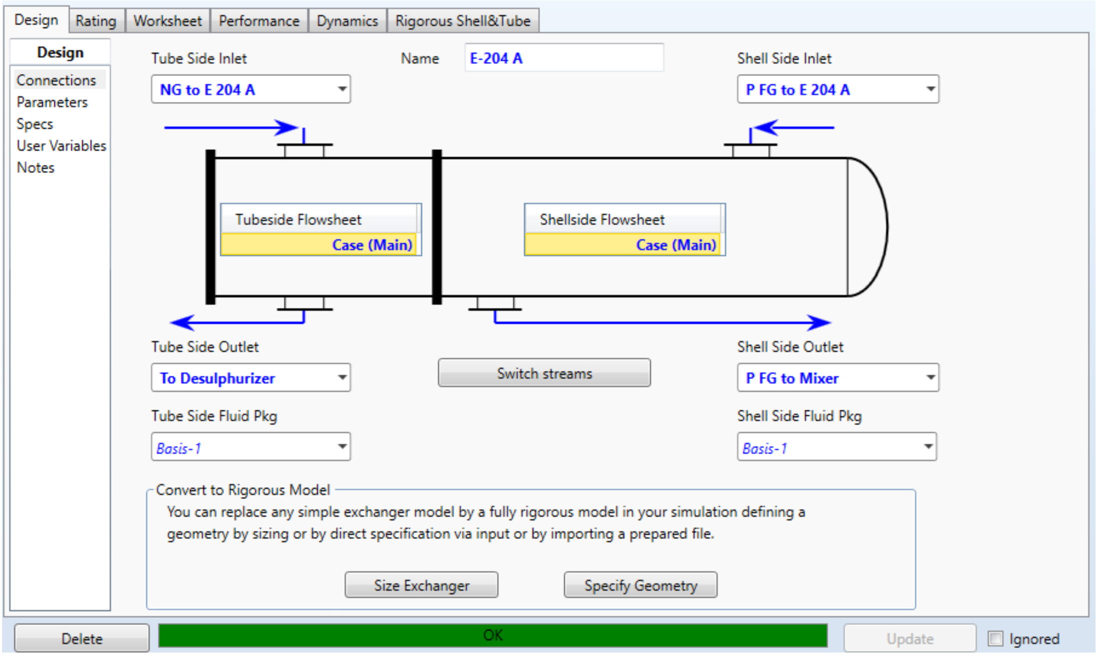


Name	From Steam Drum	Steam to F 202 A	P FG to E 203	P FG to E 204 A
Vapour	1.0000	1.0000	1.0000	1.0000
Temperature [C]	319.0	348.0	814.8	736.5
Pressure [kPa]	1.108e+004	1.108e+004	1.100e+004	1.100e+004
Molar Flow [kgmole/h]	1.418e+004	1.418e+004	9040	9040
Mass Flow [kg/h]	2.554e+005	2.554e+005	2.530e+005	2.530e+005
Std Ideal Liq Vol Flow [m3/h]	255.9	255.9	302.9	302.9
Molar Enthalpy [kJ/kgmole]	-2.368e+005	-2.350e+005	-5.192e+004	-5.479e+004
Molar Entropy [kJ/kgmole-C]	151.9	155.0	165.5	162.8
Heat Flow [kJ/h]	-3.357e+009	-3.331e+009	-4.694e+008	-4.953e+008

Figure 8.27: Exchanger E 203 Worksheet

Exchanger E 204 A

In this exchanger, Natural and Recycle gas is heated using Primary Reformer Flue Gas.



Tube Side Inlet: NG to E 204 A

Name: E-204 A

Shell Side Inlet: P FG to E 204 A

Tube Side Outlet: To Desulphurizer

Shell Side Outlet: P FG to Mixer

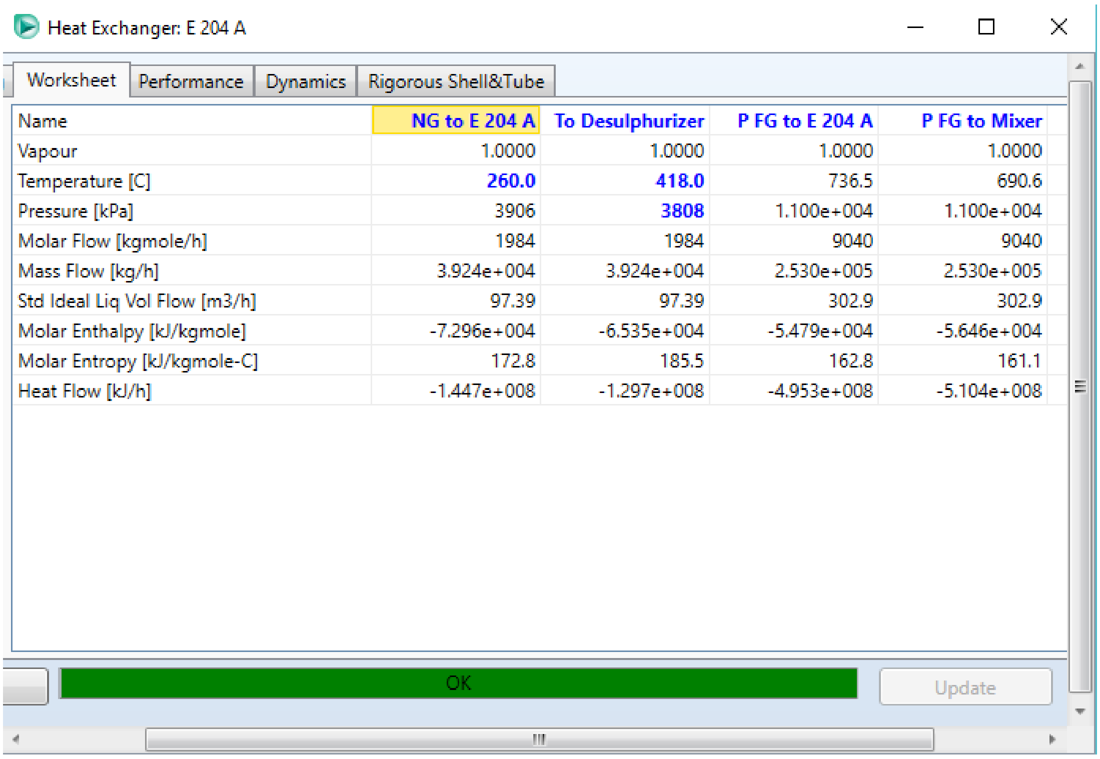
Tube Side Fluid Pkg: *Basis-1*

Shell Side Fluid Pkg: *Basis-1*

Convert to Rigorous Model

You can replace any simple exchanger model by a fully rigorous model in your simulation defining a geometry by sizing or by direct specification via input or by importing a prepared file.

Figure 8.28: Exchanger E 204 A Design

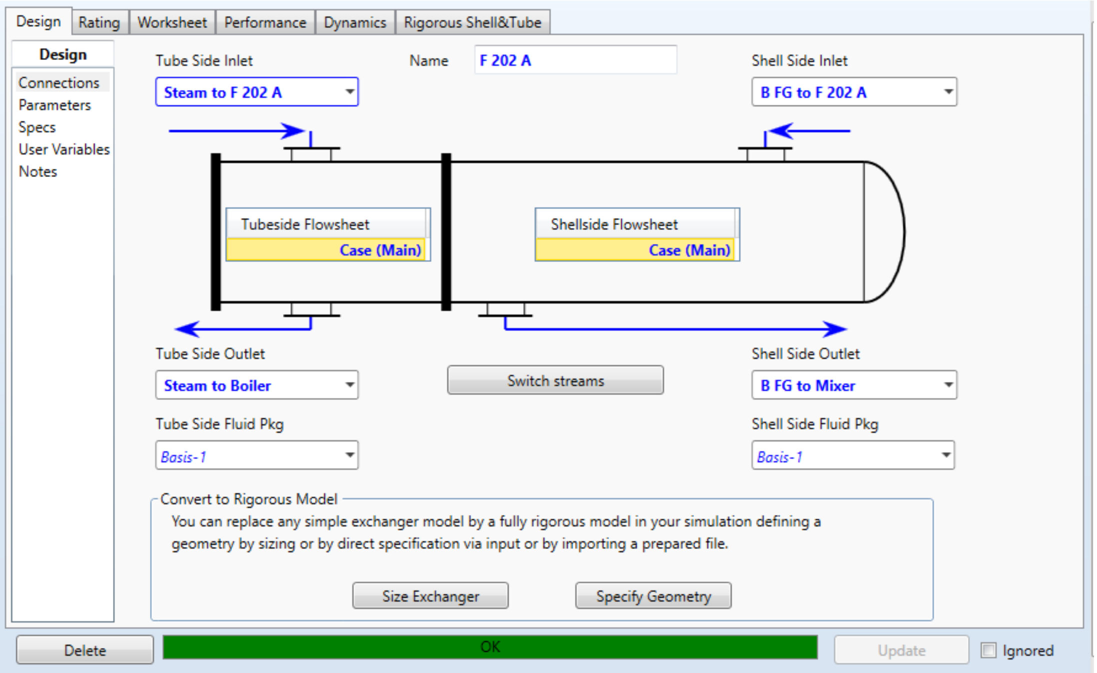


Name	NG to E 204 A	To Desulphurizer	P FG to E 204 A	P FG to Mixer
Vapour	1.0000	1.0000	1.0000	1.0000
Temperature [C]	260.0	418.0	736.5	690.6
Pressure [kPa]	3906	3808	1.100e+004	1.100e+004
Molar Flow [kgmole/h]	1984	1984	9040	9040
Mass Flow [kg/h]	3.924e+004	3.924e+004	2.530e+005	2.530e+005
Std Ideal Liq Vol Flow [m3/h]	97.39	97.39	302.9	302.9
Molar Enthalpy [kJ/kgmole]	-7.296e+004	-6.535e+004	-5.479e+004	-5.646e+004
Molar Entropy [kJ/kgmole-C]	172.8	185.5	162.8	161.1
Heat Flow [kJ/h]	-1.447e+008	-1.297e+008	-4.953e+008	-5.104e+008

Figure 8.29: Exchanger E 204 A Worksheet

Exchanger F 202 A

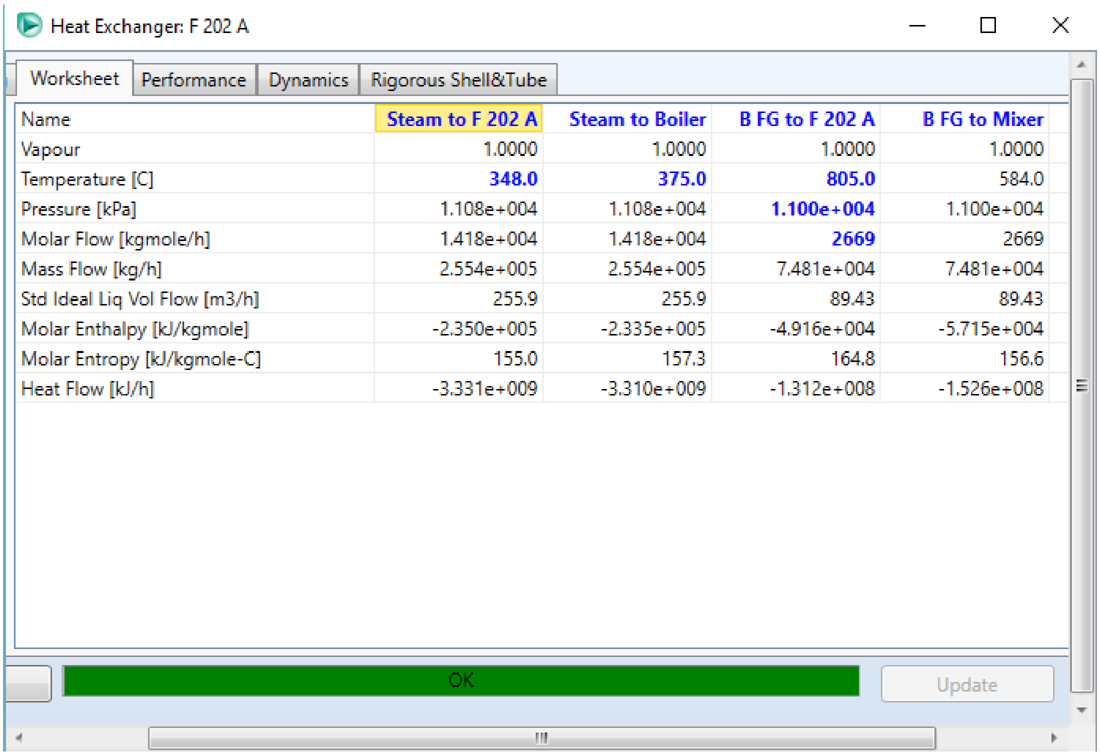
In this exchanger, Steam from Steam Drum is heated using Boiler Reformer Flue Gas.



Tube Side Inlet: Steam to F 202 A
Name: F 202 A
Shell Side Inlet: B FG to F 202 A
Tube Side Outlet: Steam to Boiler
Shell Side Outlet: B FG to Mixer
Tube Side Fluid Pkg: Basis-1
Shell Side Fluid Pkg: Basis-1

Convert to Rigorous Model
You can replace any simple exchanger model by a fully rigorous model in your simulation defining a geometry by sizing or by direct specification via input or by importing a prepared file.

Figure 8.30: Exchanger F 202 A Design

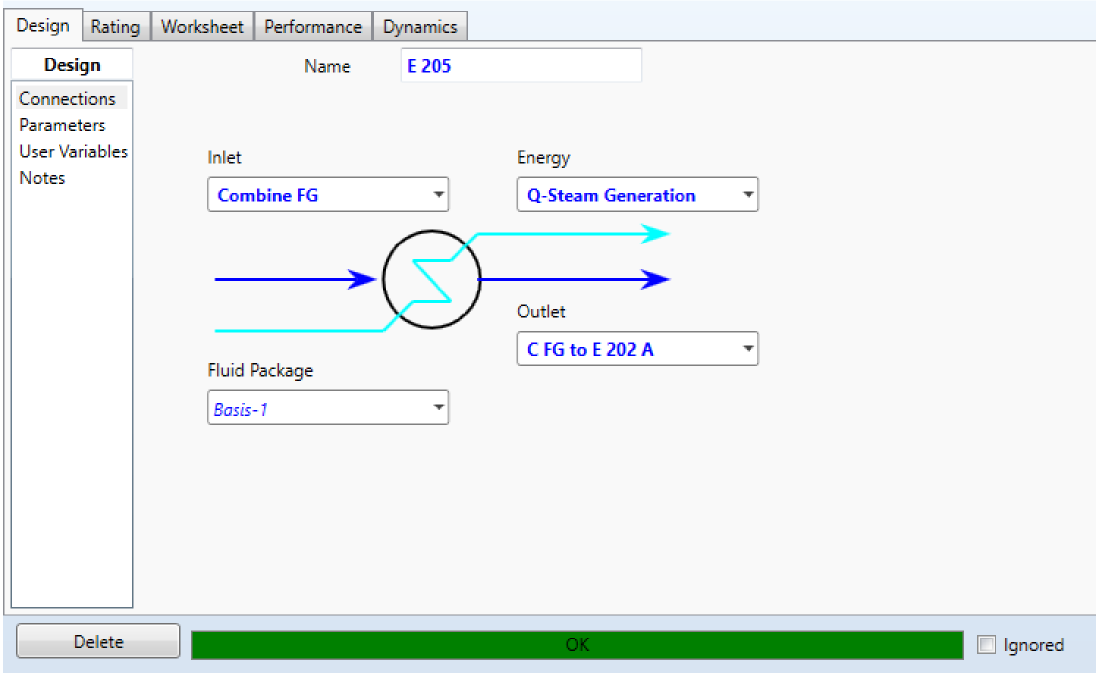


Name	Steam to F 202 A	Steam to Boiler	B FG to F 202 A	B FG to Mixer
Vapour	1.0000	1.0000	1.0000	1.0000
Temperature [C]	348.0	375.0	805.0	584.0
Pressure [kPa]	1.108e+004	1.108e+004	1.100e+004	1.100e+004
Molar Flow [kgmole/h]	1.418e+004	1.418e+004	2669	2669
Mass Flow [kg/h]	2.554e+005	2.554e+005	7.481e+004	7.481e+004
Std Ideal Liq Vol Flow [m3/h]	255.9	255.9	89.43	89.43
Molar Enthalpy [kJ/kgmole]	-2.350e+005	-2.335e+005	-4.916e+004	-5.715e+004
Molar Entropy [kJ/kgmole-C]	155.0	157.3	164.8	156.6
Heat Flow [kJ/h]	-3.331e+009	-3.310e+009	-1.312e+008	-1.526e+008

Figure 8.31: Exchanger F 202 A Worksheet

Exchanger E 205

In this exchanger, High Pressure Steam Generator is installed for cooling Combine Flue Gas.



Name: E 205

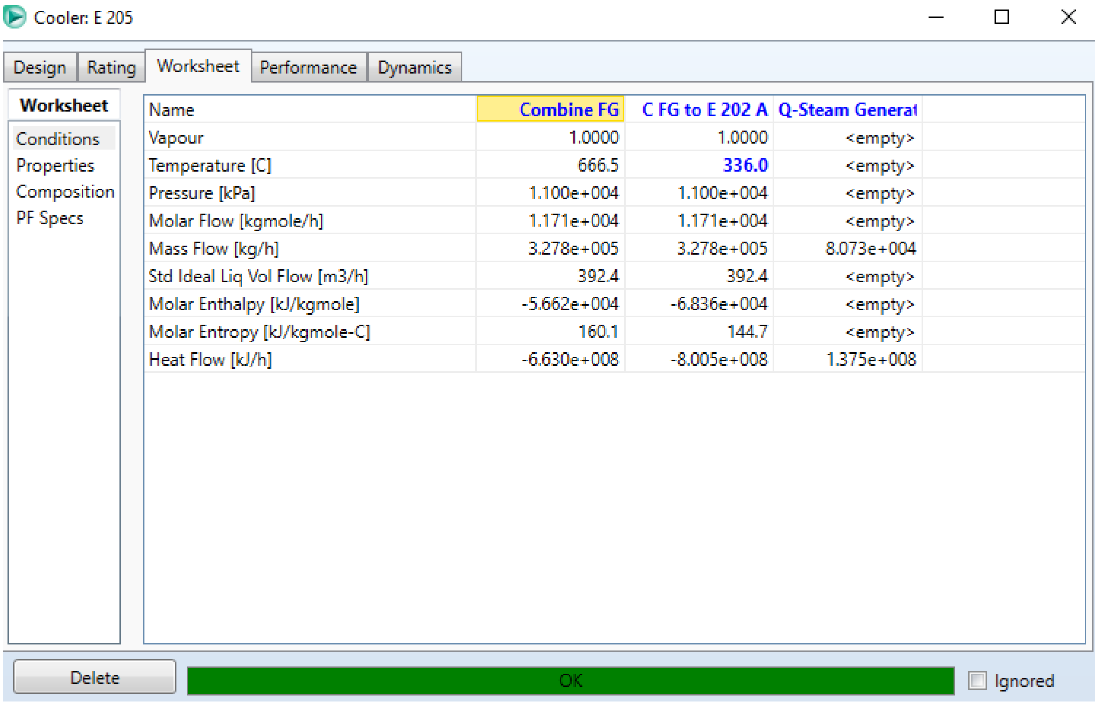
Inlet: Combine FG

Energy: Q-Steam Generation

Outlet: C FG to E 202 A

Fluid Package: *Basis-1*

Figure 8.32: Exchanger E 205 Design

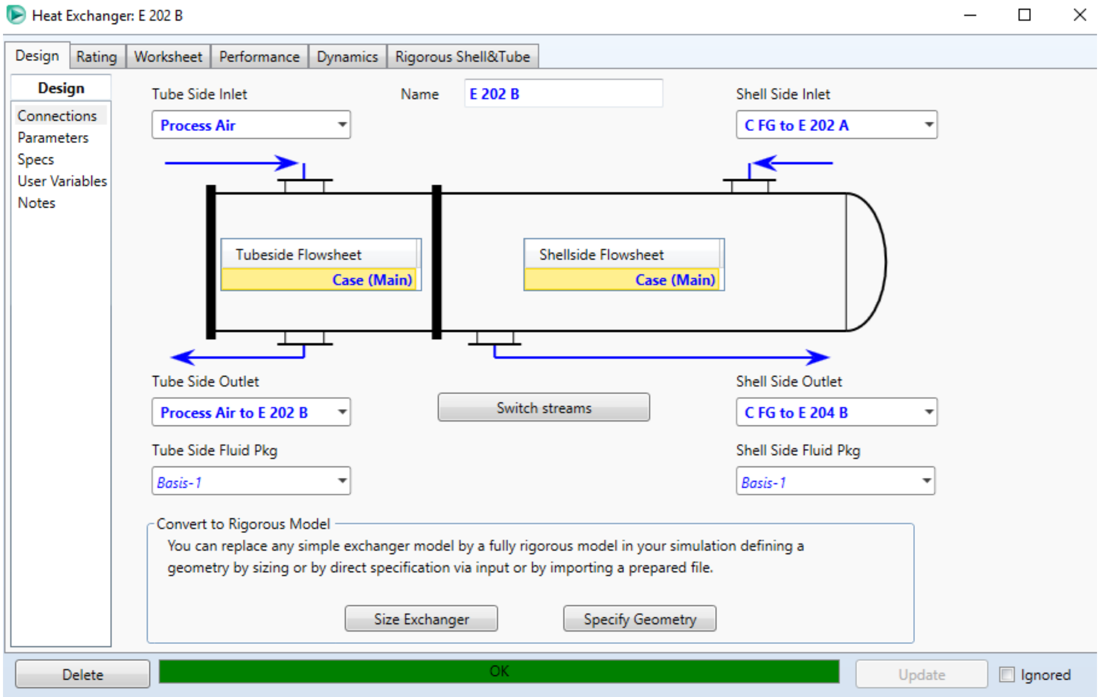


Name	Combine FG	C FG to E 202 A	Q-Steam Generat
Vapour	1.0000	1.0000	<empty>
Temperature [C]	666.5	336.0	<empty>
Pressure [kPa]	1.100e+004	1.100e+004	<empty>
Molar Flow [kgmole/h]	1.171e+004	1.171e+004	<empty>
Mass Flow [kg/h]	3.278e+005	3.278e+005	8.073e+004
Std Ideal Liq Vol Flow [m3/h]	392.4	392.4	<empty>
Molar Enthalpy [kJ/kgmole]	-5.662e+004	-6.836e+004	<empty>
Molar Entropy [kJ/kgmole-C]	160.1	144.7	<empty>
Heat Flow [kJ/h]	-6.630e+008	-8.005e+008	1.375e+008

Figure 8.33: Exchanger E 205 Worksheet

Exchanger E 202 B

In this exchanger, Process Air is heated using Combine Flue Gas.



Tube Side Inlet: Process Air

Name: E 202 B

Shell Side Inlet: C FG to E 202 A

Tube Side Outlet: Process Air to E 202 B

Shell Side Outlet: C FG to E 204 B

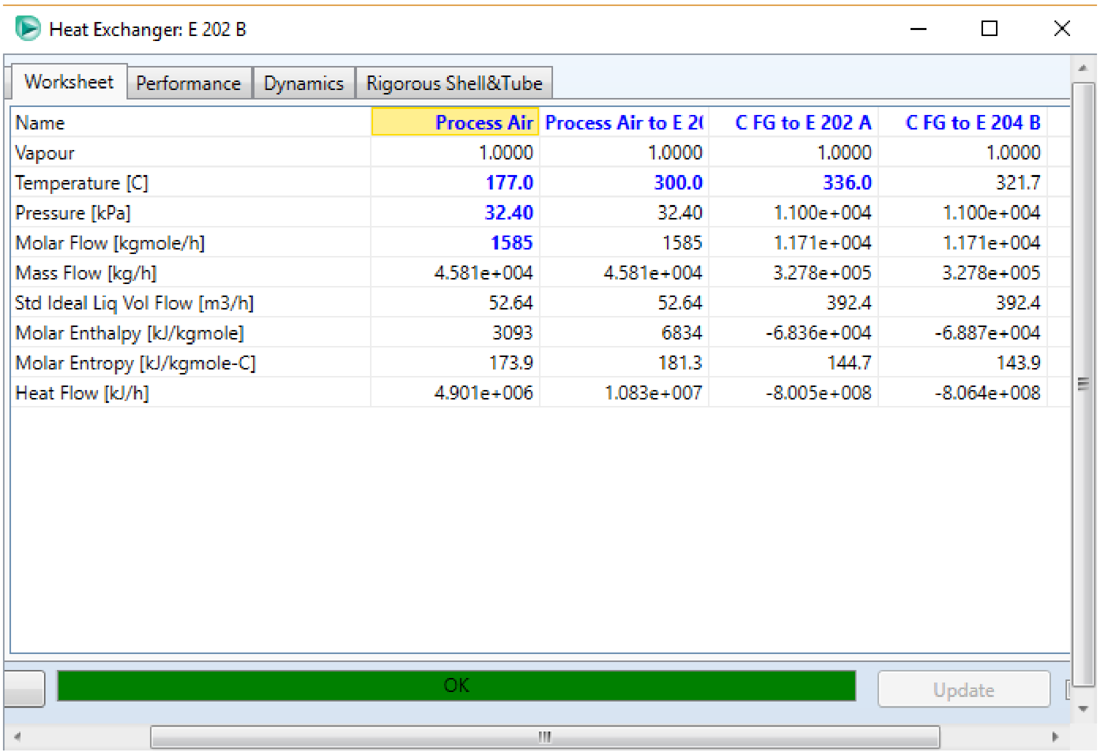
Tube Side Fluid Pkg: Basis-1

Shell Side Fluid Pkg: Basis-1

Convert to Rigorous Model

You can replace any simple exchanger model by a fully rigorous model in your simulation defining a geometry by sizing or by direct specification via input or by importing a prepared file.

Figure 8.34: Exchanger E 202 B Design

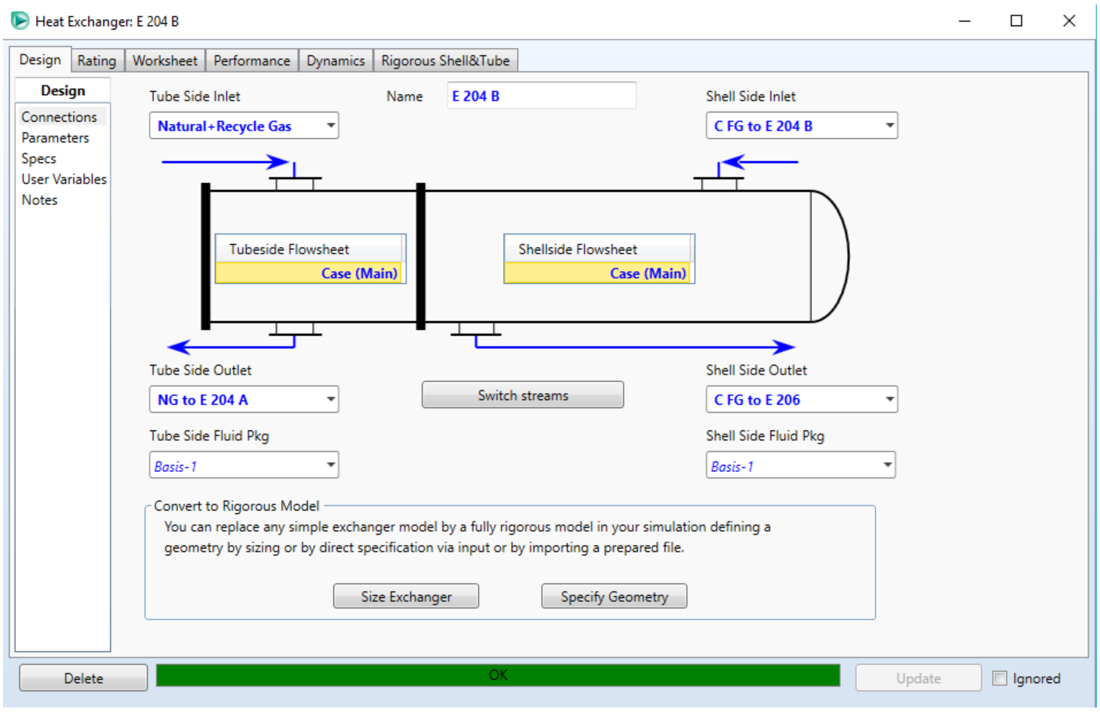


Name	Process Air	Process Air to E 2(	C FG to E 202 A	C FG to E 204 B
Vapour	1.0000	1.0000	1.0000	1.0000
Temperature [C]	177.0	300.0	336.0	321.7
Pressure [kPa]	32.40	32.40	1.100e+004	1.100e+004
Molar Flow [kgmole/h]	1585	1585	1.171e+004	1.171e+004
Mass Flow [kg/h]	4.581e+004	4.581e+004	3.278e+005	3.278e+005
Std Ideal Liq Vol Flow [m3/h]	52.64	52.64	392.4	392.4
Molar Enthalpy [kJ/kgmole]	3093	6834	-6.836e+004	-6.887e+004
Molar Entropy [kJ/kgmole-C]	173.9	181.3	144.7	143.9
Heat Flow [kJ/h]	4.901e+006	1.083e+007	-8.005e+008	-8.064e+008

Figure 8.35: Exchanger E 202 B Worksheet

Exchanger E 204 B

In this exchanger, Natural and Recycle gas is heated using Combine Flue Gas.



Tube Side Inlet: Natural+Recycle Gas

Name: E 204 B

Shell Side Inlet: C FG to E 204 B

Tube Side Outlet: NG to E 204 A

Shell Side Outlet: C FG to E 206

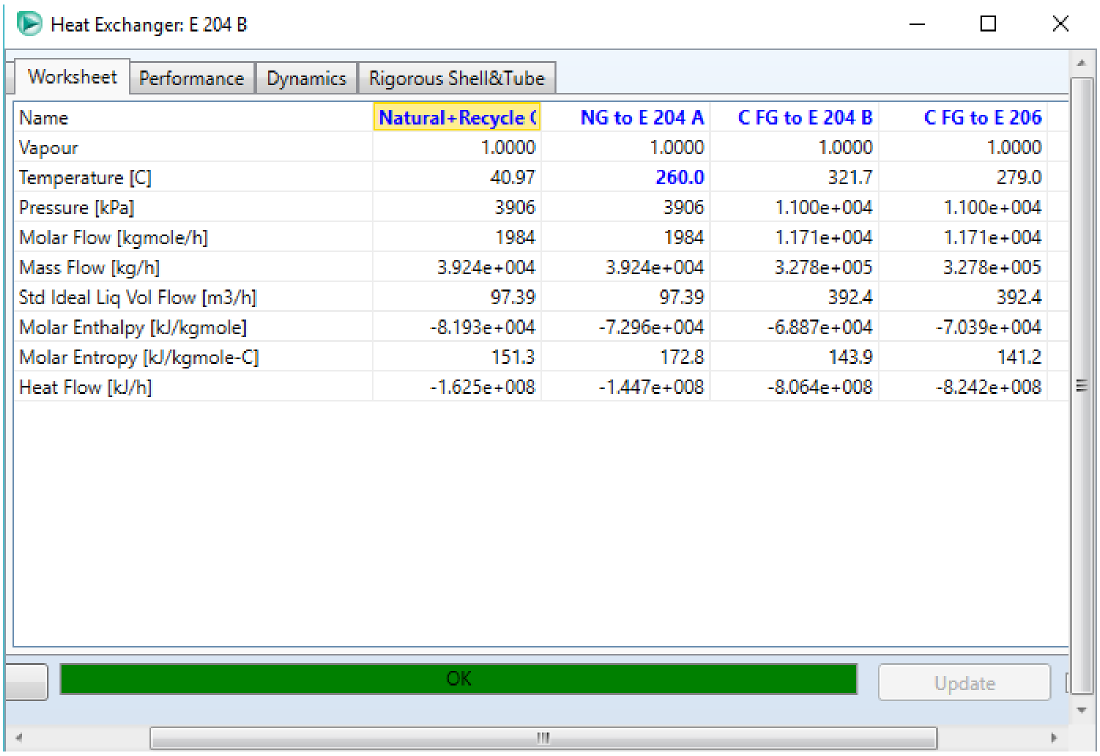
Tube Side Fluid Pkg: Basis-1

Shell Side Fluid Pkg: Basis-1

Convert to Rigorous Model

You can replace any simple exchanger model by a fully rigorous model in your simulation defining a geometry by sizing or by direct specification via input or by importing a prepared file.

Figure 8.36: Exchanger E 204 B Design

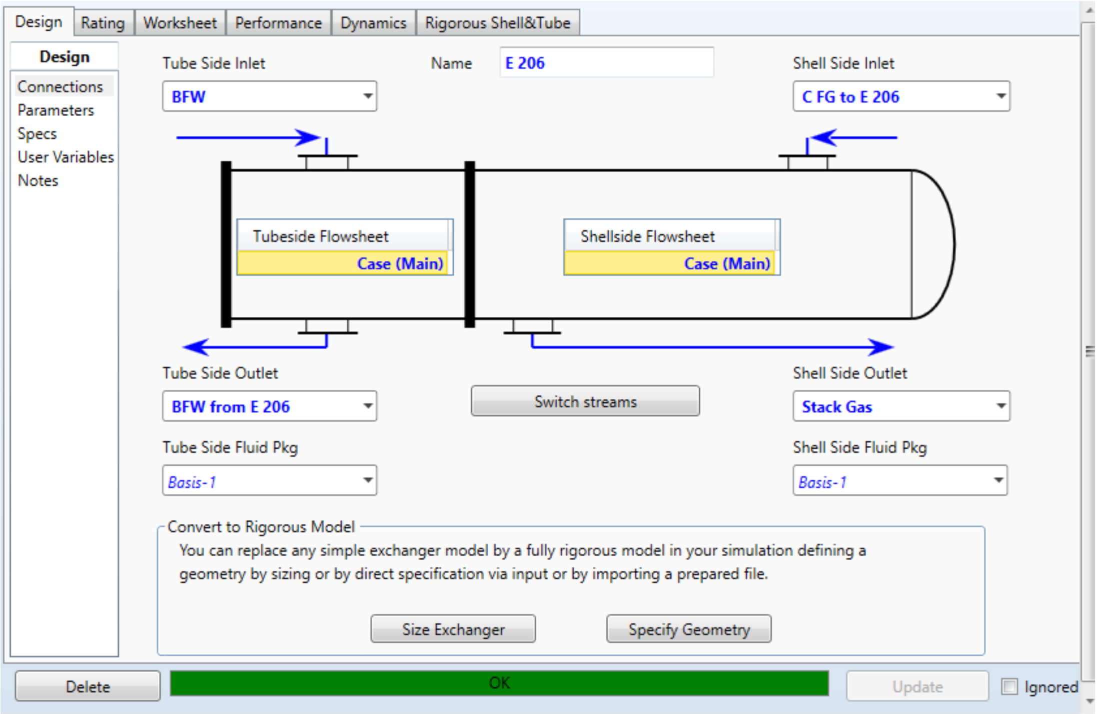


Name	Natural+Recycle (	NG to E 204 A	C FG to E 204 B	C FG to E 206
Vapour	1.0000	1.0000	1.0000	1.0000
Temperature [C]	40.97	260.0	321.7	279.0
Pressure [kPa]	3906	3906	1.100e+004	1.100e+004
Molar Flow [kgmole/h]	1984	1984	1.171e+004	1.171e+004
Mass Flow [kg/h]	3.924e+004	3.924e+004	3.278e+005	3.278e+005
Std Ideal Liq Vol Flow [m3/h]	97.39	97.39	392.4	392.4
Molar Enthalpy [kJ/kgmole]	-8.193e+004	-7.296e+004	-6.887e+004	-7.039e+004
Molar Entropy [kJ/kgmole-C]	151.3	172.8	143.9	141.2
Heat Flow [kJ/h]	-1.625e+008	-1.447e+008	-8.064e+008	-8.242e+008

Figure 8.37: Exchanger E 204 B Worksheet

Exchanger E 206

In this exchanger, Boiler Feed Water is heated using Combine Flue Gas.



Tube Side Inlet: BFW

Name: E 206

Shell Side Inlet: C FG to E 206

Tube Side Outlet: BFW from E 206

Shell Side Outlet: Stack Gas

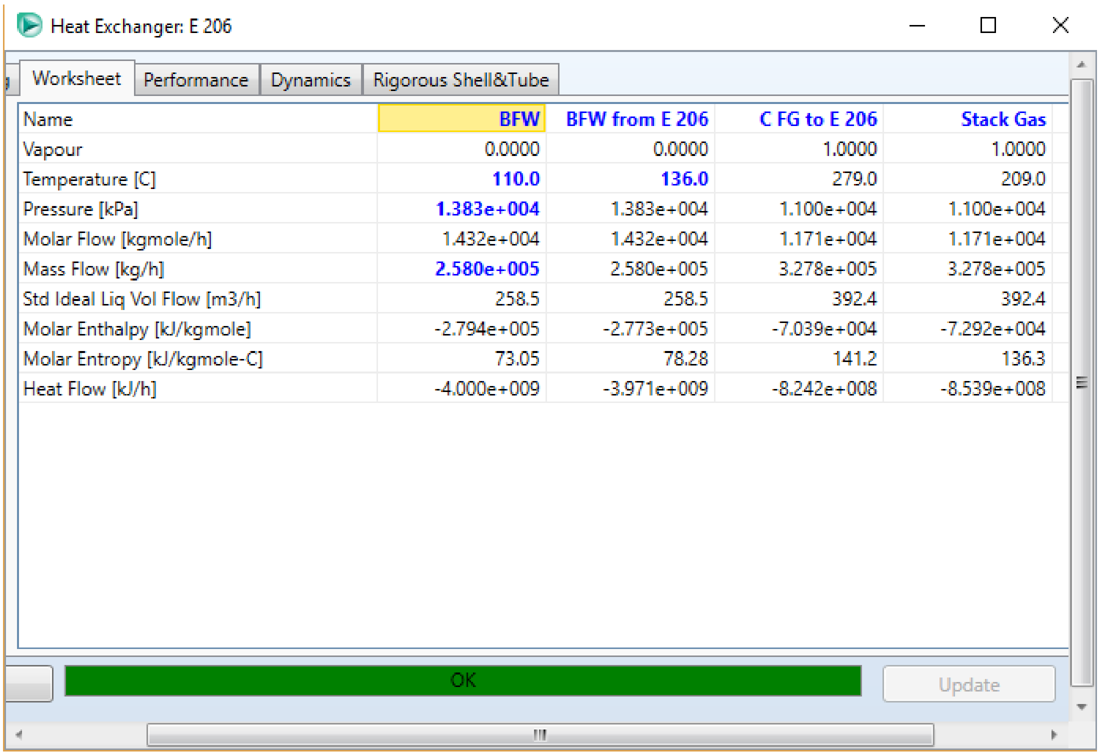
Tube Side Fluid Pkg: Basis-1

Shell Side Fluid Pkg: Basis-1

Convert to Rigorous Model

You can replace any simple exchanger model by a fully rigorous model in your simulation defining a geometry by sizing or by direct specification via input or by importing a prepared file.

Figure 8.38: Exchanger E 206 Design



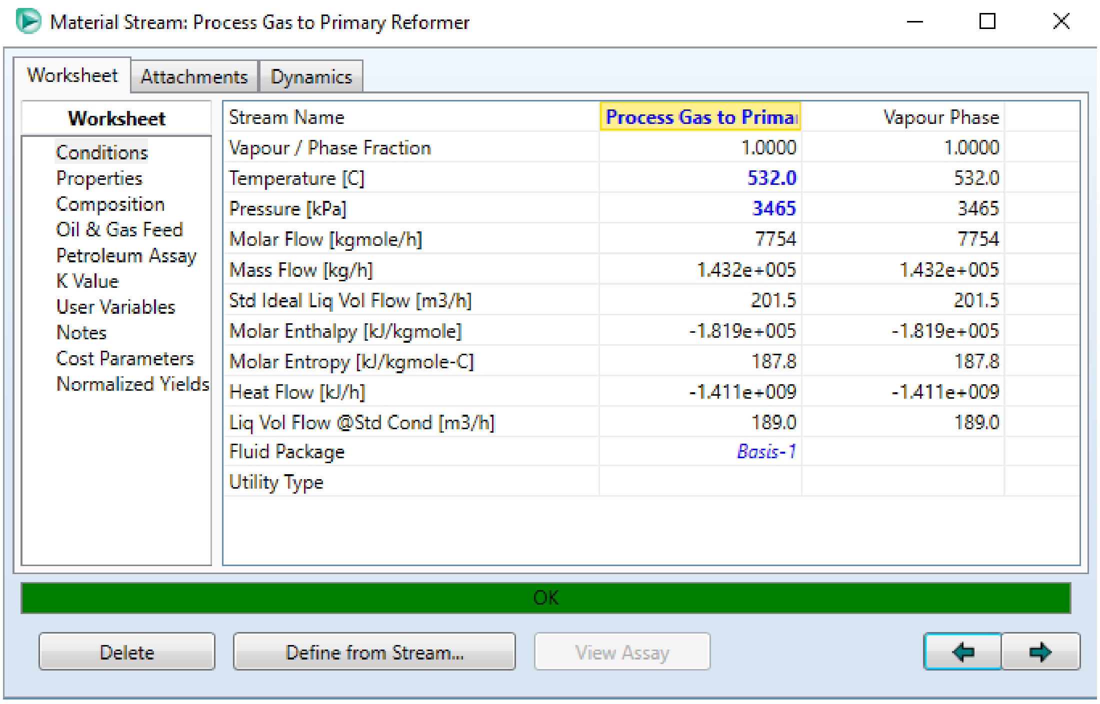
Name	BFW	BFW from E 206	C FG to E 206	Stack Gas
Vapour	0.0000	0.0000	1.0000	1.0000
Temperature [C]	110.0	136.0	279.0	209.0
Pressure [kPa]	1.383e+004	1.383e+004	1.100e+004	1.100e+004
Molar Flow [kgmole/h]	1.432e+004	1.432e+004	1.171e+004	1.171e+004
Mass Flow [kg/h]	2.580e+005	2.580e+005	3.278e+005	3.278e+005
Std Ideal Liq Vol Flow [m3/h]	258.5	258.5	392.4	392.4
Molar Enthalpy [kJ/kgmole]	-2.794e+005	-2.773e+005	-7.039e+004	-7.292e+004
Molar Entropy [kJ/kgmole-C]	73.05	78.28	141.2	136.3
Heat Flow [kJ/h]	-4.000e+009	-3.971e+009	-8.242e+008	-8.539e+008

Figure 8.39: Exchanger E 206 Worksheet

8.1.7 Product Streams

Following are the heated product streams.

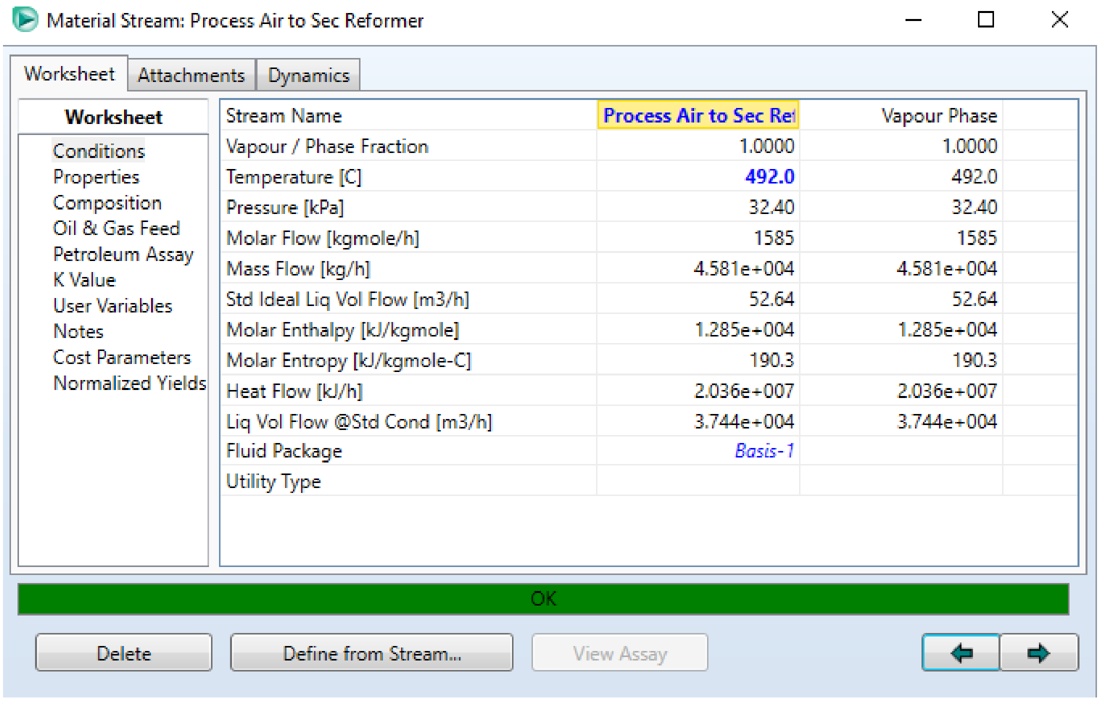
Process Gas to Primary Reformer



Stream Name	Process Gas to Prima	Vapour Phase
Vapour / Phase Fraction	1.0000	1.0000
Temperature [C]	532.0	532.0
Pressure [kPa]	3465	3465
Molar Flow [kgmole/h]	7754	7754
Mass Flow [kg/h]	1.432e+005	1.432e+005
Std Ideal Liq Vol Flow [m3/h]	201.5	201.5
Molar Enthalpy [kJ/kgmole]	-1.819e+005	-1.819e+005
Molar Entropy [kJ/kgmole-C]	187.8	187.8
Heat Flow [kJ/h]	-1.411e+009	-1.411e+009
Liq Vol Flow @Std Cond [m3/h]	189.0	189.0
Fluid Package	*Basis-1*	
Utility Type		

Figure 8.40: Process Gas to Primary Reformer

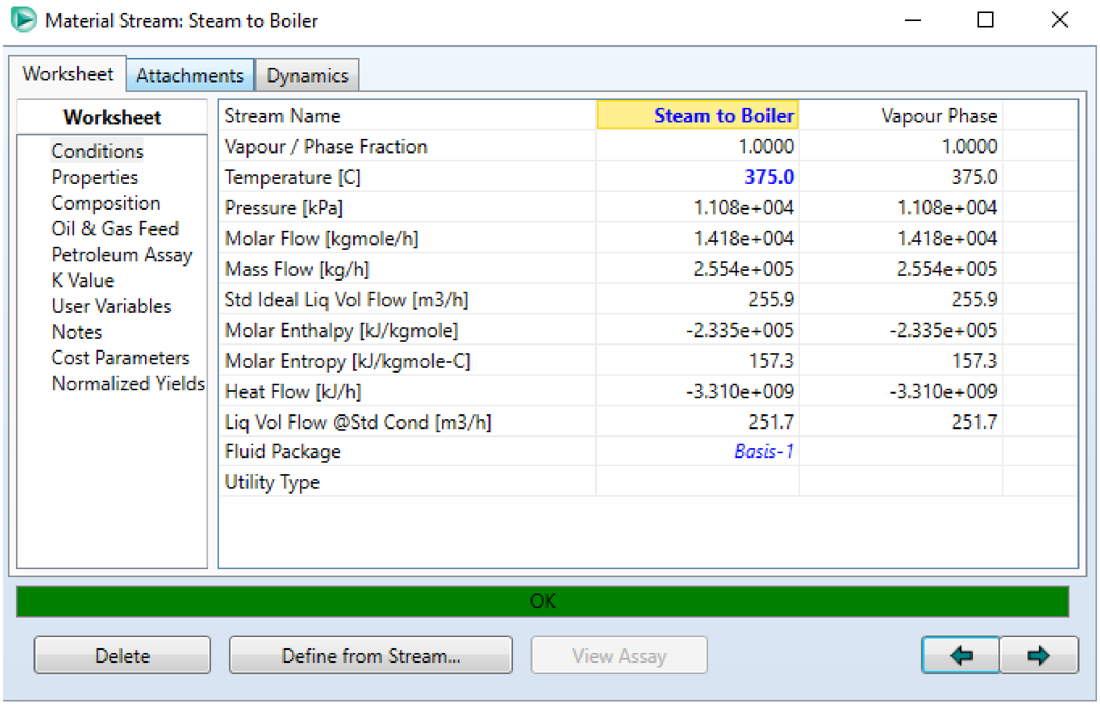
Process Air to Secondary Reformer



Stream Name	Process Air to Sec Re	Vapour Phase
Vapour / Phase Fraction	1.0000	1.0000
Temperature [C]	492.0	492.0
Pressure [kPa]	32.40	32.40
Molar Flow [kgmole/h]	1585	1585
Mass Flow [kg/h]	4.581e+004	4.581e+004
Std Ideal Liq Vol Flow [m3/h]	52.64	52.64
Molar Enthalpy [kJ/kgmole]	1.285e+004	1.285e+004
Molar Entropy [kJ/kgmole-C]	190.3	190.3
Heat Flow [kJ/h]	2.036e+007	2.036e+007
Liq Vol Flow @Std Cond [m3/h]	3.744e+004	3.744e+004
Fluid Package	*Basis-1*	
Utility Type		

Figure 8.41: Process Air to Secondary Reformer

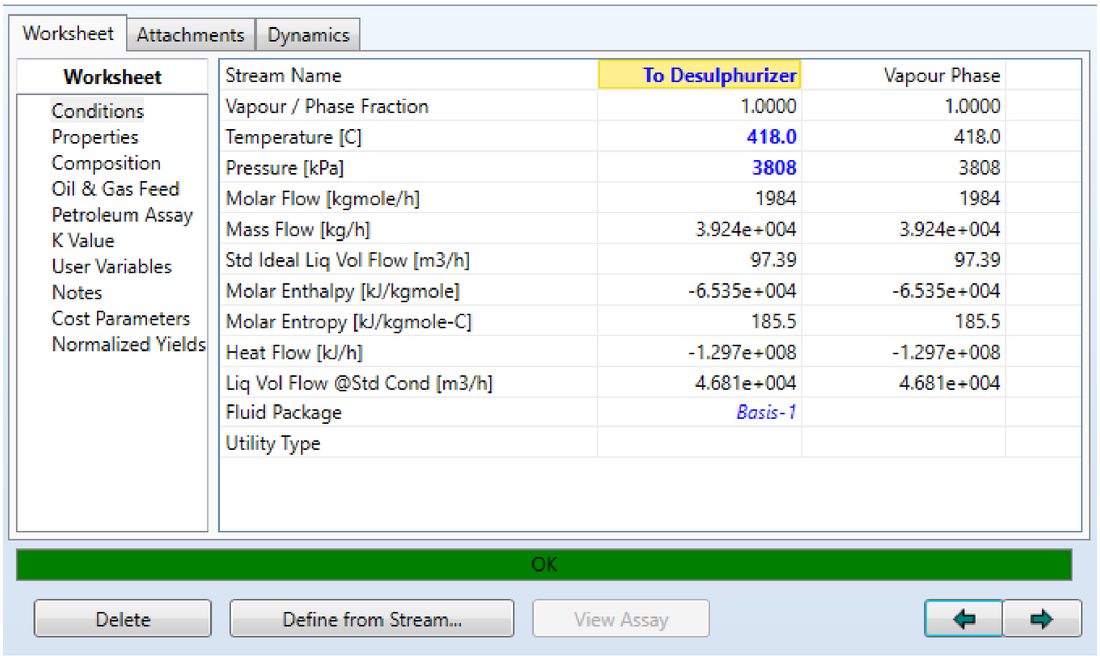
Steam to Boiler



Stream Name	Steam to Boiler	Vapour Phase
Vapour / Phase Fraction	1.0000	1.0000
Temperature [C]	375.0	375.0
Pressure [kPa]	1.108e+004	1.108e+004
Molar Flow [kgmole/h]	1.418e+004	1.418e+004
Mass Flow [kg/h]	2.554e+005	2.554e+005
Std Ideal Liq Vol Flow [m3/h]	255.9	255.9
Molar Enthalpy [kJ/kgmole]	-2.335e+005	-2.335e+005
Molar Entropy [kJ/kgmole-C]	157.3	157.3
Heat Flow [kJ/h]	-3.310e+009	-3.310e+009
Liq Vol Flow @Std Cond [m3/h]	251.7	251.7
Fluid Package	Basis-1	
Utility Type		

Figure 8.42: Steam to Boiler

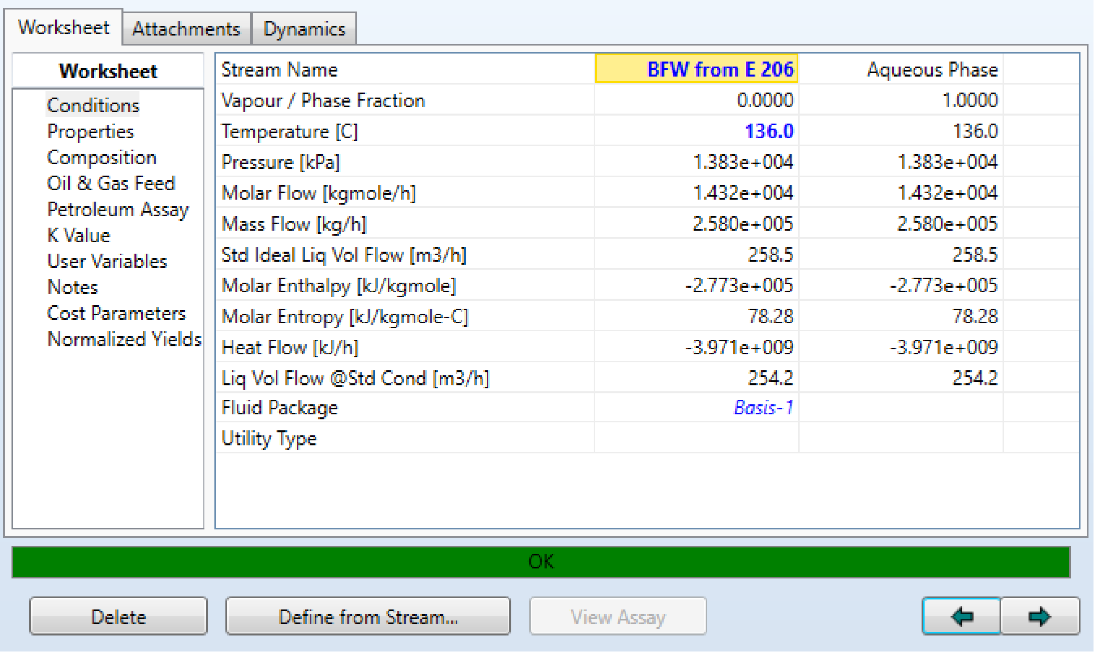
Natural and Recycle Gas to Desulphurizer



Stream Name	To Desulphurizer	Vapour Phase
Vapour / Phase Fraction	1.0000	1.0000
Temperature [C]	418.0	418.0
Pressure [kPa]	3808	3808
Molar Flow [kgmole/h]	1984	1984
Mass Flow [kg/h]	3.924e+004	3.924e+004
Std Ideal Liq Vol Flow [m3/h]	97.39	97.39
Molar Enthalpy [kJ/kgmole]	-6.535e+004	-6.535e+004
Molar Entropy [kJ/kgmole-C]	185.5	185.5
Heat Flow [kJ/h]	-1.297e+008	-1.297e+008
Liq Vol Flow @Std Cond [m3/h]	4.681e+004	4.681e+004
Fluid Package	Basis-1	
Utility Type		

Figure 8.43: Natural & Recycle Gas to Desulphurizer

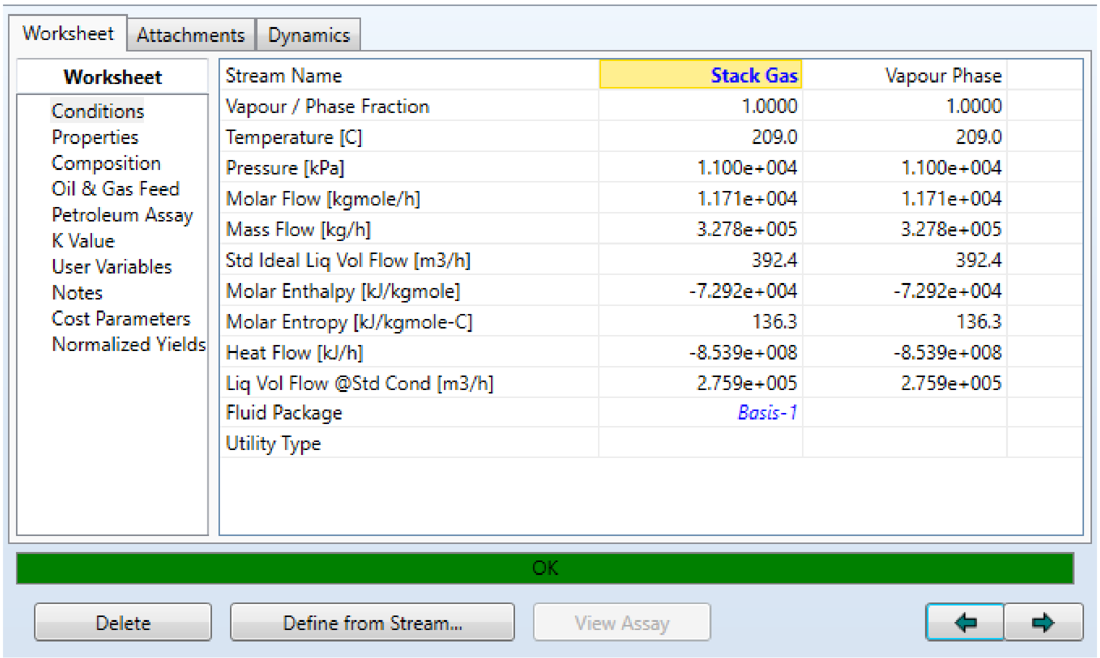
Boiler Feed Water Outlet



Stream Name	BFW from E 206	Aqueous Phase
Vapour / Phase Fraction	0.0000	1.0000
Temperature [C]	136.0	136.0
Pressure [kPa]	1.383e+004	1.383e+004
Molar Flow [kgmole/h]	1.432e+004	1.432e+004
Mass Flow [kg/h]	2.580e+005	2.580e+005
Std Ideal Liq Vol Flow [m3/h]	258.5	258.5
Molar Enthalpy [kJ/kgmole]	-2.773e+005	-2.773e+005
Molar Entropy [kJ/kgmole-C]	78.28	78.28
Heat Flow [kJ/h]	-3.971e+009	-3.971e+009
Liq Vol Flow @Std Cond [m3/h]	254.2	254.2
Fluid Package	Basis-1	
Utility Type		

Figure 8.44: Boiler Feed Water Outlet

Stack Gas

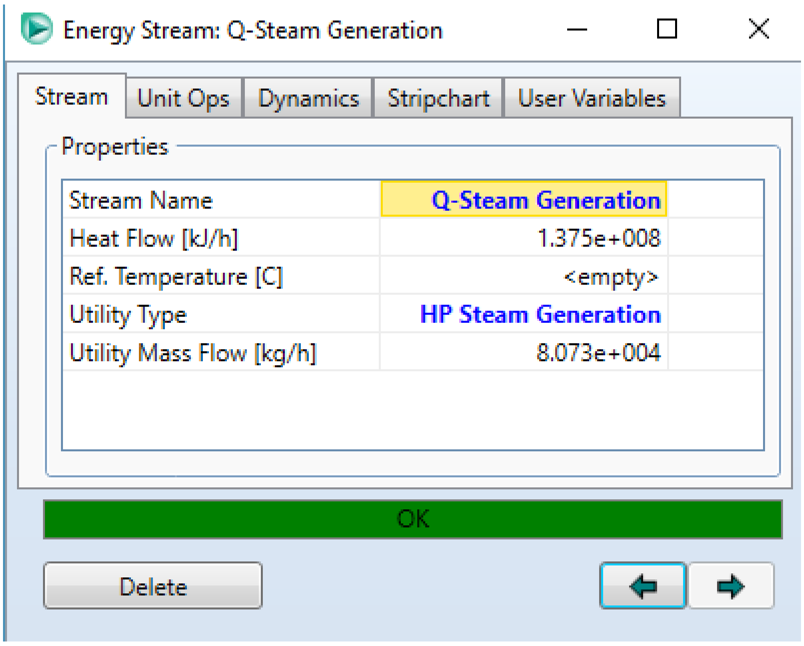


Stream Name	Stack Gas	Vapour Phase
Vapour / Phase Fraction	1.0000	1.0000
Temperature [C]	209.0	209.0
Pressure [kPa]	1.100e+004	1.100e+004
Molar Flow [kgmole/h]	1.171e+004	1.171e+004
Mass Flow [kg/h]	3.278e+005	3.278e+005
Std Ideal Liq Vol Flow [m3/h]	392.4	392.4
Molar Enthalpy [kJ/kgmole]	-7.292e+004	-7.292e+004
Molar Entropy [kJ/kgmole-C]	136.3	136.3
Heat Flow [kJ/h]	-8.539e+008	-8.539e+008
Liq Vol Flow @Std Cond [m3/h]	2.759e+005	2.759e+005
Fluid Package	Basis-1	
Utility Type		

Figure 8.45: Stack Gas

8.1.6 Higher Pressure Steam Generation

Following is the production rate of steam in E 205 simulated in Aspen HYSYS.



Stream Name	Q-Steam Generation
Heat Flow [kJ/h]	1.375e+008
Ref. Temperature [C]	<empty>
Utility Type	HP Steam Generation
Utility Mass Flow [kg/h]	8.073e+004

Figure 8.46: High Pressure Generation

8.2 Results

The temperature of stack gas calculated using Aspen Energy Analyzer was 201°C. However, temperature of stack gas simulated on Aspen HYSYS is 209°C. Slight difference in results can be due to fluid package selected or pressure drop factors. Results are significantly close and proposed modifications are supported by Aspen HYSYS simulation.

Chapter 9
HAZOP Analysis

HAZOP analysis is used as a fragment of a Quantitative Risk Assessment. The aim is to investigate the divergence of the plant, from the design intent and formulate risk for workforce and equipment and operability complications.

9.1 Why HAZOP?

The motives for carrying out this study, are primarily, to identify hazards, and to resolve them.

The following Guidewords along with the meanings are illustrated below.

Guidewords	Meaning
NO	Complete negation of the design intent
MORE	Quantitative Increase
LESS	Quantitative Decrease
REVERSE	Logical opposite of the design intent
INSTEAD	Complete substitution

Table 9-1: Guide Words for HAZOP

9.2 Analysis

HAZOP Analysis is done on the major equipment used in convective section of primary reformer.

9.2.1 Valve

Equipment: Valve

Purpose: To control the flow of inlet streams in the reformer

Parameter	Deviation	Causes	Consequences	Action
Flow	More	Control Valve fails open	Low heat recovery	Installation of Flow sensors
Flow	No	Control Valve fails closed	No heat exchanges in the coil	Replace or clean valve

Table 9-2: HAZOP Study on Valve

9.2.2 Coils

Equipment: Coils

Purpose: To exchanges heat with Flue gas

Parameter	Deviation	Causes	Consequences	Action
Flow	No	Control Valve fails to open	No heat exchanges in coils	Install temperature sensors
Flow	More	Failure of valve to be close	Output temperature too low	Install temperature sensors before and after process streams
Flow	Less	Pipe leakage	Temperature of process steam is low	Install flow sensors
Flow	Reverse	Failure of process fluid inlet valve	Product offset	Installation of check valve
Contamination	high	Poor separation	Low heat recovery	Proper maintenance
Pressure (shell side)	low	Inlet pump failed to work properly	No significant variation	Installation of pressure sensors
Pressure (shell side)	high	Exchanger outlet discharge valve fails to open	Exchanger shell side will be over pressurized	High pressure security must be installed on shell outlet which if actuated will close all valves.

Table 9-3: HAZOP Study of Coils

9.2.3 Boiler

Equipment: Boiler

Purpose: To produce steam

Parameter	Deviation	Causes	Consequences	Action
Flow	Low	Failure of inlet control valve	Boiler heated the reactants to high temperature than required	Install flow sensors
Flow	High	Inlet valve opens	Poor heating of reactants	Install Control valves
Temperature	High	Failure of temperature control valve	Chances of explosion	Install temperature sensors
Temperature	low	High flow rate of inlet streams	Low temperature of process streams	Install Control Valves

Table 9-4: HAZOP Study on Boiler

Chapter 10

Conclusion

By using pinch analysis, not only the source of heat wastage was identified, that is high temperature flue gas being released into environment, but also it suggested away to recover waste heat. Hence led to better heat integration. Around 7% more steam can be produced and stack temperature of the flue gas can be reduced by 19°C using proposed heat exchanger network design. Proposed design also achieves 91.5% of the area target calculated using Bath Algorithm.

Economic analysis shows the payback period after applying these modifications is just 2.06 years with the cost saving of 0.45 million dollars per year. Due to reduction in stack temperature of flue gases, environmental impact is also reduced.

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