RESOURCE CONSERVATION THROUGH A HIERARCHICAL

APPROACH OF MASS AND ENERGY INTEGRATION

A Dissertation

by

RUBAYAT MAHMUD

Submitted to the Office of Graduate Studies of
Texas A&M University
in partial fulfillment of the requirements for the degree of

DOCTOR OF PHILOSOPHY

December 2005

Major Subject: Chemical Engineering

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Approved by:

Chair of Committee,	Mahmoud M. El-Halwagi
Committee Members,	Sam Mannan
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	Thomas Lalk
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ABSTRACT

Resource Conservation Through a Hierarchical Approach of

Mass and Energy Integration. (December 2005)

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The objective of this work was to develop a systematic methodology for simultaneously targeting and optimizing heating, cooling, power cogeneration, and waste management for any processing facility. A systems approach was used to characterize the complex interactions between the various forms of material and energy utilities as well as their interactions with the core processing units. Two approaches were developed: graphical and mathematical. In both approaches, a hierarchical procedure was developed to decompose the problem into successive stages that were globally solvable then. The solution fragments were then merged into overall process solutions and targets. The whole approach was a systems approach of solving problems. The methodology was developed from the insights from several state of the art process integration techniques. In particular, the dissertation introduced a consistent framework for simultaneously addressing heat-exchange networks, material-recovery networks, combined heat and power, fuel optimization, and waste management. The graphical approach relied on decomposing the problem into sequential tasks that could be

addressed using visualization tools. The mathematical approach enabled the simultaneous solution of critical subproblems. Because of the non-convexity of the mathematical formulation, a global optimization technique was developed through problem reformulation and discretization. A case study was solved and analyzed to illustrate the effectiveness of the devised methodology.

DEDICATION

To my parents, for all their supports and sacrifices, so that I could pursue my dreams.

For their unconditional love. For making me, what I am today.

To Diba, my dear wife, for her love and understanding. For being an inspiration to me.

For supporting me in every situation.

To Raisa, my lovely daughter. For being the light in our life.

ACKNOWLEDGEMENTS

I convey my profound gratitude to my advisor Dr. Mahmoud M. El-Halwagi for his timely guidance, advice and encouragement throughout my study at A&M, my dissertation work and also in personal matters. I am really thankful to him for his constant support and appreciation of my work. I thank him for making my life easier while studying in A&M. It was a great learning experience for me to work under Dr. El-Halwagi.

I thank my committee members Dr. Mannan, Dr. Baldwin and Dr. Lalk for their time and constructive advice regarding my dissertation work.

I convey my appreciation to my group members for providing a very suitable and interesting environment. I convey my thanks to all the faculty members and students of the Chemical Engineering Department at A&M for creating such a fabulous academic environment.

Finally I would like to thank my parents, my wife, my younger sister and my little angel for being the motivation throughout my life.

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CHAPTER I

INTRODUCTION

Reduction of operating cost is among the principal objectives of chemical manufacturing processes. Process utilities are among the key contributors to the operating cost of the process. One way of classifying process utilities is to categorize them into energy utilities and material utilities. Energy utilities include fuel, heating media (e.g., steam, heating oil, etc.), cooling media (e.g., refrigerant, cooling water, etc.), and electric power. Material utilities include mass-separating agents (MSAs), catalysts, waste-treatment agents, and process consumables (other than raw materials).

Over the past two decades, significant progress has been made in optimizing several subsystems of energy and material utilities. In this context, process integration and optimization techniques have been utilized to aid the chemical process industry in utility reduction, pollution abatement, and cost cutting efforts. State of the art process integration tools and techniques have been developed to address issues like optimum mass allocation, waste reduction, consumption of fresh resources, optimum heating requirement, and optimum cogeneration potential. Mass integration techniques have been developed for addressing the material utilities of the process analysis. El-Halwagi and Manousiothakis (1989a) introduced the problem of synthesizing mass exchange networks (MENs) and developed a pinch-based targeting technique.

This dissertation follows the style of *Chemical Engineering Communication.*

This technique targeted minimum usage of external MSAs by maximizing mass exchange within the process streams. Multicomponent MENs can also be systematically synthesized (e.g., El-Halwagi and Manousiouthakis, 1989b; Alva-Argaez et al, 1999). Additional research has also been conducted on broader classes and techniques for the MEN problem. These include genetic algorithms (e.g., Garrard and Fraga, 1998; Xue et al., 2000), reactive MENs (e.g., El-Halwagi and Srinivas, 1992; Srinivas and El-Halwagi, 1994a), the simultaneous design of mass- and heat-exchange networks (e.g., Srinivas and El-Halwagi, 1994b and Sebastian et al, 1996), the synthesis of MENs with fixed-load removal (e.g., Kiperstok and Sharratt, 1995), MENs with variable supply and target compositions (Garrison et al., 1995), fixed-cost targeting techniques (e.g., Hallale and Fraser, 1997, 1999), MEN with flexible performance (Zhu and El-Halwagi, 1995; Papalexandri and Pistikopoulos, 1994), controllable MENs (Huang and Edgar, 1995; Huang and Fan, 1995) and batch MENs (Foo et al., 2004).

Another important category in optimizing material utilities and waste discharge is the identification of recycle/reuse strategies. An important variation of MENs, wastewater minimization, was introduced by Wang and Smith (1994). They proposed a graphical approach to target the minimum fresh water consumption and wastewater discharged by the transfer of contaminants from process streams to water streams. Dhole et. al. (1996) and El-Halwagi and Spriggs (1996) addressed the recycle/reuse problem through a source-sink representation. Polley and Polley (2000) proposed a set of rules for sequencing mixing and recycle options. Additionally, Sorin and Bedard (1999) proposed an algebraic method called the Evolutionary Table which is based on locating

the global pinch based on mixing source streams with closer concentration differences first, and then going to the stream with the next nearest concentration. Hallale (2002) developed the Water Surplus Diagram with a graphical representation of purity versus flowrate graphical representation. The concept of surplus was also used by Alves (1999) for the application of hydrogen recovery systems in refineries. Both methods rely on extensive calculations to create the surplus diagram in order to target minimal consumption of resources (water in the case of Hallale (2002) and hydrogen in the case of Alves and Towler (2002). Targeting techniques have been developed to identify minimum usage of fresh resources using a cascade diagram (Manan et al., 2004) and using pinch-based composite representations (El-Halwagi et al., 2003). Mathematical programming techniques have also been used to solve the recycle/reuse problems (Savelski and Bagajewicz (2000, 2001)) including multicomponent systems (e.g., Alva-Argaez et. al. (1999), Benko et. al. (2000) and Dunn et. al. (2001a, 2001b)), and systems with interception (El-Halwagi et al., 1997). Additionally, similar methods have been developed for unsteady-state and batch systems (e.g., Wang and Smith, (1995), Almato et. al. (1997), and Zhou et al. (2001)). Parthasarathy (2004) introduced a design procedure to optimize the recovery of water and energy via recycle/reuse. Review of mass integration can be found in literature (e.g., El-Halwagi, 1997; El-Halwagi and Spriggs, 1998, Dunn and El-Halwagi, 2003).

The process integration techniques utilized for addressing the energy utility side of the operations are mainly based on energy integration techniques. These techniques can be divided into two groups, one addressing the thermal demand (heating and

cooling) of the process, and the other one addressing combined heat and power applications (e.g., heat pumps, heat engines, cogeneration). For thermal utilities, numerous methods for the synthesis of heat exchange networks (HENs) have been developed. Among the graphical techniques thermal pinch techniques (Hohman, 1971; Umeda et.al. 1979, Linnhoff and Hindmarsh, 1983) can be used to identify minimum heating and cooling utility requirements for the process. Linnhoff and Flower (1978) introduced an algebraic technique for the same purpose. The concept of temperature interval diagram and cascade diagram was developed to address the same problem. Another graphical method called the Grand Composite Curve (GCC) was developed by Linnhoff et. al. (1982) to determine the optimum levels and types of heating and cooling utilities. For cogeneration targeting, Dhole and Linnhoff (1993) introduced a method of coupling the concept of exergy with existing graphical energy integration technique. The technique examines multiple processes at once by constructing overall composite source and sink profiles through the individual process GCC's. Raissi (1994) introduced TH-Shaftwork targeting model for cogeneration targeting. Harell (2003) introduced single stage graphical technique for the determination of optimum cogeneration potential before the detailed design.

Mavromatis (1996) and Mavromatis and Kokossis (1998a, 1998b) introduced Turbine Hardware Model for targeting the cogeneration potential. Varbanov et. al. (2004a) introduced improved turbine hardware model by considering changes in turbine efficiency with the changing load. They utilize their improved model in modeling and optimization of utility systems. Later Varbanov et. al. (2004b) utilized the improved

turbine hardware model and industrial R-curve concept in analyzing the total site utility systems. The R-curve which derived by cogeneration efficiency vs. heat-to-power ratio was introduced by Kenney (1984) and later developed by Kimura and Zhu (2000).

There exists some literature on the integration of steam driven chillers into the total site utility system to enhance the efficiency of the total cogeneration. Zanis (1986) presented the idea of integration of cogeneration and refrigeration systems into total energy facilities. Kirsner (1986) presented a comparison study between the steam driven chillers and the centrifugal chillers for process industries. Poredos et. al. (2002) studied the energy efficiency of chillers in a trigeneration plant. In this study they presented an exergetic efficiency determination different kind of chillers. Bruno et. al. (2000) proposed an optimization of energy plants including absorption chillers. They propose a modeling and optimization tool to study the economic viability of absorption chillers integration in energy systems for different process conditions.

Nishio and Johnson (1977) formulated an algorithm for the expansion problem for a steam and power plant as a synthesis problem with constraint. They introduced a simple LP logarithm for solving multi-time period linear programming problem. Nishio et. al. (1980) introduced a simple LP algorithm for design method of utility system. Papoulias and Grossman (1983) proposed an MILP (mixed integer linear programming) approach for total site utility systems by considering constant process heat and power demand for a site. Bruno et. al. (1998) introduced an MINLP (mixed integer nonlinear programming) approach with some improvements with regard to complex steam turbines and more accurate steam property characterization.

In spite of the usefulness of the above-mentioned techniques they suffer from one or more of the following limitations:

- Addressing one subsystem of the utilities without interaction with the rest of the material and energy utilities. For instance, in many cases material and energy utilities are interdependent. It is important to integrate and reconcile both categories.

- Isolation of the utility system from the core processing units. This is normally achieved by subjugating the utility system to the core-process requirements without establishing a tradeoff. Nonetheless, in a typical process, the utility side of the operation and the core process units are interdependent. For instance, the process has demand for water, steam and power, which are supplied by the utility system. At the same time, the process can supply the utility system with process fuel and waste heat. Additionally, the design and operation of core process units may be adjusted so as to optimize the combined performance of the process and the utility system. This interaction is schematically illustrated by figure 1.1.

- Ineffective formulation and/or solution technique: while the overall process and utility integration may be formulated as a mixed-integer nonlinear program, the global solution of such programs may be an elusive task. At present, there is no general-purpose formulation and computationally efficient global-solution technique for the optimization of process and utility systems.
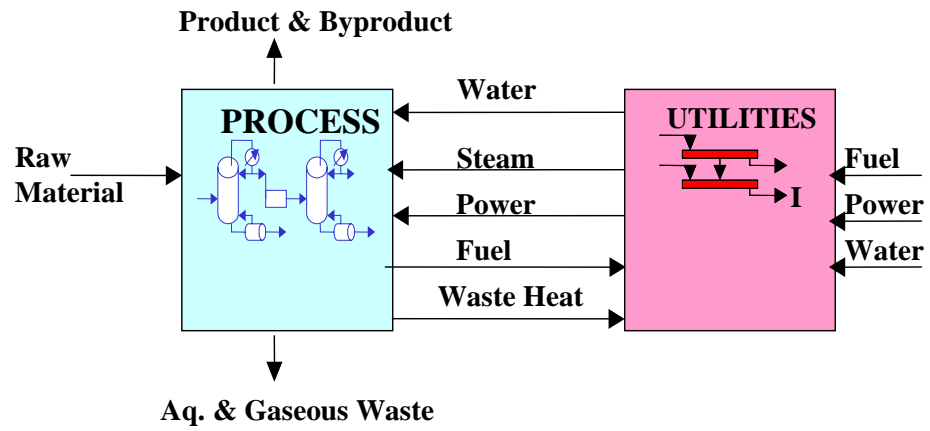
Figure 1.1: Interaction between Core Process and Utilities

CHAPTER II

PROBLEM STATEMENT

The problem to be addressed by this dissertation is formally introduced here. Consider a process with:

- A set of heating demands expressed as quantities and temperature levels. The set is given by HEATING_DEMANDS = {QH_i | i=1,2,…,$N_{Heating}$}. Each heating demand is to receive heat to increase its temperature from a supply temperature ts_i to a target temperature tt_i.

- A set of cooling demands expressed as quantities and temperature levels. The set is given by COOLING_DEMANDS = {QC_i | i=1,2,…,$N_{Cooling}$}. Each cooling demand is to receive heat to reduce its temperature from a supply temperature TS_i to a target temperature TT_i.

- A certain requirement for electric power, P.

- A certain requirement for external fuel, F.

- A set of demands for material utilities: MATERIAL_UTILITIES = {$M_{u,v}$| u = 1,2,…,$N_{Material\ Utilities}$, v=1,2,…,N_{Sinks}} where $M_{u,v}$ is the flowrate requirement of the uth utility in the vth sink, $N_{Material\ Utilities}$ is the number of material utilities (e.g., an MSA, non-heating steam, water, etc.) and N_{Sinks} is the number of process units (sinks) that employ these material utilities. Each sink has constraints on the concentration of the material utility expressed as:

$$z_{u,v}^{min} \leq z_{u,v} \leq z_{u,v}^{max}$$

- A set of process sources that may be considered for recycle to the process sinks to replace some of the fresh usage of material utilities. The set is given by PROCESS_SOURCES = {W_p| p = 1,2,...,N$_{Process\ sources}$ } where W_p is the flowrate of the pth process source. The composition of the pth source is given by y_p.

- A set of process wastes. The set is given by PROCESS_WASTES = {W_w| w = 1,2,...,N$_{Process\ wastes}$ } where W_w is the flowrate of the wth process waste. A subset of this set involves the combustible wastes that may be used to supplement the usage of the external fuel.

The objective is to integrate, reconcile, and optimize the utility usage for the process. Towards this end, the following important design challenges to address:

- What are the optimum quantities and levels of heating and cooling utilities?

- Is there a potential for power cogeneration? What is the cogeneration target?

- What is the optimum scheme for recycling/reusing process sources?

- What is the optimum strategy for waste discharge?

- Can some of the combustible wastes be used instead of fresh fuel? To what extent? Where?

- What refrigeration technologies may be used (e.g., cooling towers, refrigeration cycles, absorptive refrigeration, etc.)?

- What are the necessary process modifications that are required to trade off the core-process units with the utility system?

The abovementioned design challenges are highly interactive, complex, and combinatorial. Therefore, it is necessary to develop a systematic and generally applicable approach to the problem.

This work introduces a hierarchical procedure that decomposes the problem into successive stages that are globally solvable then merges the solution fragments into an overall process target. In particular, the dissertation introduces a detailed methodology for simultaneous integration of process utilities and core process units. The problem is decomposed into stages each of which is addressed using state of the art process integration techniques. First visualization tools are used to develop global insights of the system. Subsequently, mathematical programming techniques are employed to develop an optimization formulation. Global optimization techniques were also employed to improve the quality of the solution. The devised visualization tools and optimization formulations constitute effective tools to aid engineers in determining the optimum fuel, recycle/recovery, heating/cooling requirement, utility selection, and power cogeneration satisfying process requirements.

CHAPTER III

GRAPHICAL HIERARCHICAL APPROACH

3.1 OVERALL APPROACH

In this chapter, a visualization approach is adopted to address the problem stated in Chapter II. In any manufacturing process the required process data are usually available or retrievable. From the process data mass integration and heat integration analysis are performed. Mass integration analysis precedes heat integration analysis due to the assumption that for any waste material the recoverable material value is higher then the recoverable thermal value. Also it is important that both mass and heat integration analyses are performed before considering the cogeneration study, because, the results from this analysis would affect the availability of steam for the cogeneration.

Mass integration analysis provides insights into optimum recycle/reuse opportunity which results in optimum raw material consumption and optimum waste disposal. Also, mass integration analysis provides information regarding any existing fuel substitution opportunity. Based on the mass integration analysis, it is possible to evaluate the steam demand for mass purposes.

After mass integration, heat integration analysis is performed. It provides the minimum heating and cooling demand of the process. Additionally, heat integration analysis provides information regarding optimum conditions of the steam (i.e. required temperature and pressure of the steam). From the results of mass and heat integration analysis and also from the given process information, total steam demand of the plant for

heat, mass, and other process requirements can be determined. The result of these analyses is that the steam header balance for the total manufacturing site can be generated.

From the header balance, using extractable work method, optimum cogeneration potential for the manufacturing process can be targeted. First the cogeneration potential is targeted considering the heating demand for the steam only. Extractable work method also provides information regarding the presence of excess steam (steam produced within the process) in the process. Reconciliation of excess steam is targeted by supplying the cooling demand of the process via steam using any absorption refrigeration technique. The results of these analyses provide a balanced cooling, heating and cogeneration with optimum material recovery. The flowchart of the proposed procedure is schematically shown by figure 3.1.

3.2 MASS INTEGRATION AND HEAT INTEGRATION APPROACH

The process plant utilizes the high value feedstocks to generate products, byproducts, and wastes. Waste streams may contain some components which if recovered can be utilized to reduce the consumption of higher value material. This is identified as material value of the waste and represented as M_{value}. Furthermore, some waste streams can be burned to produce some thermal value, which would result in reduction of fresh fuel consumption. Recoverable thermal value from waste is represented as T_{value}. Waste streams that are neither recovered as recyclable materials nor utilized for thermal value are discharged or sent to waste treatment. Figure 3.2 illustrates a plant with waste utilization alternatives.

Process Requirements and Data

Mass Integration ⬌ **Heat Integration**

| Raw Material Recycle | Min Waste | Fuel Substitution | Non-heating Demand | | Heating Demand | Utilities Selection | Cooling Demand |

Header Balance

CHP Extractable Work

Cogen Target   **Excess Steam**   **Updated Header Balance**

Steam Reconciliation
Optimal Utilities
Cogen Potential (CCHP)

Solution Strategy
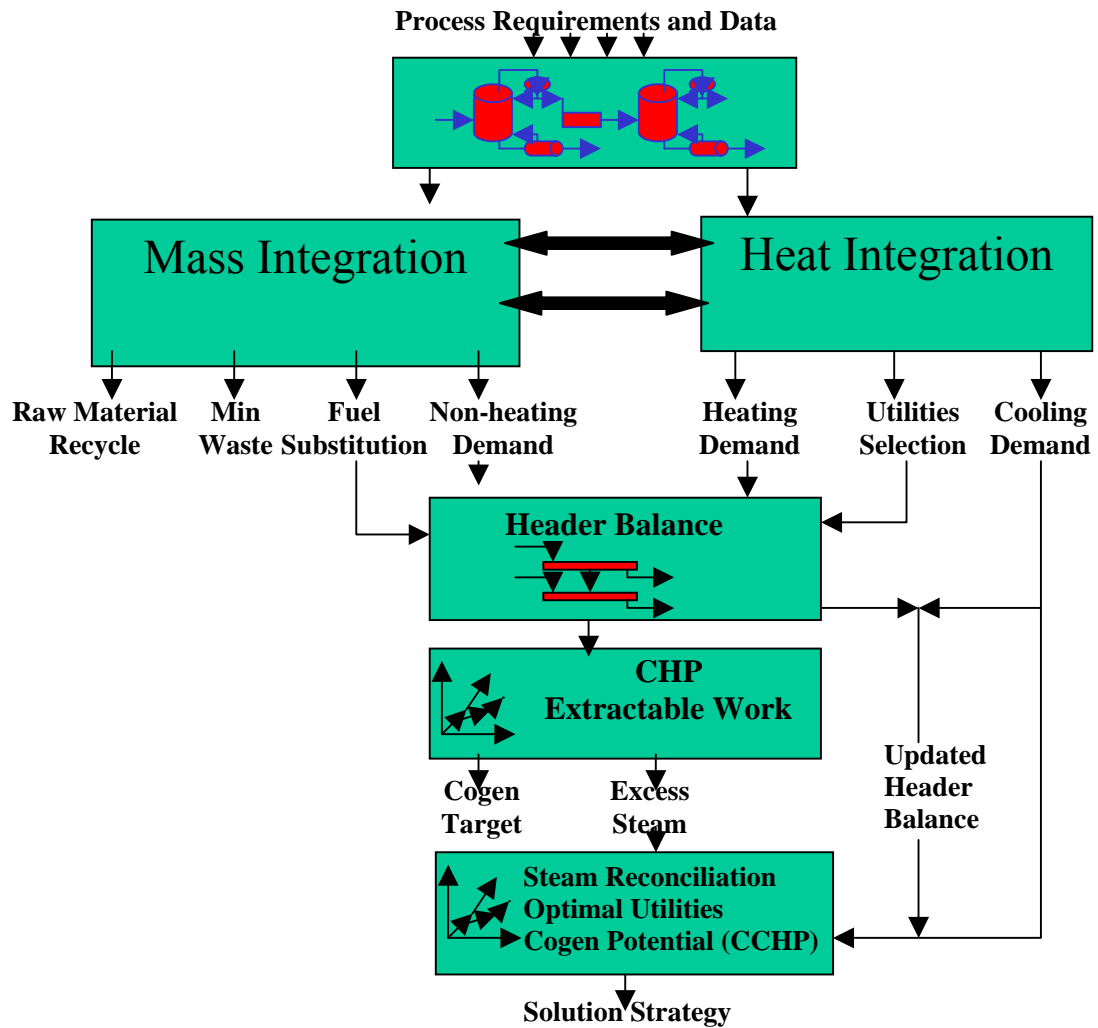
Figure 3.1: Overall Approach (Graphical Technique)
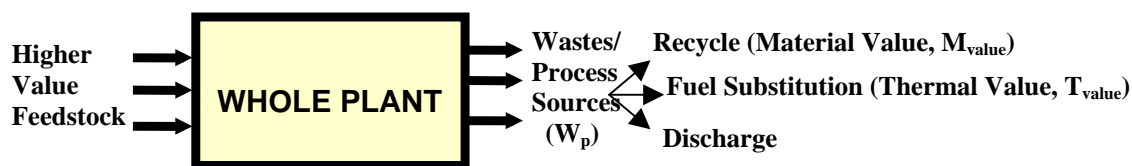
Figure 3.2: Overall Plant with Waste Utilization Alternatives

The waste streams that are generated can be targeted for recycle and recovery via mass integration techniques. The unrecycled materials or process wastes can be evaluated for any thermal utilization or sent for waste treatment or can be discharged into the atmosphere. Figure 3.3 illustrates the recovery network.
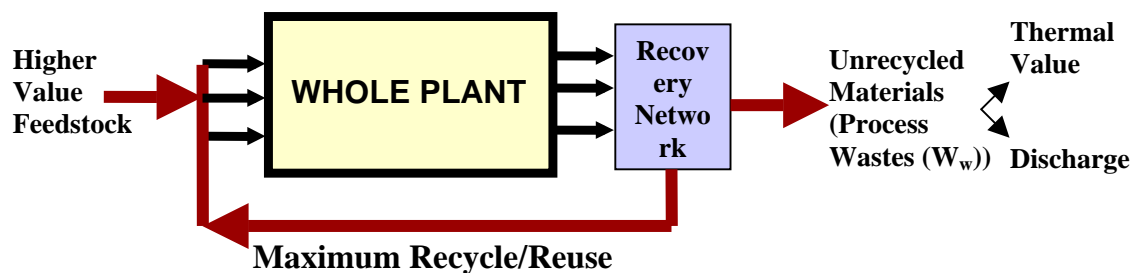
Figure 3.3: Recovery Network

Typically, the value of the waste streams when used to replace fresh feedstocks is higher than its thermal value. This observation is based on the rationale that if the thermal value of a chemical is higher than its value as a feedstock, it would be used as a fuel. Therefore, first an attempt is made to recover as much material value as possible

from the waste streams beginning from low investment options to higher investment options. Figure 3.4 illustrates overall mass integration techniques.
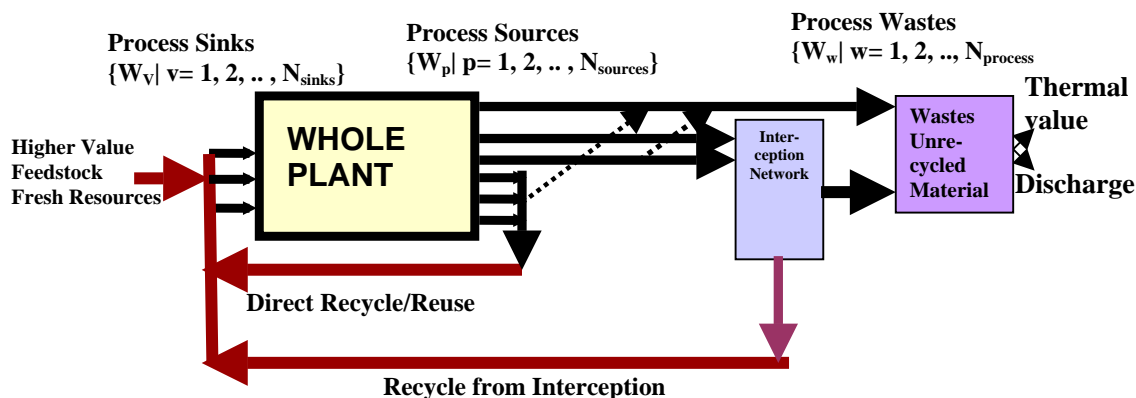
Figure 3.4: Overall Mass Integration Techniques

Keeping these issues in mind, first the opportunity for direct recycle is evaluated. An effective approach to the identification of waste recycle targets is the material-recycle pinch diagram shown by figure 3.5. For the direct recycle evaluation the waste streams are considered as process sources. A set of process sinks are identified which can utilize these process sources. Apart from the process sources, the sinks also get additional required amount from fresh sources. Unused sources become wastes.

The material recovery problem can be defined as follows:

Given:

- A set of process sources $\{W_P|\ P = 1, 2, .., N_{sources}\}$ which can be recycled and reused in process sinks. Each source has a given flowrate W_P. The composition of the Pth source is given by Y_P.

- A set of process sinks $\{W_V| V= 1, 2, .. , N_{sinks}\}$. Each sink requires a flowrate W_V and a composition of a single targeted species, z_j^{in} , that satisfies the following constraint: $z_j^{min} \leq z_j^{in} \leq z_j^{max}$

- Available for service is a fresh external resource that can be purchased to supplement the use of process sources in sinks.
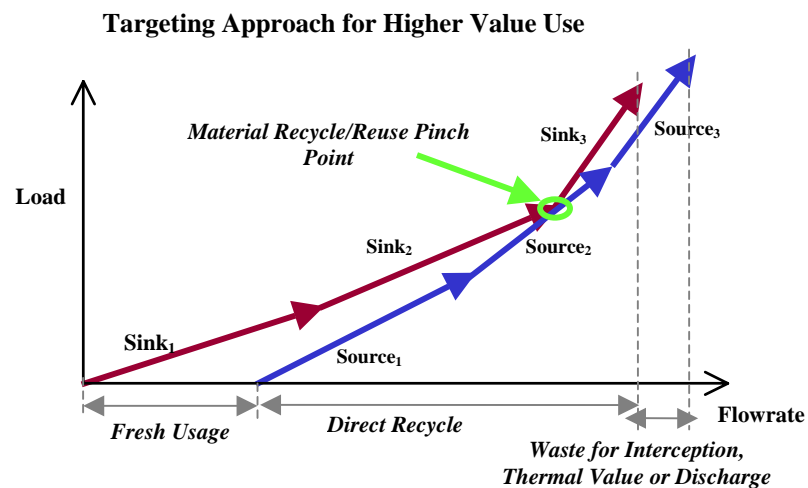
Figure 3.5: Material Recycle Pinch Diagram (El-Halwagi et al., 2003)

The objective of the waste recycle and material recovery problem is to determine the target of maximum direct recycle possible and hence determine the minimum usage of fresh resources and also the minimum waste.

Since in this section the dissertation focuses mainly on state of the art graphical techniques, for the direct recycle/reuse problem, the material recycle/reuse pinch diagram method developed by El-Halwagi et. al. (2003) is used for targeting waste recovery and disposal. This is a single stage, systematic, and graphical method for

identifying rigorous targets for recycle/reuse problem. Material recycle/reuse pinch diagram provides information regarding minimum fresh usage, maximum recycle and minimum waste disposal.

The target for minimum waste streams identified from the direct recycle/recovery analysis can be subjected for further material recovery by utilizing mass integration techniques such as species interception network (Figure 3.6). Interception denotes the utilization of separation unit operations to adjust the composition of pollutant-laden streams to make them acceptable for the sinks (Figure 3.7). The separation can be induced by mass-separating agents (MSAs) and Energy Separating Agents (ESAs, like non-heating steam). The synthesis of mass-exchange networks (MENs) (El-Halwagi and Manousiouthakis 1989b) can be utilized as a systematic technique to screen the multitude of separating agents and separating technologies to determine the optimum separation system. The synthesis of MENs will result in an optimum demand of material utilities $\{M_{u,v}|$ u= 1, 2, .., $N_{material\ utilities}$, v= 1, 2, .., $N_{sinks}\}$ in a number of process sinks N_{sinks} that utilizes that material utilities.

Recovery of material value from waste streams through species/mass interception networks may involve capital and operating costs. Therefore, a thorough cost/benefit analysis should be pursued on the available technology to be utilized. Also, an economic evaluation is required to target the amount of waste stream to be recovered and also the external mass separating agents to be utilized. For any interception device to become economically feasible the overall value of recovered waste should be greater then the overall cost of recovery.

Targeting Approach for Higher Value Use
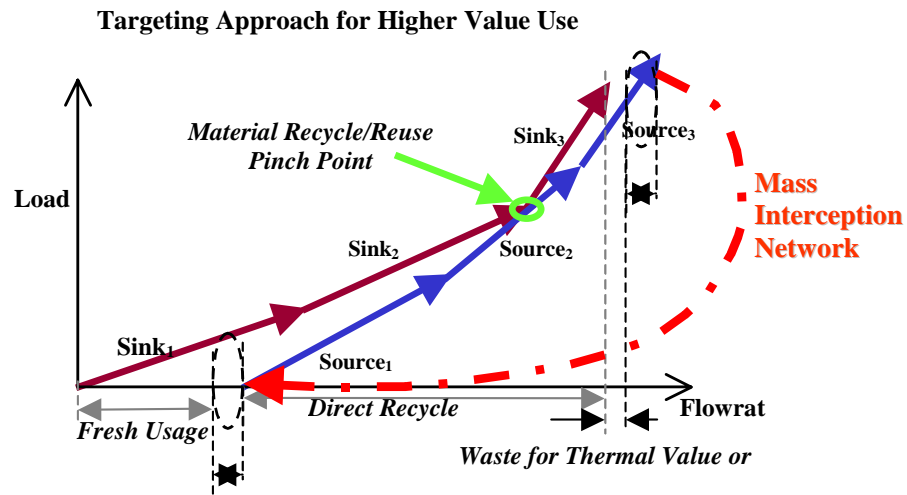
Figure 3.6: Material Recycle Pinch Diagram (Interception)
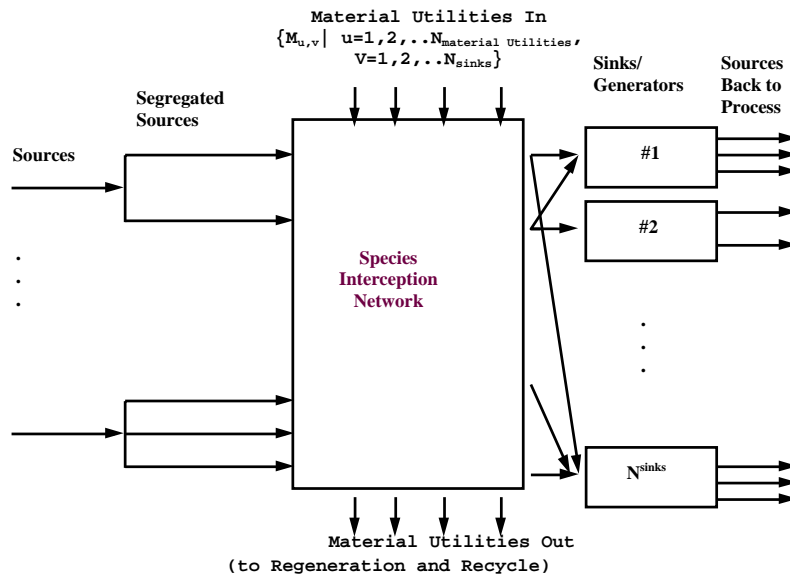
Figure 3.7: Species Interception Network

The overall material value of the recovered waste is determined by the following factors:

$$M_{value} = VM_{\text{fresh material replaced}} + VM_{\text{reduction in waste treatment}} + VM_{\text{environmental incentives}} \qquad (3.1)$$

The overall cost involves in the recovery process is determined by the following factors:

$$RM_{cost} = CM_{\text{interception unit}} + CM_{\text{external MSA}} + CM_{\text{operation \& maintenance}} \qquad (3.2)$$

The interception network can only be feasible for recovering material value in the cases where $M_{value} > RM_{cost}$

As mentioned earlier, the waste streams may have recoverable material value as well as thermal value. The direct recycle/reuse pinch method results in recovery of material value at no or low cost. When an interception network is used, a certain amount of investment is required for the separation units and also for the external material utilities. Higher recycle and recovery would result in lesser amount of waste targeted by the MEN due to diminishing economic value. Although thermal value is assumed to be lower then the material value of a given waste streams, the investment required to recover the thermal value may also be lower than that for material value. Although interception is feasible when considered independently to recover waste for material value up to certain target, the thermal value recovery may become more economically beneficial beyond a certain extent of recovery. Therefore, the optimum extent of interception is determined in conjunction of determining the optimum extent of thermal value recovery. The interception network should be utilized as long as the net value of the recovered materials is higher than their thermal values.

For targeting thermal value recovery from waste, it is important to identify the thermal requirement of the process. Heat integration techniques are utilized to determine the optimum heat load to be added/removed by certain utilities. The process has certain heating demands $\{QH_i | i= 1,2, .., N_{Heating}\}$ and a number of cooling demands $\{QC_i | i= 1, 2, .., N_{Cooling}\}$. The heat integration techniques determine the requirement of minimum heating utility and minimum cooling utility and also determine the possibility of integrated heat exchange opportunities. Any general heat integration technique such as heat pinch diagram, cascade analysis or grand composite curve can be utilized.

As this Chapter is primarily focused on graphical techniques that are being used, the heat pinch diagram (Figure 3.8) is chosen for getting the heating values. The heat pinch diagram shows the requirements of minimum heating and cooling utilities by maximizing the integrated heat exchange within the process.
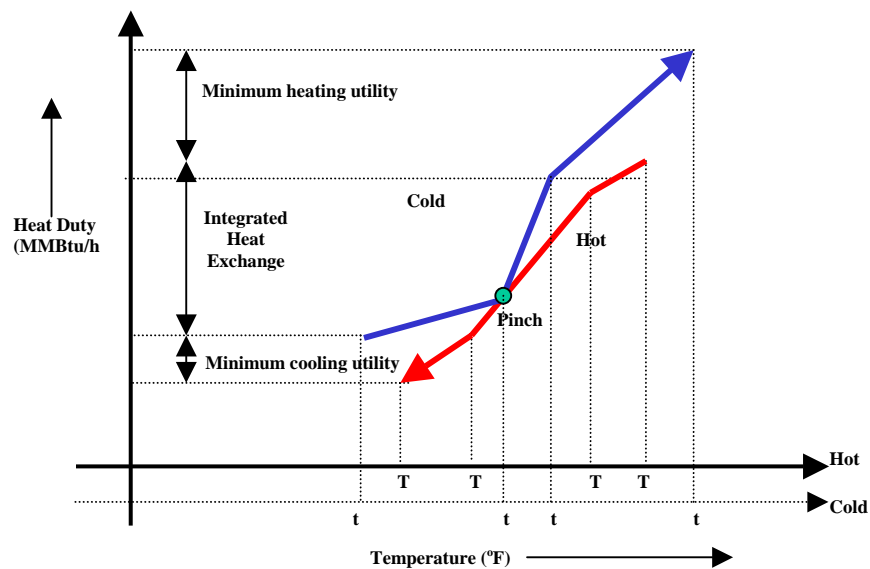
Figure 3.8: Heat Pinch Diagram

The waste streams should be targeted for thermal value recovery, if and only if, the process has a demand for heating utility. The combustible waste can be burned in appropriate units (e.g., incinerators), and the heat generated can be recovered by waste heat boiler (WHB) or heat recovery steam generator to produce steam to supply the heating demand of the process. Similar to the material value recovery from waste through interception network, the thermal value recovery system also has some cost and benefits. Therefore, a thorough cost/benefit analysis should be pursued on thermal value recovery from the waste streams.

The overall thermal value of the recovered waste is determined by the following factors:

$$T_{value} = VT_{high\ cost\ fuel\ substituted} + VT_{reduction\ in\ waste\ treatment} + VT_{environmental\ incentives} \quad (3.3)$$

The overall cost involves in the recovery process is determined by the following factors:

$$RT_{cost} = CT_{inceneration\ unit} + CT_{external\ fuel} + CT_{operation\ \&\ maintenance} \quad (3.4)$$

The thermal value recovery system is only feasible where the T_{value} is greater than RT_{cost}. The cost/benefit analysis will provide an insight on the extent of the waste that can be targeted for thermal value recovery.

As discussed earlier, a coupled cost/benefit analysis for both material and thermal value recovery should be undertaken. Independent analyses result in determination of feasible region or feasible target for thermal and material recovery network. But within the feasible region of one, optimum solution may be identified to tradeoff the optimum extent of recovery for mass versus thermal values. To understand

such result, or to target the extent of recovery of material and thermal value from waste from overall economic impact analysis, the two independent cost/benefit analyses should be compared.

The comparative cost/benefit analysis of material and thermal value recovery system results in identifying targets for optimum mass interception, thermal value recovery, and minimum waste. The minimum waste identified should be send for waste treatment or discharged into the environment according to given conditions and regulations.

Based on the foregoing steps, the mass and heat integration techniques result in:

- Maximum recycle

- Minimum fresh consumption

- Minimum waste disposal

- Optimum material value recovery from waste

- Optimum thermal value recovery from waste

- Fuel substitution opportunity

- Portion of steam demand for the mass purposes

- Determination of minimum heating and cooling demand

- Availability of process steam from waste

These aspects show how the mass and heat integration analysis provide important information regarding optimum utilization of resources and recovery of material and energy utilities. An overview of the proposed procedure is summarized through a flowchart in figure 3.9.
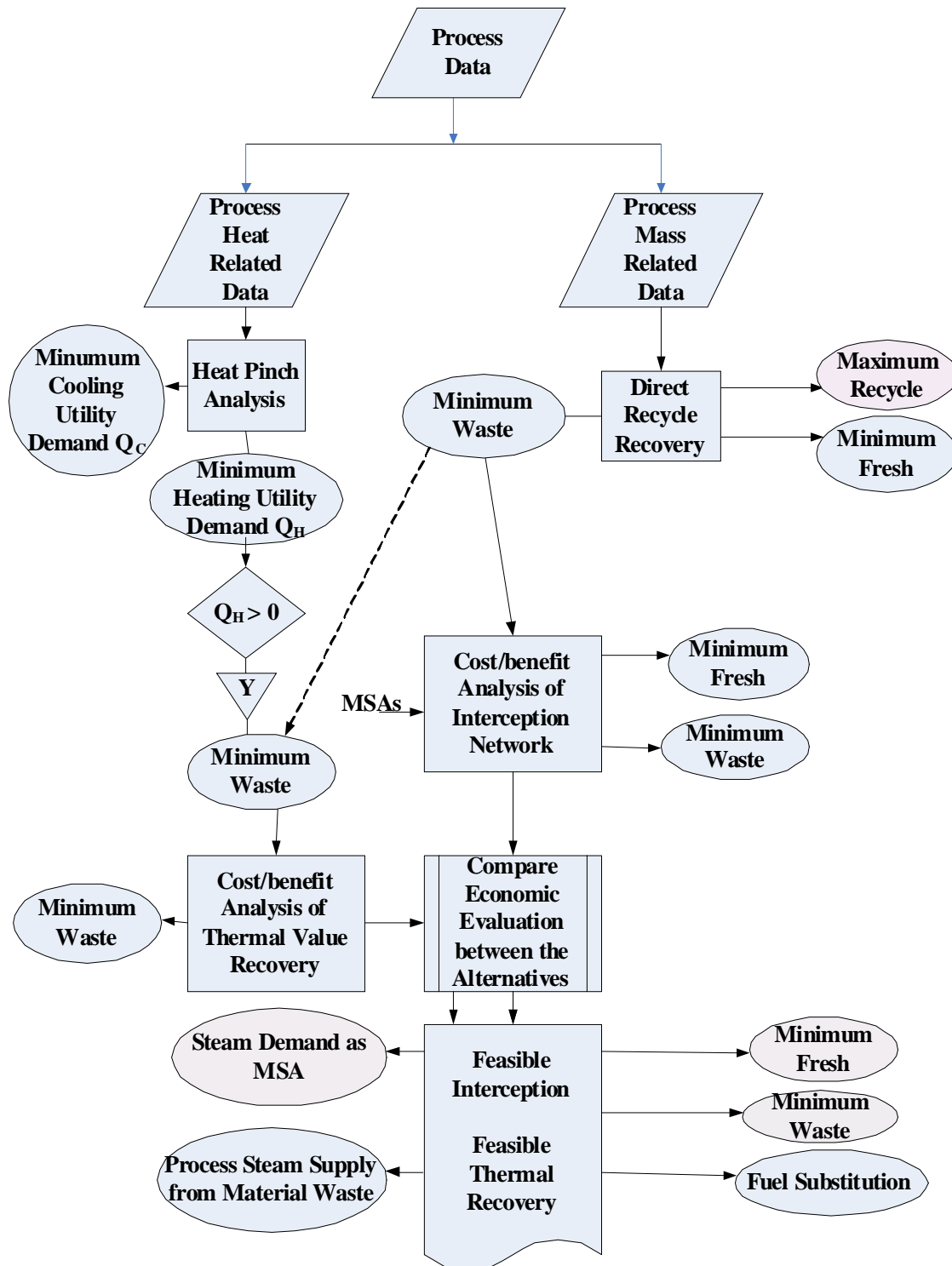
Figure 3.9: Hierarchical Mass and Heat Integration Analysis

3.3 STEAM HEADER BALANCE

The steam header balance is an important step in tracking steam levels, sources, and demands. In a process plant, steam may be required at different levels of pressure and temperature. Steam may be required for heating, non-heating and several mass purposes. There are several possible ways of generating steam for meeting the steam demand. In a typical process plant, steam is generated in a central utility/boiler plant by burning coal or gas as fuel. Then, the steam is transferred to different process areas through steam headers according to the requirement. Also, steam can be purchased from an outside source (e.g., a nearby power plant) to meet the plant demand. Additionally, there may be several possibilities within the process to generate steam. Any opportunity to produce steam within the process is extremely significant. It provides positive economic effects and may also render the plant more environment friendly. Generation of steam using process sources reduces the fuel consumption of the process. It also affects the cogeneration efficiency, waste discharge (including green house gases GHGs), waste disposal, and waste treatment cost as will be discussed in next section. Some of the possibilities of generating steam within the process are:

- Steam may be generated within the process as a byproduct of a exothermic reaction system

- There may exists hot streams that require cooling and generate steams to satisfy this need

- Steam may be generated within the process by heat recovery steam generator (HRSG) using the hot exhaust gases from different equipments

- Process plants usually lose a lot of heat through stack. This heat can be recovered economically via HRSG in producing steam.

- Steam may be generated using waste heat boiler (WHB) where heat is generated through burning combustible waste in the incinerator. This makes the steam production as a function of mass integration techniques.

In the chemical manufacturing processes the utility steam is usually available through different fixed steam headers according to the process requirements. Steam physical properties vary between different headers but they are usually fixed for any given header.

For developing the balance of the steam headers, steam generated within the process and steam demand within the process are considered. As discussed earlier, both mass and heat integration analyses affect both steam generation and steam demand for the whole process. Generally, steam generated and required by the processes is allocated through a system of steam headers.

The steam header represents the levels at which steam is required by the process for various purposes. The levels are determined by the process requirements of steam for mass, heat and other purposes. In a chemical manufacturing process, steam is mainly utilized for heating. Different levels of steam requirement for heating purposes can be targeted through Grand Composite Curve (GCC) analysis. In addition to the minimum heating and cooling utility demand GCC graphically provides important insight into the levels of required steam (figure 3.10). A typical steam header system is shown in figure 3.11.

Each header level has steam supplied to it through generation within process and each header level has to satisfy a certain process steam demand. Surplus and deficit header levels can be determined by performing steam balance against each header level.
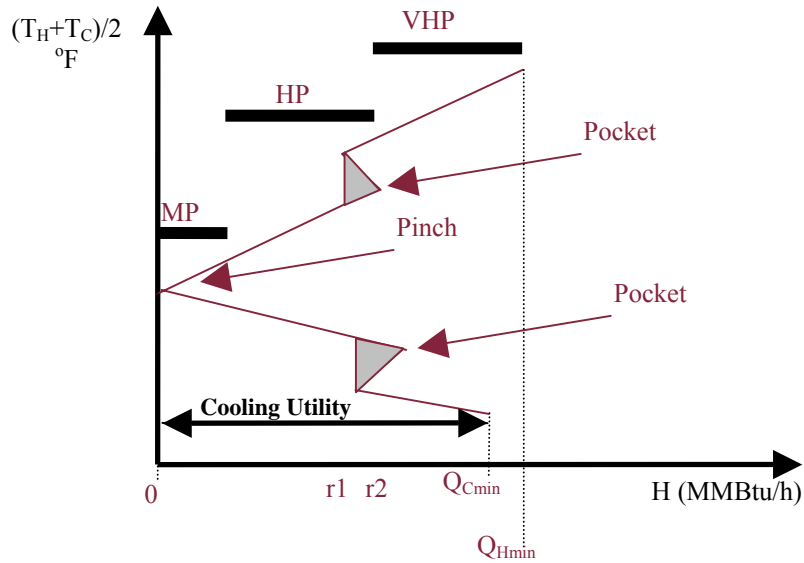
Fig 3.10: Grand Composite Curve
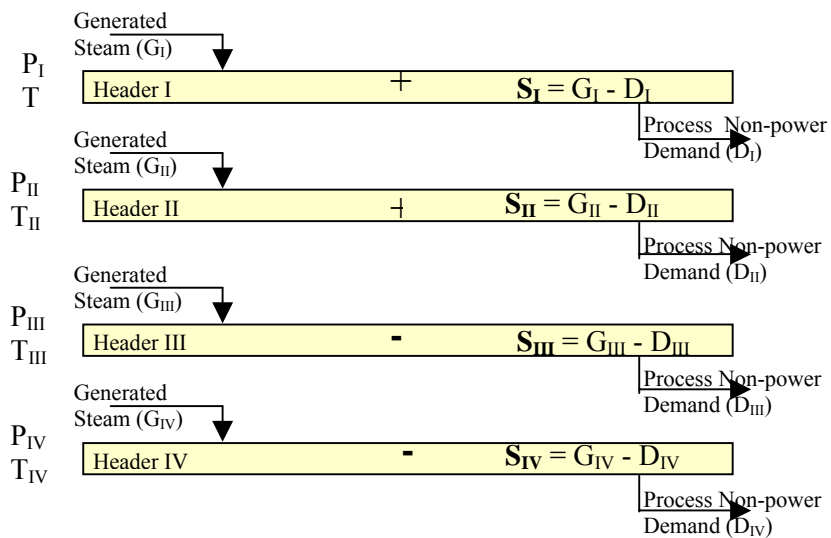
Fig 3.11: Typical Header System

The generated steam header system provides important information regarding whether the process by itself can generate enough steam to satisfy required steam demand. It also provides insights regarding additional steam requirements if any. This steam should be generated by burning external fuel or should be purchased directly from outside sources. In summary, the steam header system provides the following valuable aspects:

- Existence of surplus steam at different levels

- Requirement of steam at different levels

- Necessity of outside steam

- External fuel requirement

- Possible cogeneration opportunity (discussed in detail in next section)

Steam header balance is optimized through mass and heat integration analyses and also from process data. Mass integration provides information regarding mass demand of steam and also about the fuel substitution opportunity. On the other hand, heat integration analysis provides the required heating demand, and determines the different levels of the steam header system. The approach of steam header generation is illustrated through figure 3.12.

3.4 ENERGY INTEGRATION APPROACH

In addition to demand for material utilities, heating and cooling utilities, and fuel the process plant also has a demand for electricity or power. Energy integration techniques are utilized to address the power consumption issue for a process plant. Power consumption is affected by the cogeneration potential of the plant.
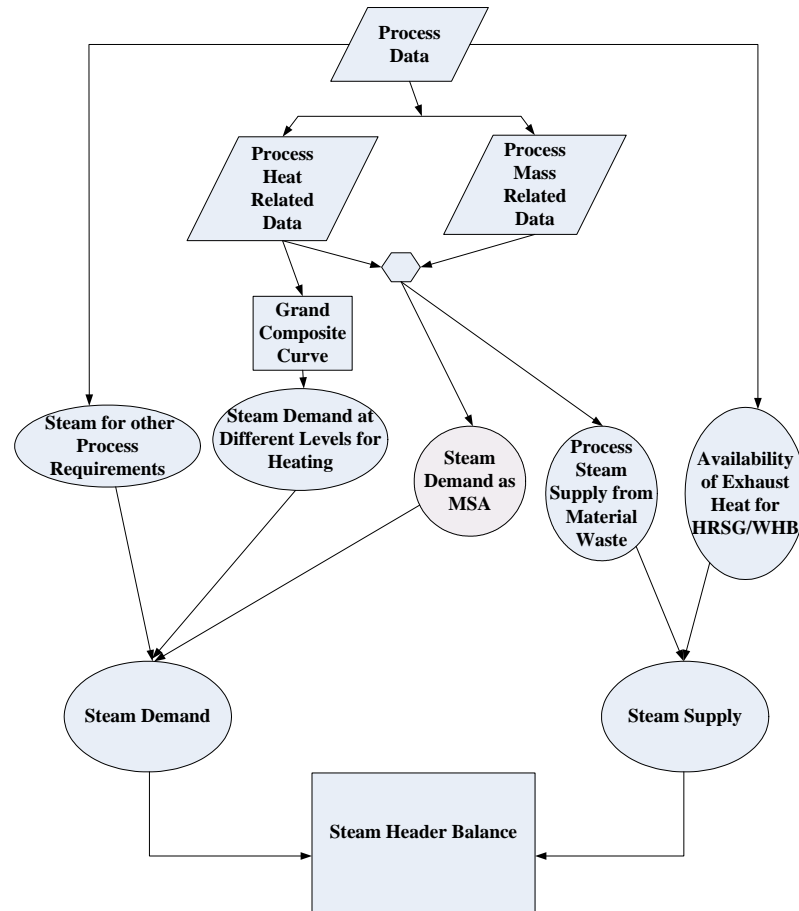
Figure 3.12: Steam Header Balance from Mass, Heat Analysis and Process Data

The steam headers presented in the previous section not only provide information about steam and fuel requirement, but also provide very important insight into the cogeneration capability of the process or the cogeneration capability of central utility/boiler plant.

As mentioned earlier, the process plant has steam demand at various pressure and temperature levels. Within a steam header system, the temperature and pressure of each

header are fixed. Steam can be passed from any higher-pressure header to any lower pressure header according to the process requirement. From the higher pressure header steam can be passed through a pressure relief valve (PRV) to supply it to the required lower pressure header. Such relief involves useless loss of energy. For cogeneration purpose, steam is passed through a turbine to reduce the pressure according to the requirement, and at the same time generate power.

Within a steam header system, the temperature and pressure of each header are known values. With this information, specific enthalpy of the steam at a header condition can be determined. Also from the header balance, required mass flow rate of steam within headers are known. From the header balance, steam enthalpy and required flowrate, optimum cogeneration potential can be targeted. First, the cogeneration potential is targeted considering only the plant heating demand for steam. Extractable work method (Harell, 2004) is utilized for targeting cogeneration potential . Now graphical extractable work method will be introduced here briefly.

3.4.1 EXTRACTABLE WORK METHOD

Extractable work method is based on the enthalpy difference between actual inlet and outlet condition (i.e. temperature and pressure) of the turbine. Since turbines are placed between steam headers with known temperature and pressure, this method is convenient then the Mollier diagram method. In Mollier diagram method it is required to calculate the enthalpy at isentropic condition at outlet pressure. Determining enthalpy at isentropic condition sometimes become cumbersome. Figure 3.13 illustrates the difference between extractable work method and Mollier diagram.

30
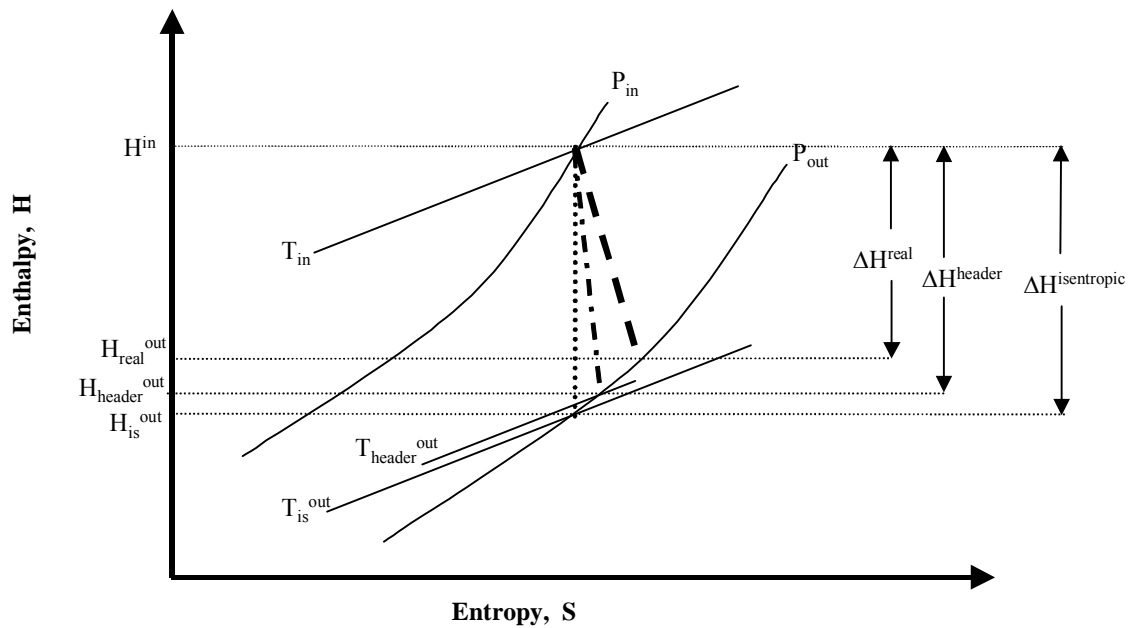
Figure 3.13: Mollier Diagram with Header and Isentropic Outlet Condition

From the figure 3.13, enthalpy difference between turbine inlet and outlet condition is

$$\Delta H^{header} = H^{in} - H^{out}_{header} \tag{3.5}$$

Here, ΔH^{header} is the specific enthalpy difference between the turbine inlet and
outlet header

H^{out}_{header} is the enthalpy at the outlet header temperature and pressure

H^{in} is the enthalpy at the inlet header temperature and pressure

Now, the efficiency term is utilized to relate the header difference to the actual enthalpy
difference. The efficiency term is defined by the following equation.

$$\eta_{header} = \frac{\Delta H^{real}}{\Delta H^{header}} \tag{3.6}$$

Here, η_{header} is the efficiency of the system,

ΔH^{real} is the actual enthalpy difference,

ΔH^{header} is the enthalpy difference between header conditions.

Form equation (3.6) and equation (3.7) we get,

$$w = \Delta H^{real} = \eta_{header}\left(H^{in} - H^{out}_{header}\right) \tag{3.7}$$

Here, w is the specific power produced by the turbine.

The actual power can be determined by multiplying the steam mass flowrate passing through the turbine with the specific power.

$$W = \overset{\bullet}{m}\,\eta_{header}\left(H^{in} - H^{out}_{header}\right) \tag{3.8}$$

Here, W is the actual power generated by the turbine,

$\overset{\bullet}{m}$ is the mass flowrate of steam passing through the turbine

Now, the concept of extractable energy is defined by,

$$e = \eta H \tag{3.9}$$

Here, e is the extractable energy,

η is an efficiency term,

H is the specific enthalpy at a given set of conditions.

The power generation expression can be written as,

$$W = \overset{\bullet}{m}\left(e^{in} - e^{out}_{header}\right) \tag{3.10}$$

Here, e^{in} is the extractable energy at inlet condition,

e_{header}^{out} is the extractable energy at outlet header condition.

Now, combining steam mass flowrate passing through the turbine with the extractable energy term, we get the power generation term to be the difference between the inlet and outlet extractable power:

$$W = E^{in} - E_{header}^{out} \qquad (3.11)$$

Here, E^{in} is the extractable power at inlet condition,

E_{header}^{out} is the extractable power at outlet header condition.

Now, for illustrating the graphical approach for extractable work method, consider the header balance showed in figure 3.14.

Supply

| VHP | Supply – Demand = (+) |

Supply

Demand

| HP | Supply – Demand = (+) |

Supply

Demand

| MP | Supply – Demand = (-) |

Supply

Demand

| LP | Supply – Demand = (-) |

Demand
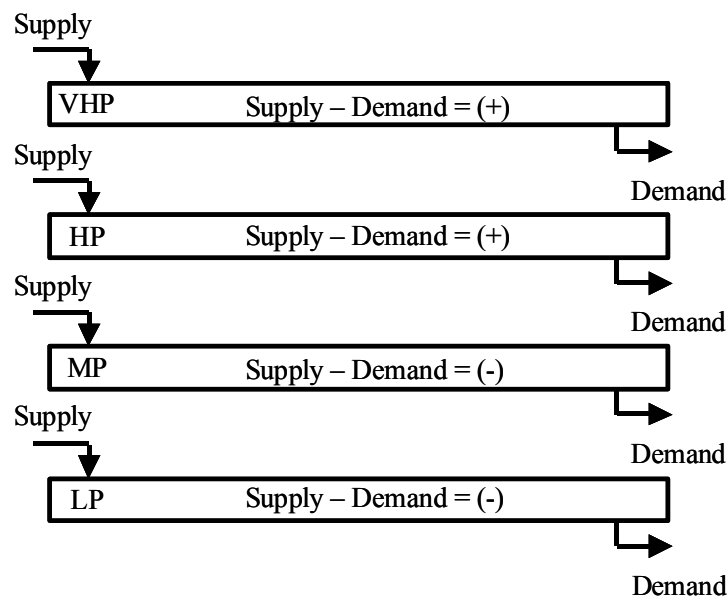
Figure 3.14: Steam Header Balance

Assuming, the magnitude of extractable power at the VHP, HP, MP and LP levels will be E1, E2, E3 and E4, and the mass flow rates will be M1, M2, M3 and M4 respectively. First, consider the surplus VHP and HP headers, which are ranked in order of ascending pressure levels. Figure 3.15 illustrates the generation of surplus composite line.
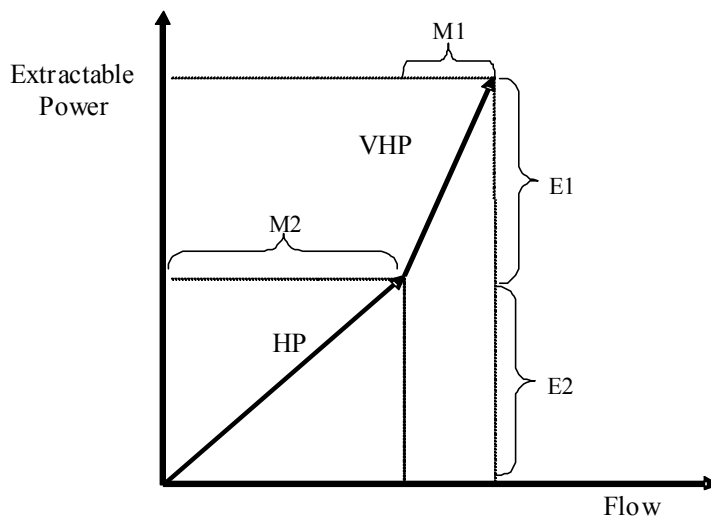
Figure 3.15: Extractable Power Surplus Composite Line

In a similar fashion, the deficit composite line can also be constructed on the same graph by plotting the LP and then MP deficit headers. The addition of the deficit composite line can be seen in figure 3.16.

After constructing the surplus and deficit lines, the cogeneration potential of the system is easily determined by shifting the deficit composite line to the right and up until it is directly below the uppermost region of the surplus line (figure 3.17). The extractable power method is shown schematically by figure 3.18.
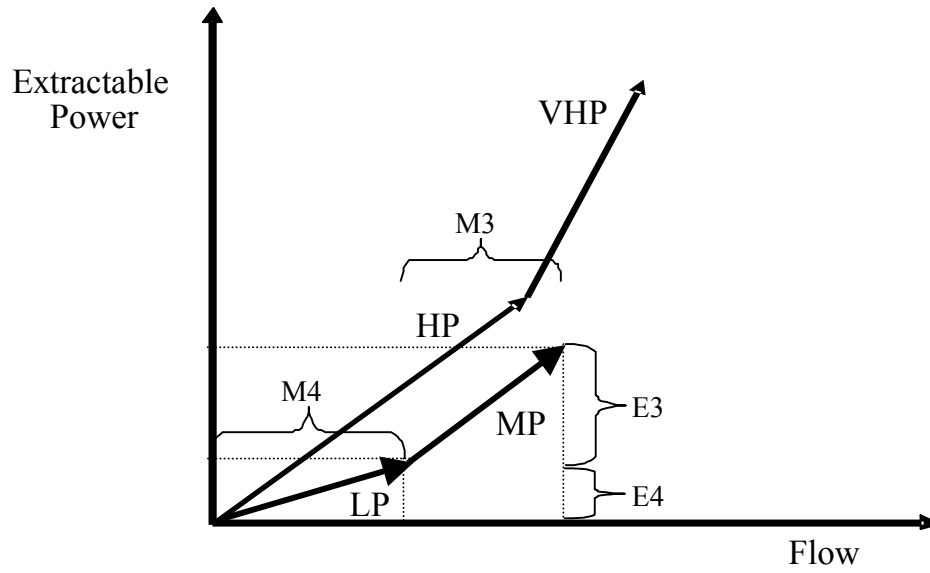
Figure 3.16: Extractable Power Surplus and Deficit Composite Lines
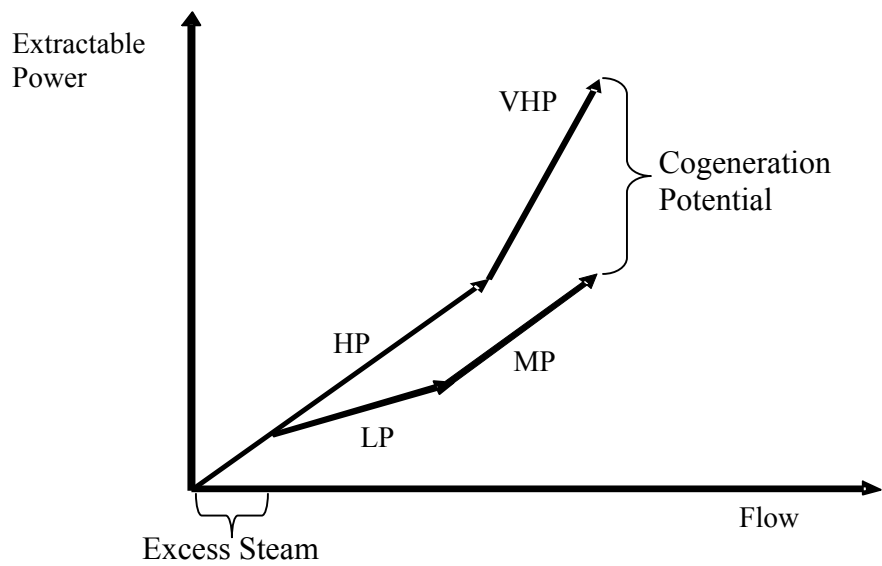
Figure 3.17: Shifted Extractable Power versus Flow Diagram
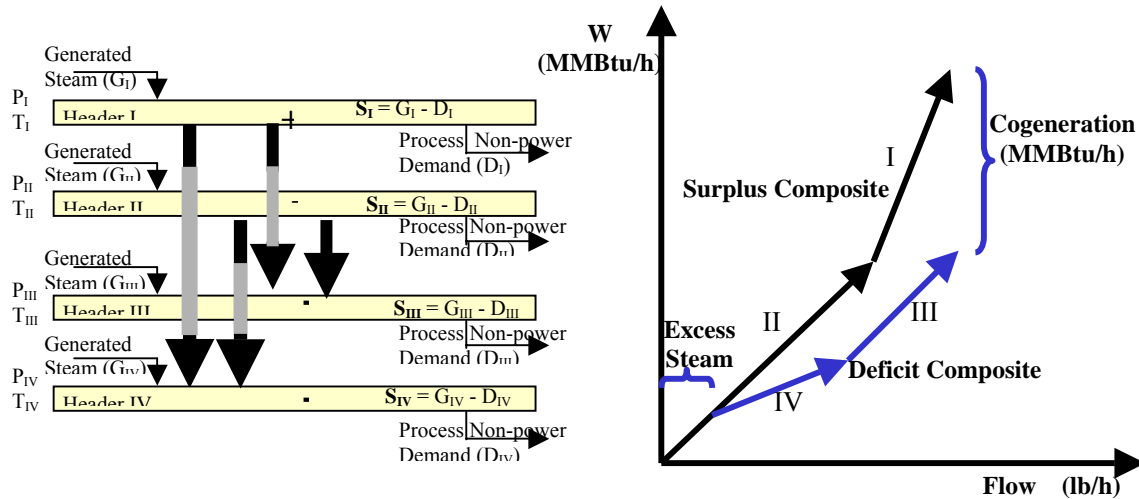
Fig 3.18: Extractable Power Method

The gap between the surplus and deficit line determines the cogeneration potential of the system. Furthermore, the region for which there is no deficit line below the surplus line indicates the amount of excess steam within the process.

The extractable work method provides the information about the optimum cogeneration potential from a given header system. Also, the method provides information regarding existence/availability of excess/free steam at any given level within the process. The information regarding the excess steam can be utilized in further optimization of the process. There can be two distinct situations regarding how to handle the availability of excess/free steam. These situations can be derived from the steam header balances. The situations are described elaborately utilizing two different cases with different scenarios.

3.4.2 CASE 1

The plant may have excess steam available from the process. In such cases, the plant satisfies all the steam demand by generating steam within the process without burning any external fuel. This case may be encountered when the mass integration analysis results in sufficient amount of combustible waste, and also the heat integration analysis results in existence of sufficient amount of exhaust heat. In such cases the excess steam is generated virtually for free (since no outside fuel is involved in producing such steam). Such free steam can be passed through a condensing turbine to generate more power for the process which will decrease the overall power consumption. Also, any such free steam can be considered for supplying the cooling load of the process by absorption refrigeration. In the absorption refrigeration case, not only the cogeneration will be enhanced, but also the amount of excess steam will reduce drastically, and also it would reduce the electricity consumption by replacing electricity driven chillers for cooling. Such production of heating, cooling and power is called trigeneration system. This is an advancement of cogeneration where only heat and power is considered. Trigeneration system can really make the economy if there exists enough free steam or exhaust heat within the process to run the absorption refrigeration system. As illustrated in the figure 3.19, the extractable work method can be readily utilized to determine the trigeneration opportunity within the process.

The absorption refrigeration system entails the use of steam as part of the cycle. As such, the cooling demand is associated with a certain amount of steam consumption, which affects the balance of the steam headers. While adding the cooling demand to the

steam header, it is necessary to check whether or not the existing excess steam can supply the whole cooling demand. If the existing excess steam is capable of supplying the cooling demand, then the result of CHP (where the excess steam is condensed through condensing turbine or exhausted to atmosphere, only heat and power) and CCHP (where excess steam is utilized for supplying the cooling demand, and any excess after that is passed through condensing turbine or exhausted to atmosphere, heat, cool and power) should be evaluated economically to find out the feasibility of either option.
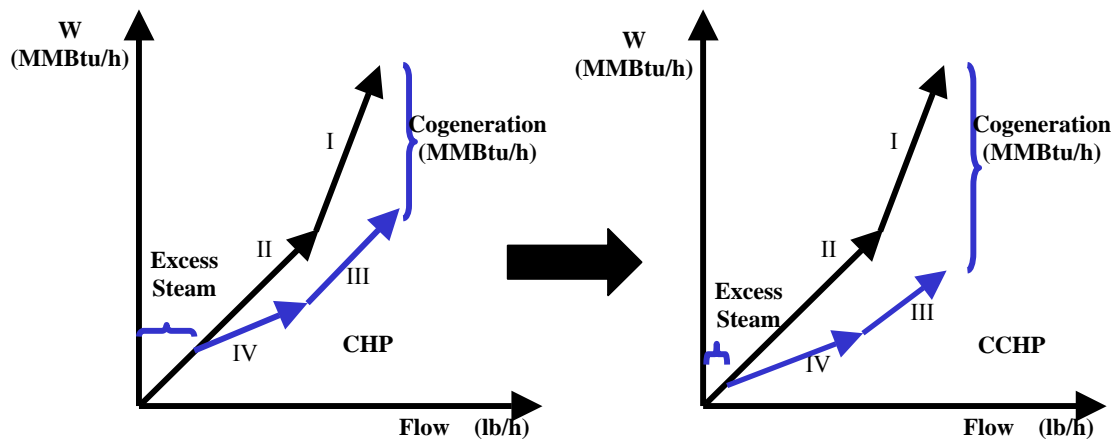
Figure 3.19: CHP versus CCHP

In another related situation, if the existing excess steam cannot supply the full cooling demand, then the outside fuel option should be economically evaluated for generating extra steam for supplying the cooling demand. Such evaluation will result in

the extent of cooling load that should be taken care by absorptive refrigeration system. Again this option should be economically evaluated against the situation of only utilizing condensing turbine.

3.4.3 CASE 2

A second case of excess steam may occur when the plant is burning external fuel to produce steam. Releasing excess steam produced from external fuel is typically an indication of opportunities for saving. Such situation can occur, if the plant is poorly designed or also after some energy conservation study on existing plant. Usually the successful energy conservation study results in reduction of heating and cooling utility demand by the application of heat and mass integration techniques. There might be several options, the steam production can be reduced to match the reduced demand of the plant, but it will affect the cogeneration potential of the plant and result in additional power consumption. Also at load less then optimum design load the turbine will not work efficiently, further hampering the cogeneration potential of the plant. On the other hand, it might not be economical to burn excess fuel to produce excess steam to keep the cogeneration at a given level. Again, this excess steam could be targeted towards supplying the cooling demand through absorption refrigeration, which would result in demand of steam and also keep the cogeneration at a given level. Considering the marginal price of steam and the cogeneration efficiency of the plant should be utilized in performing any such evaluation of cooling load. The cogeneration efficiency is a function of fuel consumption, which is a function of mass and heat integration. The whole scenario is illustrated by a flow chart in figure 3.20**.**
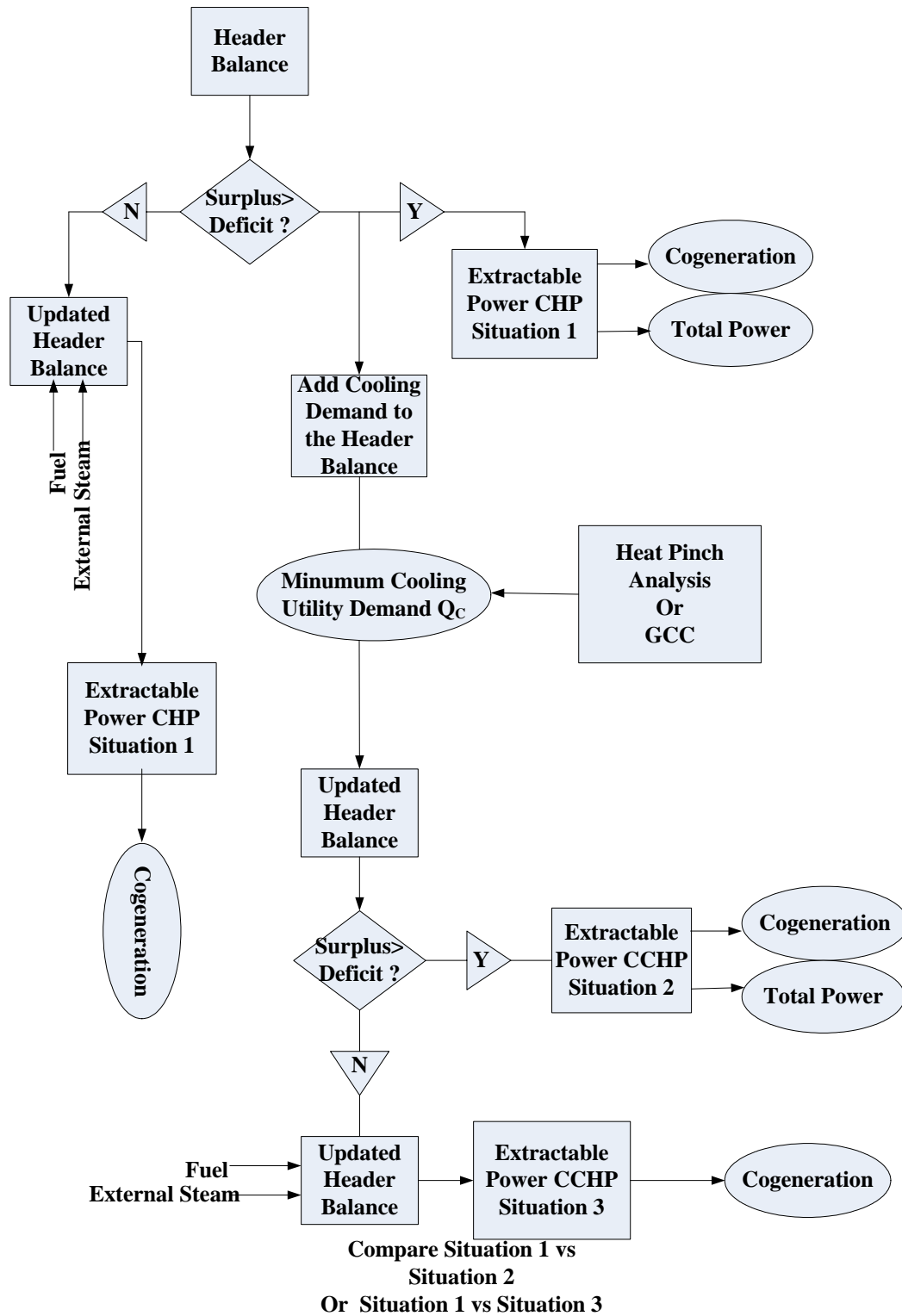
Figure 3.20: Different Cases for Power Cogeneration from a Given Steam Header

3.5 CASE STUDY

The graphical hierarchical approach introduced in the previous section will be utilized to solve an industrial case study to illustrate the applicability of the solution approach. For the case study a propylene manufacturing process by catalytic de-hydrogenation of propane is selected. The process for the production of propylene by catalytic dehydrogenation of propane is shown in figure 3.21. Fresh propane feed is mixed with recycle propane and heated to $1220^{o}F$, and fed to the dehydrogenation reactors. The reaction is endothermic, but the temperature is maintained at $1200\ ^{o}F$ by the heat stored in catalyst bed by both burning of coke and from the sensible heat of large quantities of regeneration air. The hot reactor gases are quenched successively in two spray towers and are compressed. The gases are then fed to the absorption column to recover most of the C_3 hydrocarbon by absorption with naphtha. The top of the absorber contains hydrogen, methane and C_2 hydrocarbons, which is purged as flue gas. The rich absorption oil is charged into stripping column, where the absorbed gases are separated from oil. The oil is returned to the absorber. The gases separated in the stripping column are sent to the de-ethanizer, where the ethane and lighter components are distilled off. The bottom product is charged to the depropanizer, where propylene-propane fraction and C_4^{+} fraction are separated. Propylene-propane mixture from the top of the depropanizer is further fed to the propylene columns to produce polymer grade propylene as the distillates. Unconverted propane recovered from the bottom of the propylene column is recycled to the dehydrogenation reactors after it is combined with fresh propane feed.
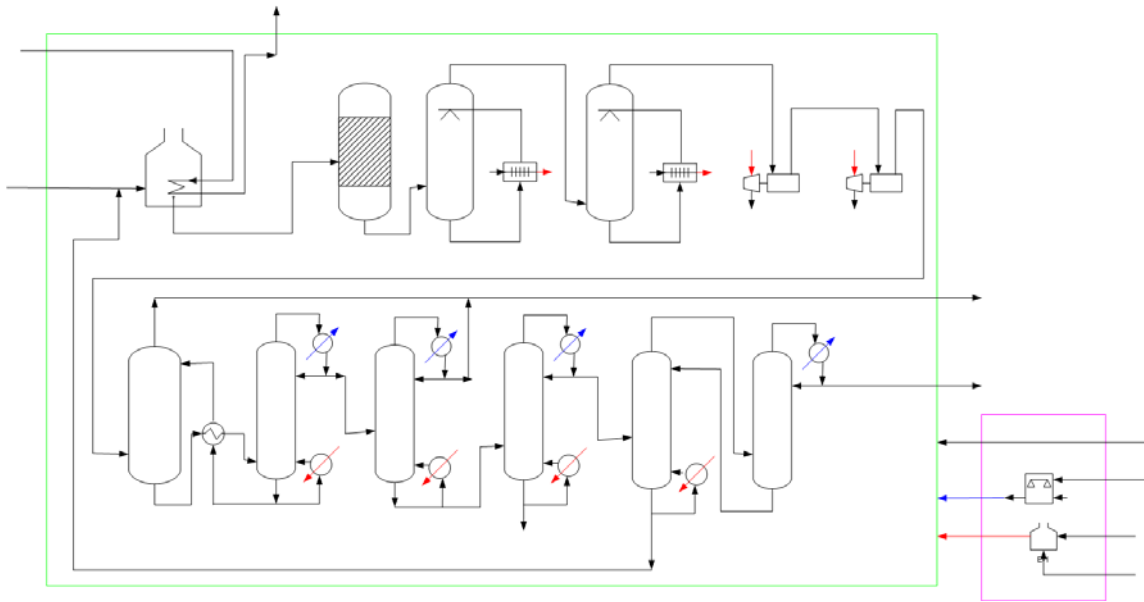
Figure 3.21: Propylene by Catalytic De-Hydrogenation of Propane

First the process is analyzed for collecting data required to initiate mass and heat integration analysis. Also data regarding the raw material and different utility consumption for various purposes are collected to determine the existing annual operating cost. The plant is assumed to run for 8,000 hours per year. It is also assumed that in the existing situation, the plant purchases all the required fuel and electricity from outside sources without considering the potential of utilizing process fuel for heating and cogeneration. The objective of the case study is to develop a revised process configuration optimizing fresh and utility consumption, waste recycle, recovery of material and thermal value from waste, external fuel and electricity consumption.

Tables 3.1-3.7 summarize the data for the heating, cooling, process sinks, process sources, and external resources for the case study.

Table 3.1: Process Heating/Cooling Data

Stream	T_{Supply} (^{o}F)	T_{Target} (^{o}F)
H1	170	130
H2	150	100
H3	190	100
H4	160	80
C1	90	170
C2	100	140
C3	130	190
C4	150	170

Table 3.2: Process Sink Data

Sink	Flowrate (lb/hr)	Maximum Inlet Mass Fraction (Impurities)	Maximum Inlet Load (kg/hr)
VA Process Reactor	34000	0.20	6800

Table 3.3: Process Source Data

Source	Flowrate (lb/hr)	Outlet Mass Fraction (Impurities)	Maximum Inlet Load (kg/hr)	Heating Values (Btu/lb)
De-ethanizer	10000	0.48	4800	1000
Absorption Column	20000	0.65	13000	1500

Table 3.4: Raw Material Data

Raw Material	Cost ($/lb)	Mass Fraction (Impurities)
Fresh	0.11	0.00

Table 3.5: Fresh Fuel Data

Fresh Fuel	Cost ($/MMBTU)	Heating Value (Btu/lb)
Fuel	2.6	13400

Table 3.6: Process Fuel Data

Process Fuel	Flow (lb/hr)	Heating Value (Btu/lb)
Depropanizer Bottom	20000	13400

Table3.7: Electricity Consumption Data

	Demand (MW)	Cost ($/kwhr)
Electricity	12	0.06

Other related data include the cost of waste treatment, which is given as $0.0022/lb of waste.

Determining heating utility demand from the process data is obtained through the grand composite curve analysis described earlier. As shown by figure 3.22, the minimum required heating utility is 182 MM Btu/hr of low-pressure steam. On the other hand, the minimum required cooling utility is determined to be 166 MM Btu/hr.

As mentioned earlier, the plant purchases all its required raw material and utilities and electricity from external sources, without considering recycle, reuse opportunity and also without considering recovery of thermal or material value from waste and producing electricity through cogeneration. Table 3.8 shows the calculation of existing operating cost for the plant.
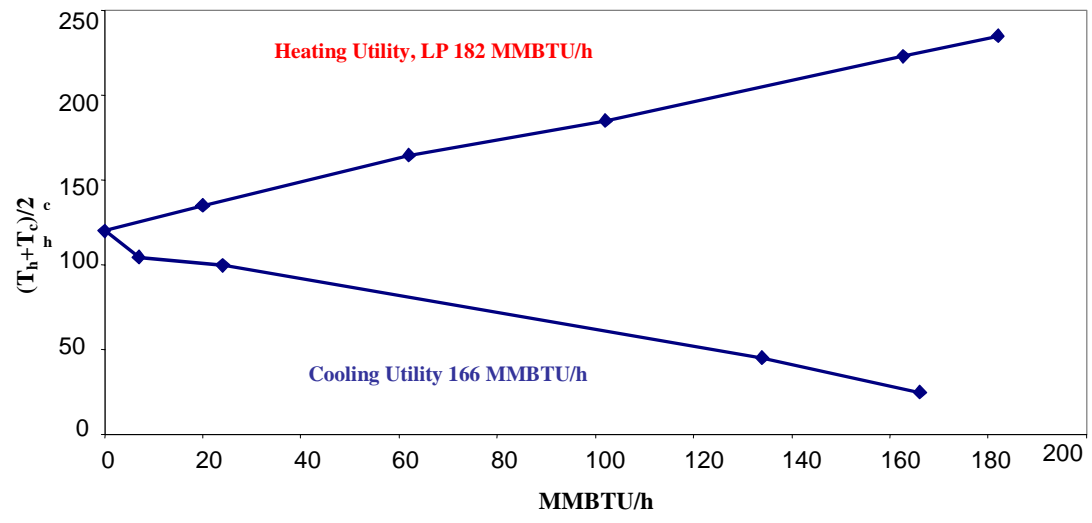
Fig 3.22: Grand Composite Curve (Case Study)

Table 3.8: Existing Operating Cost

Fuel Cost

$$\frac{182 \text{ MMBTU}}{\text{HR}} \times \frac{\$2.6}{\text{MMBTU}} \times \frac{8000 \text{ HR}}{\text{YR}} = \$3,785,600/\text{YR}$$

Raw Material Cost

$$\frac{34,000 \text{ LB}}{\text{HR}} \times \frac{\$0.11}{\text{LB}} \times \frac{8000 \text{ HR}}{\text{YR}} = \$29,992,000/\text{YR}$$

Electricity Cost

$$\frac{12 \text{ MW}}{} \times \frac{\$0.06}{\text{KWHR}} \times \frac{1000 \text{ KW}}{1 \text{ MW}} \times \frac{8000 \text{ HR}}{\text{YR}} = \$5,760,000/\text{YR}$$

Waste Treatment Cost

$$\frac{50,000 \text{ LB}}{\text{HR}} \times \frac{\$0.0022}{\text{LB}} \times \frac{8000 \text{ HR}}{\text{YR}} = \$880,000/\text{YR}$$

So, total annual operating cost for the existing situation is:

$3,785,600 + $29,992,000 + $5,760,000 + $880,000 = $40,417, 600

Figure 3.23 shows the existing situation and the existing operating cost. This is showing that all the process sources are directly going to the process waste treatment, and the plant is purchasing all its fresh raw material, fuel and power from external

sources. Now the proposed graphical methodology will be utilized to reduce the annual operating cost of the plant.
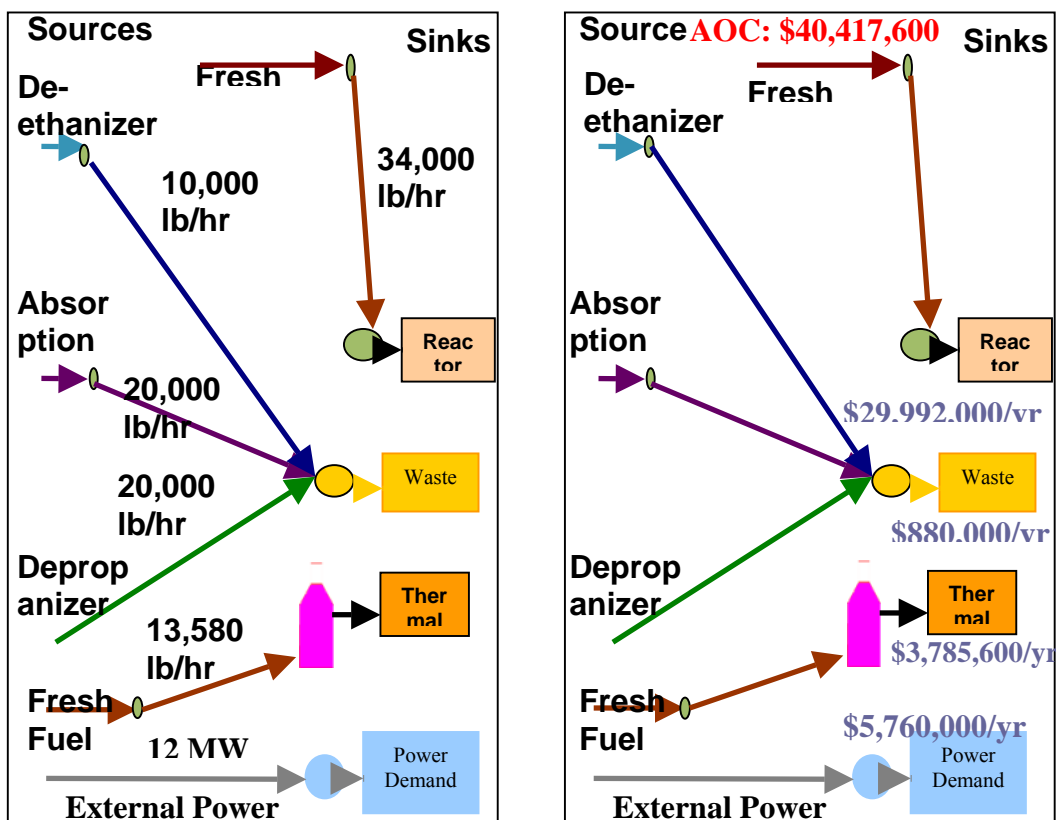
Fig 3.23: Existing Scenario Flowrate and Cost

3.5.1 MASS AND HEAT INTEGRATION

According to the methodology developed in previous section, first the mass and heat integration analysis is performed. From the source-sink data, the availability of direct recycle opportunity is determined. A material recovery pinch analysis is carried

out as shown by figure 3.24. The pinch diagram results in minimum fresh requirement of 21,000 kg/hr. Also, the diagram shows 17,000 kg/hr of waste discharge, which mainly contains recoverable organic materials.
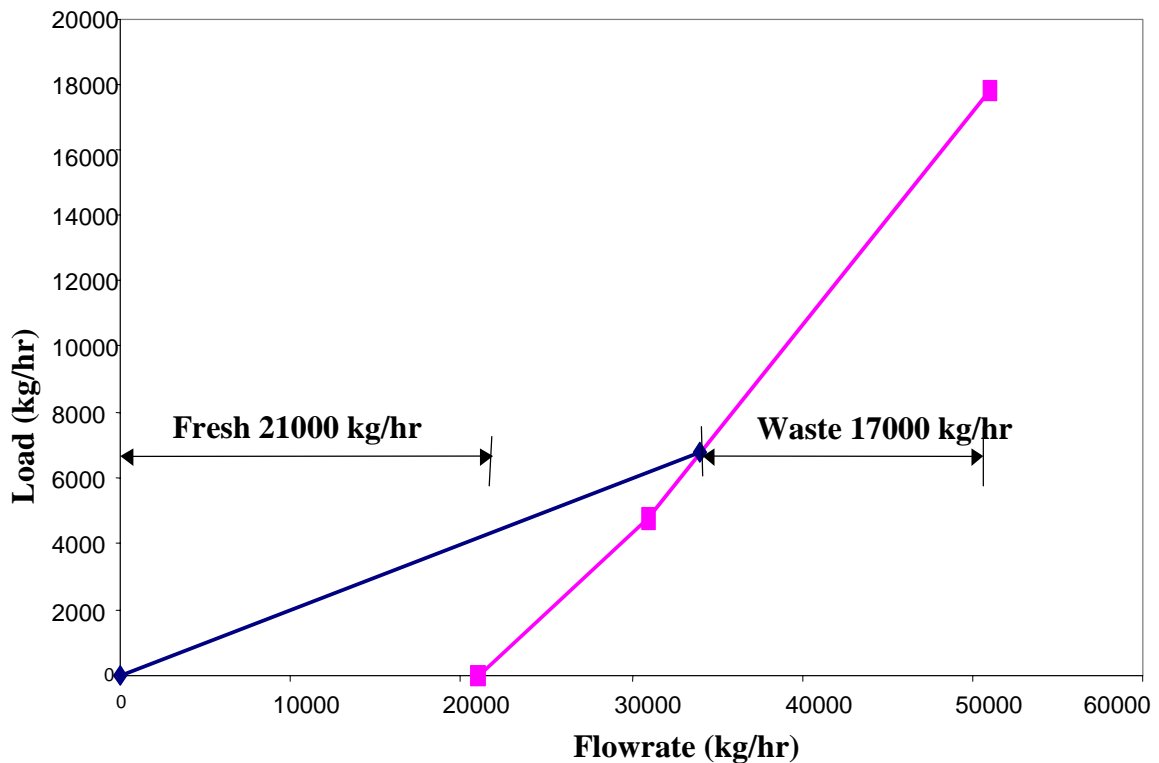
Figure 3.24: Material Recycle Pinch Diagram (Case Study)

This waste stream can be targeted for interception for material value recovery, or targeted for thermal value recovery. Both the material and thermal value recovery should be economically feasible for the given situation. The unrecoverable waste is discharge or sent for waste treatment as shown by figures 3.25 and 3.26.

Recovered Material Value

Waste → Interception Network

Impurities

WASTE 17000 kg/hr

WHB/HRS

Recovered Thermal Value

Waste

Fuel

Ash

Waste Treatment & Discharge
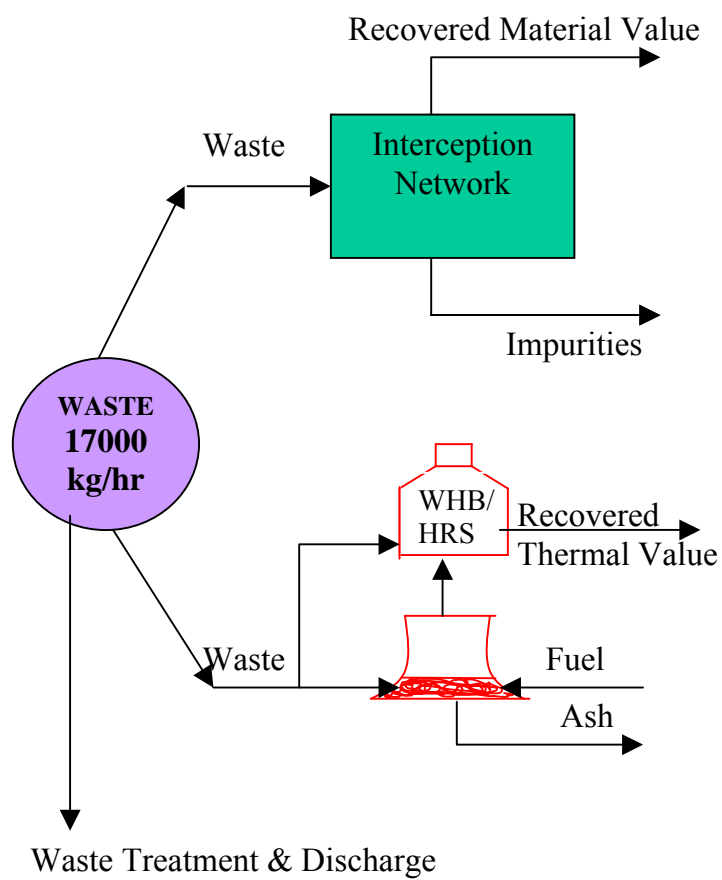
Figure 3.25: Alternatives for the Process Waste

Recovered Organic

Waste stream → Interception Network

Impurities
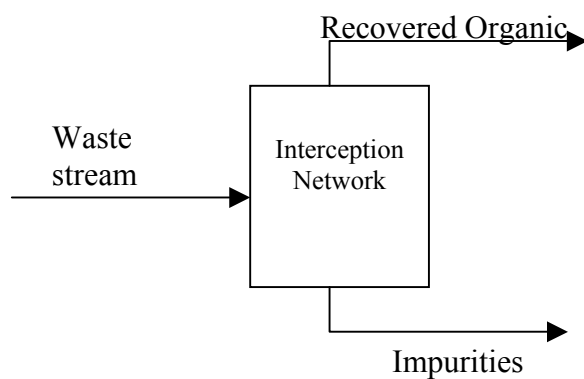
Figure 3.26: Mass Integration for Recovery of Organic Materials

First the interception network feasibility is studied. Correct technology is identified and the cost of recovery is determined. It is found that the waste stream is mainly from the absorption column bottom with the impurity mass fraction of 0.65. Further analysis revealed that this stream is not suitable for economical recovery of material value. So the whole stream is targeted either for waste treatment or thermal value recovery.

Boiler operational data revealed that this waste stream could be utilized for substituting external fuel. So total recovered thermal value from absorption column bottom is:

$$\frac{17000 \text{ lb}}{\text{hr}} \left| \frac{1500 \text{ BTU}}{\text{lb}} \right| \frac{1 \text{ MMBTU}}{10^6 \text{ BTU}} = 25.5 \text{ MMBTU/hr}$$

Also from process fuel data we get another stream, i.e. depropanizer bottom stream that can be burned into the boiler for thermal value recovery and aid in external fuel substitution. Total recovered thermal value from depropanizer bottom is:

$$\frac{20000 \text{ lb}}{\text{hr}} \left| \frac{7000 \text{ BTU}}{\text{lb}} \right| \frac{1 \text{ MMBTU}}{10^6 \text{ BTU}} = 140 \text{ MMBTU/hr}$$

After, fuel substitution, the external fuel demand comes down to:

182 MMBTU/hr – 165.5 MMBTU/hr = 16.5 MMBTU/hr.

So, the cost of external fuel purchase becomes:

$$\frac{16.5 \text{MMBTU}}{\text{HR}} \; \left| \; \frac{\$2.6}{\text{MMBTU}} \; \right| \; \frac{8000 \text{ HR}}{\text{YR}} \quad = \quad \$343,200/\text{YR}$$

Figure 3.27 shows the result of mass and heat integration analysis and the resulting process modifications and the annual operating cost. Figure 3.28 shows the difference of operating cost between the existing situation and the proposed modification. Here we see that the annual operating cost is reduced for around $ 35 million to around $19 million due to direct recycle and utilization of thermal value recovery. As a result of these actions the raw material cost is reduced, the waste treatment cost is eliminated and also the fresh fuel cost is reduced.

3.5.2 STEAM HEADER BALANCE

Further process data analysis reveals that more steam can be generated from the process utilizing waste heat boiler and heat recovery steam generators. The following are the steam production data from process utilizing WHB and HRSG.

HP: 90.5 MMBTU/hr (by heat recovery from hot exhaust)

MP: 10 MMBTU/hr (HRSG)

LP: 20 MMBTU/hr (HRSG)

So, the process has excess steam and external fuel consumption for steam production is totally eliminated. Also the process has some non-heating steam demand for the stripping column operation and steam driven equipments. Due to these additional availabilities of steam from the process it eliminates the external fuel consumption for

the plant. This reduces the annual operating cost of the plant further. So, it is seen that a very systematic analysis of process data and process situation is required, and can result in annual operating cost reduction. The updated annual operating cost is around $18.5 million. Figure 3.29 shows the updated annual operating cost.
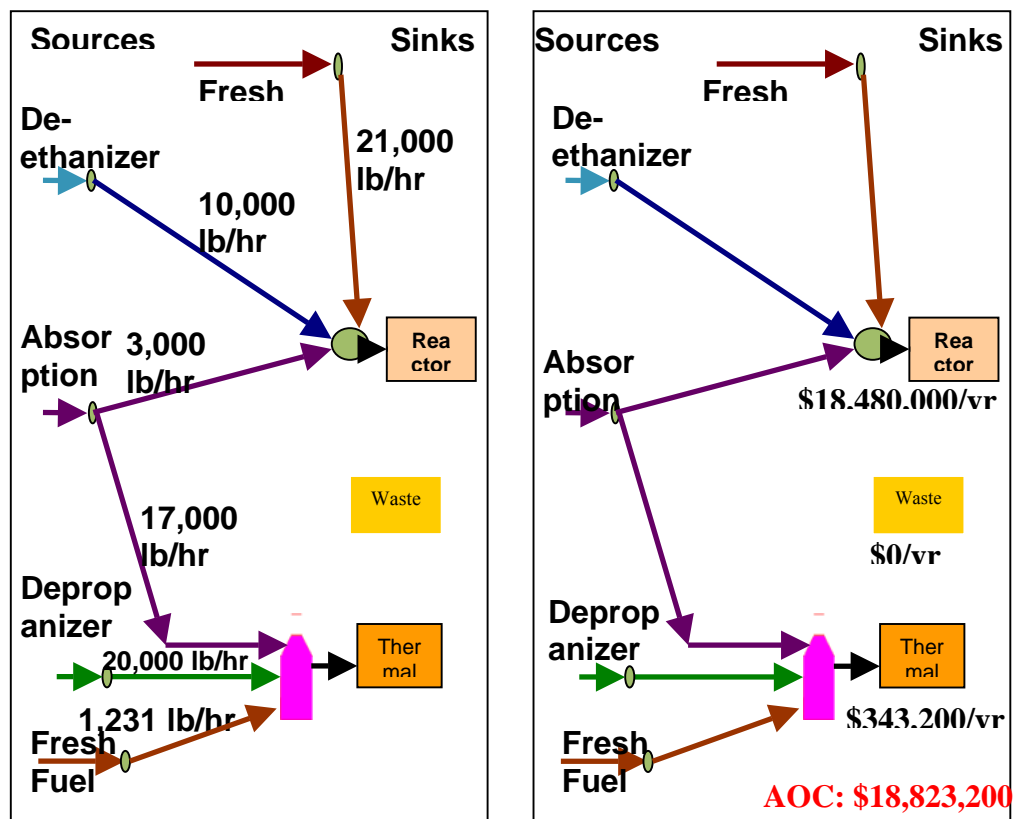
Figure 3.27: Mass and Heat Integration (Graphical Approach)

Now putting all these to generate the steam header balance for the plant. Table 3.9 shows the steam header balance of the plant.

Table 3.9: Steam Header Balance (Graphical Technique)

Steam	Pressure (psia)	Temperature (^{o}F)	Supply (MMBTU/h)	Demand (MMBtu/h)
HP	600	800	256	0
MP	130	350	10	15
LP	40	270	20	192
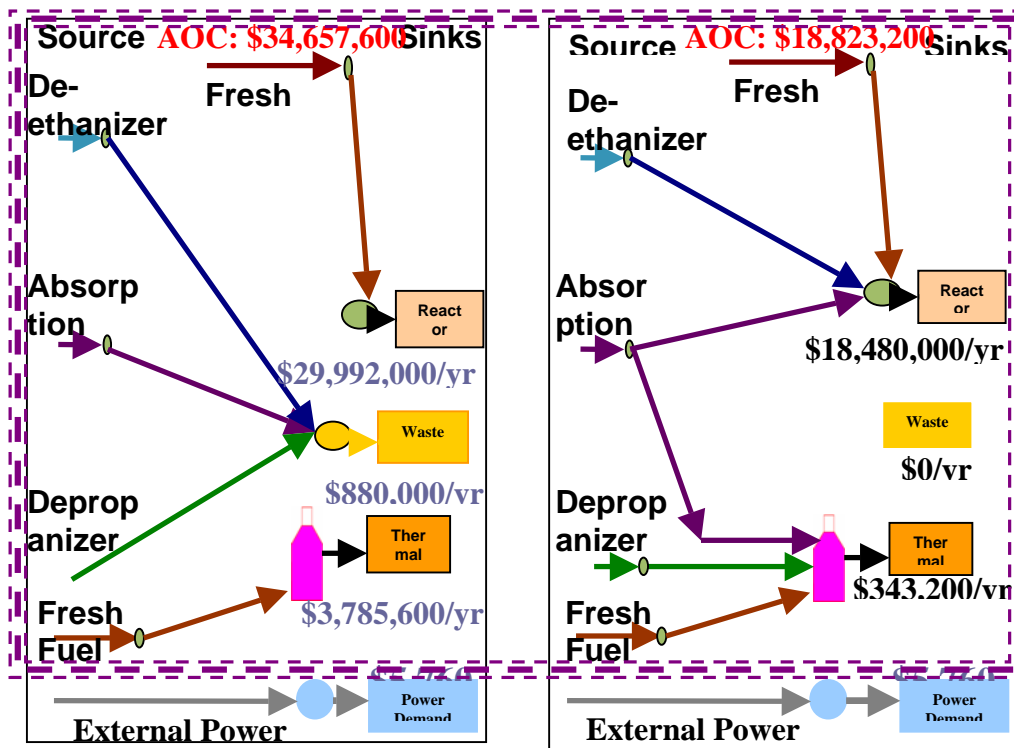
Figure 3.28: Comparison of Graphical Mass and Heat Integration Results (right) with the Existing Situation (left)
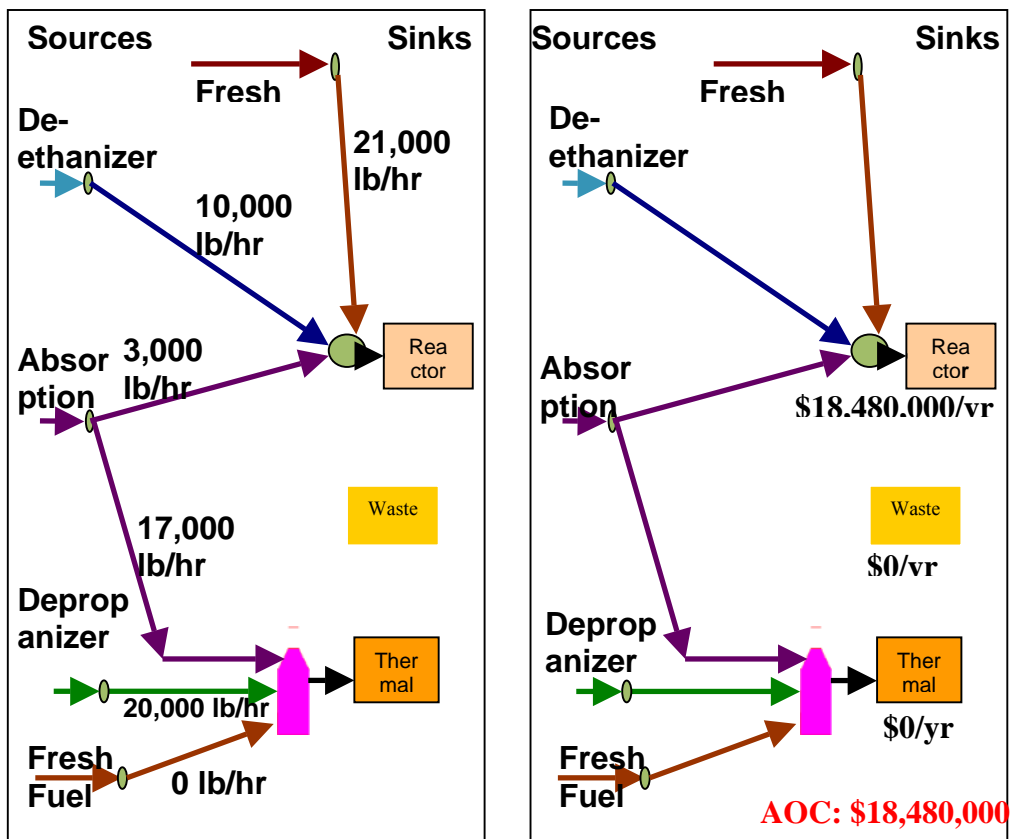
Figure 3.29: Elimination of External Fuel Cost

3.5.3 EXTRACTABLE WORK METHOD

From the header balance, the extractable work method is utilized to determine the cogeneration potential (figure 3.30). We find that, the given header balance has existing 26.3 MMBTU/hr cogeneration potential with an excess steam of 30320 lb/hr at HP header level. Since, the steam header has excess steam, it is evaluated whether the cooling load for the plant can be supplied using an absorption chiller. Absorption chiller utilizes steam to supply cooling demand.
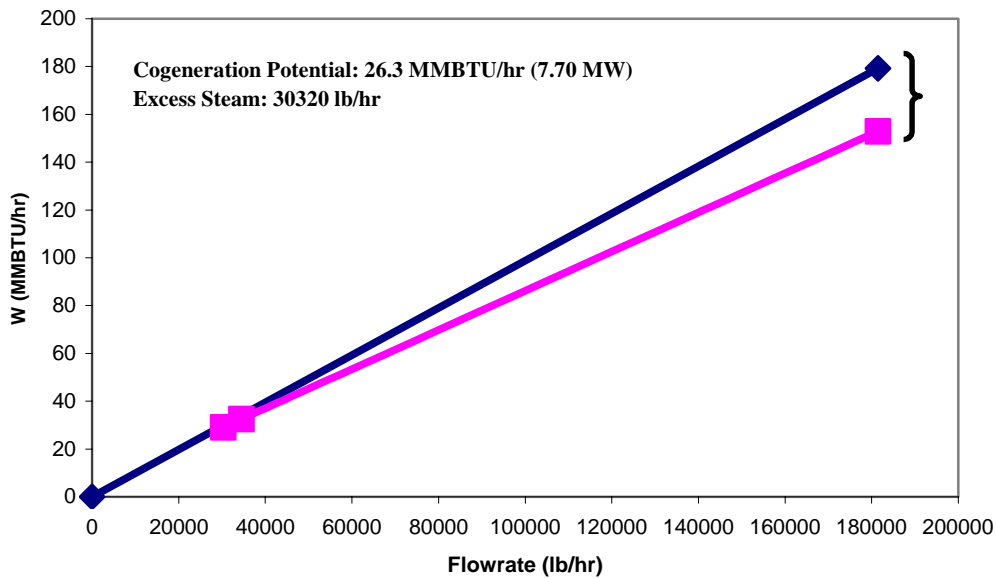
Figure 3.30: Extractable Work Method for CHP (Case Study)

Cooling requirement for the plant (as seen from GCC) is 166 MMBtu/hr. From the excess steam available within the process 43 MMBtu/hr of cooling load can be supplied. Adding this cooling load to the header balance, we get an updated header balance. Table 3.10 shows updated header balance.

Table 3.10: Updated Steam Header Balance (Graphical Technique)

Steam	Pressure (psia)	Temperature (°F)	Supply (MMBTU/h)	Demand (MMBtu/h)
HP	600	800	256	0
MP	130	350	10	51
LP	40	270	20	192

Now we need to determine the new cogeneration potential with the updated header balance. Cogeneration potential can be generated through using the extractable work method (figure 3.31).
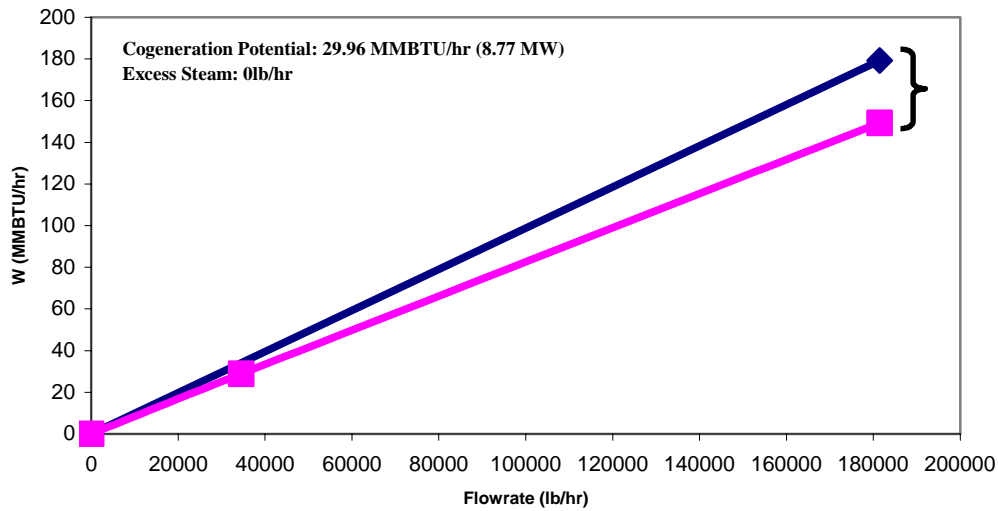
Figure 3.31: Extractable Work Method for CCHP

So the cogeneration potential is enhanced and excess steam is reduced, by considering the cooling load in the steam header balance. New cogeneration potential is 8.77 MW compared to 7.70 MW while we only considered the heating demand. Also we see that the excess steam becomes 0 lb/hr compared to 30320 lb/hr for the case only with heating demand. Now economic evaluation should be performed to determine feasible technology to supply the rest of the cooling load. While evaluating the cooling load supply technology, emphasis should be provided on local cooling demand and also any global power demand.

Now table 3.11 shows the annual operating cost for the reconfigured process
flow diagram identified by the hierarchical graphical technique.

Table 3.11: Updated Operating Cost (Graphical Technique)

Fuel Cost

$$\frac{0 \text{ MMBTU}}{\text{HR}} \quad \frac{\$2.6}{\text{MMBTU}} \quad \frac{8000 \text{ HR}}{\text{YR}} = \$0/\text{YR}$$

Raw Material Cost

$$\frac{21{,}000 \text{ LB}}{\text{HR}} \quad \frac{\$0.11}{\text{LB}} \quad \frac{8000 \text{ HR}}{\text{YR}} = \$18{,}480{,}000/\text{YR}$$

Electricity Cost

$$\frac{3.23 \text{ MW}}{} \quad \frac{\$0.06}{\text{KWHR}} \quad \frac{1000 \text{ KW}}{1 \text{ MW}} \quad \frac{8000 \text{ HR}}{\text{YR}} = \$1{,}550{,}400/\text{YR}$$

Waste Treatment Cost

$$\frac{0 \text{ LB}}{\text{HR}} \quad \frac{\$0.0022}{\text{LB}} \quad \frac{8000 \text{ HR}}{\text{YR}} = \$0/\text{YR}$$

So, total annual operating cost for the existing situation is:

$0 + \$18,480,000 + \$1,550,400 + \$0 = \$20,030,400$

Figure 3.32 shows the current flowrates and the updated operating cost. Figure
3.33 shows the reconfigured process flow diagram.
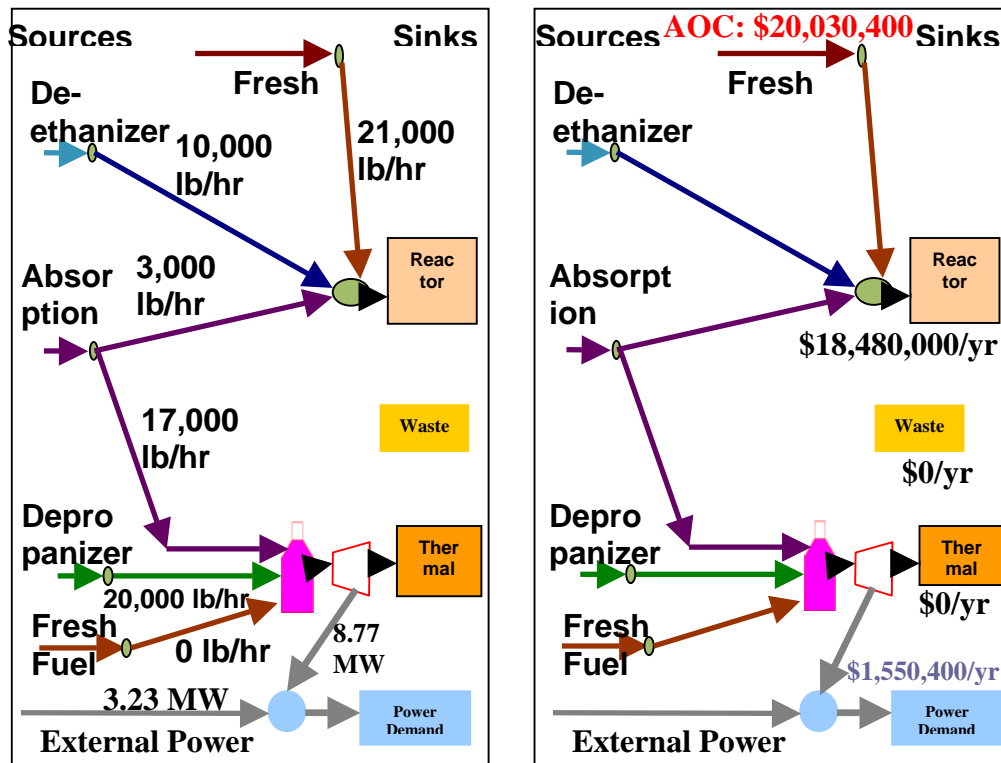
Figure 3.32: Hierarchical Graphical Approach Case Study Results
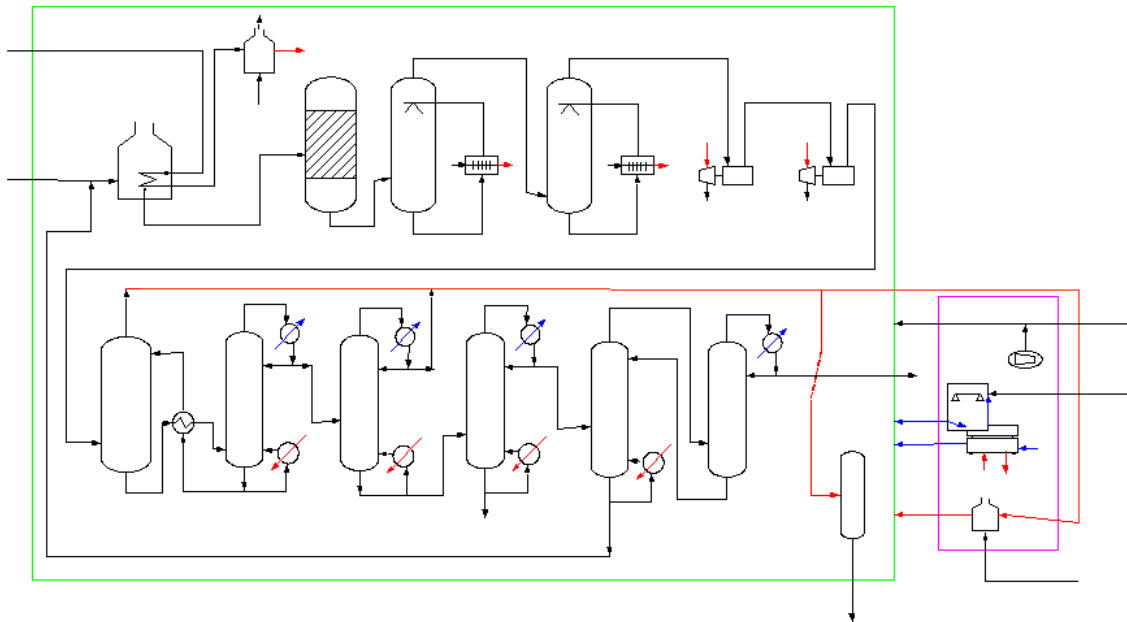
Figure 3.33: Reconfigured Process Flow Diagram (Graphical Approach)

In the reconfigured diagram it is shown that all the waste stream has been utilized either for material value or thermal value recovery. Part of the absorption column and de-ethanizer top product is utilized for material value recovery by recovering ethylene for the adjacent VAC process, and the rest is utilized as fuel in the boiler. Also depropanizer bottom product is utilized as fuel in the boiler. So the process does not consume any external fuel. Also power is cogenerated within the system reducing external power consumption.

3.5.4 ECONOMIC COMPARISON

Figure 3.34 illustrates the comparison of annual operating cost between the existing situation and the reconfigured process identified by graphical hierarchical approach.
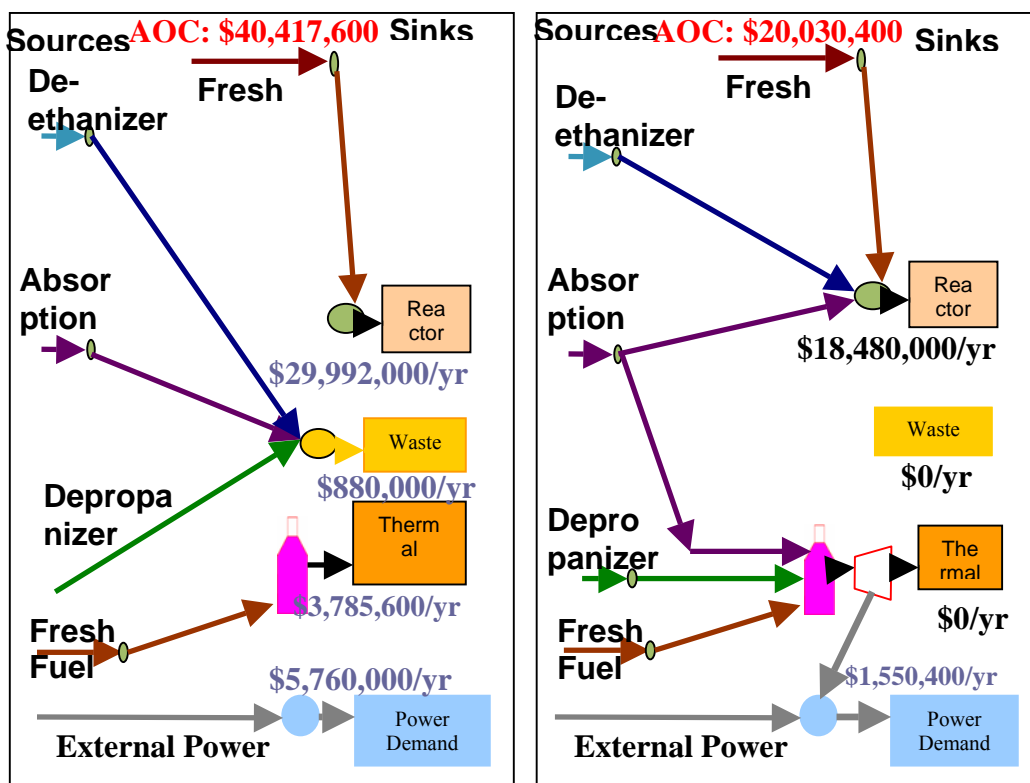
Figure 3.34: Comparison between Existing Situation and Graphical Approach Solution

The step-by-step hierarchical methodology results in elimination of external fuel consumption and elimination of waste treatment. All the process wastes have been utilized either for material value or thermal value recovery. Also the external electricity consumption is reduced resulting in reduction in electricity purchase cost. Also due to direct recycle the fresh raw material consumption is also reduced. As a result of the devised solution, the target for cumulative reduction in operating cost is approximately $20 million/yr. We see that the cost reduction is about 50% of the existing situation, thus demonstrating the applicability of the procedure.

3.6 CONCLUSIONS

This chapter has presented a systematic and hierarchical procedure for the simultaneous optimization of heating, cooling, power cogeneration, fuel consumption, and waste recovery. Visualization tools for mass and energy integration techniques have been utilized independently and as well as interdependently to generate and screen alternatives. The strong interaction between the material utilities and energy has been demonstrated.

The mass integration techniques result in optimal recycle and minimization of fresh consumption and waste disposal. The waste identified by the mass integration is subjected to economic evaluation between material and thermal value recovery. This thermal value directly affects the steam header balance and as well as fuel consumption and cogeneration efficiency.

The heat integration provided the information about the minimum cooling and heating demand. Furthermore, it provided the information regarding the state of the different utilities required. Based on the data analysis and heat integration, the existence of exhaust heat can be determined and used to optimize fuel consumption, steam header balance, and power cogeneration. The steam header balances was created from the insights derived from mass integration, heat integration, and also from the given process situation. Both mass and heat integration analyses were used to provide guidelines and insight for the fuel requirements for steam generation.

It was found that cogeneration is dependent on steam headers and the fuel consumption. Therefore, the power consumption for the plant may be affected by

material utilities and energy requirement. Additionally, the cogeneration efficiency is affected by the availability of combustible waste and waste heat for fuel substitution. The developed procedure quantifies the relationship between material utilities, energy requirements, and waste utilization for material and thermal objectives. The hierarchical nature of the developed procedure enables the decomposition of the various mass and energy integration activities into tractable tasks. The procedure results in the optimal allocation of combustible materials, the optimization of integrated heat exchange, the extent of power cogeneration, and the optimal alternatives for utility generation. The proposed procedure should be very helpful in guiding engineers as that approach the complex task of optimizing material and energy utilities, energy management, and waste allocation.

CHAPTER IV

MATHEMATICAL PROGRAMMING

4.1 OVERALL APPROACH

The previous chapter has focused on graphical tools for mass and energy integration. While these visualization tools provide valuable insights to the designer, they are not amenable for simultaneous solution. Each graph is plotted as a "snapshot" for a given set of variables. Changing the values of these variables implies the iterations over possible values of these variables. As the number of variables and the number of possible values of the variables increase, it becomes useful to develop mathematical tools which can address parts of the problem simultaneously. The purpose of this chapter is to develop a novel approach to optimizing heating, cooling, cogeneration, and waste allocation using a mathematical approach. The problem is decomposed into consecutive sub-sections where each section will be solved for global optimality and the solutions are brought together to generate optimal process.

Figure 4.1 summarizes the proposed framework for the mathematical procedure. In brief the proposed procedure is as follows- from the process data first simultaneous mass and heat integration analysis is performed. From the result of simultaneous mass and heat integration analysis, the steam header balance is generated. Then the steam header balance is utilized for determining the cogeneration potential and also determining the availability of any excess steam. The excess steam is reconciled

according to the process cooling and power requirement. The proposed approach will be discussed in detail in the following sections.
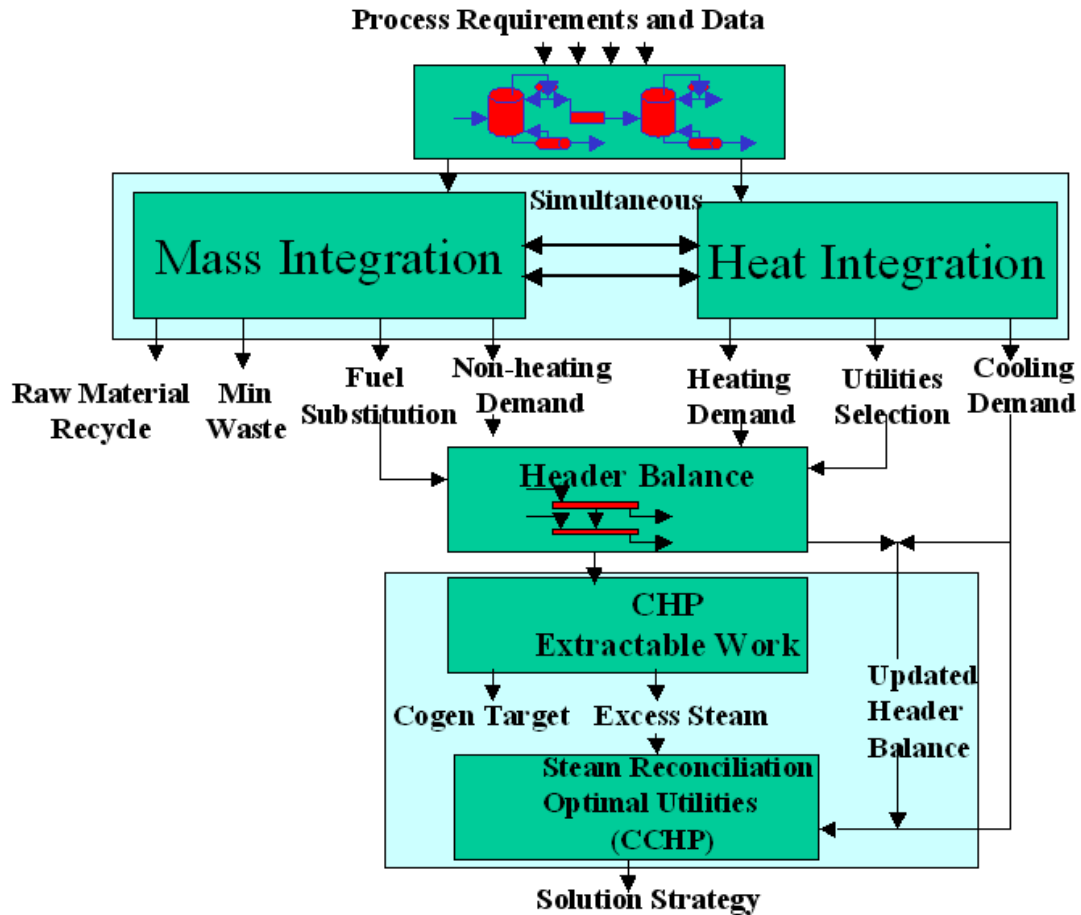
Figure 4.1: Overall Approach (Mathematical Programming)

4.1.1 OVERALL OBJECTIVE FUNCTION

Minimize: annual operating cost =

$$C_{Fresh} \bullet \sum_{j=1}^{N_{Sinks}} Fresh_J + C_{Fuel} \bullet Fuel + C_{Waste} \bullet Waste + \sum_{k=1}^{N_{int}} Interception\_Cost_k \bullet E_k + C_{Power} \bullet Power \qquad (4.1)$$

Where, C_{Fresh} is the cost of the fresh resource (\$/amount of resource), $Fresh_j$ is the amount of fresh resource fed to the jth sink (mass per year). $Interception\_Cost_k$ is the total annualized fixed equipment cost associated with interception device k. This cost is typically a nonconvex function of flowrate, inlet and outlet compositions, and design as well as operating parameters. E_k is a binary integer variable that has the value of 1 or 0 depending on whether or not unit k is used or not, respectively. C_{fuel} is the cost of fresh fuel (\$/MMBTU) fed into the boiler, $Fuel_q$ is the amount/flowrate (MMBTU/h) of fuel fed into the boiler. C_{waste} is the annual waste treatment cost and *waste* is the total amount of flow going to waste (lbs/yr). C_{power} is the annual external power (electricity) power cost (\$/kwhr) and *Power* is the amount (MW) of power purchased per year from outside sources.

4.1.2 COST FUNCTION IN THE OBJECTIVE EQUATION

The cost functions in the objective equation require some special considerations. The cost functions such as: C_{fresh}, C_{fuel}, C_{power}, C_{waste} are mainly functions of their respective flowrates. So, in general it can be defined as:

$$Cost = f(consumption) \qquad (4.2)$$

Although, cost is proportional to the flowrates and function of flowrates, but in many cases it is not a continuous function of the flowrates. Cost may change abruptly over certain decision variables, making it a discontinuous function. Also cost function can be a constant number or variable depending on the consumption capacity. The variations of cost functions are shown in figure 4.2. Modeling these cost functions

requires special considerations. Binary integer variables can be used to model these cost functions.
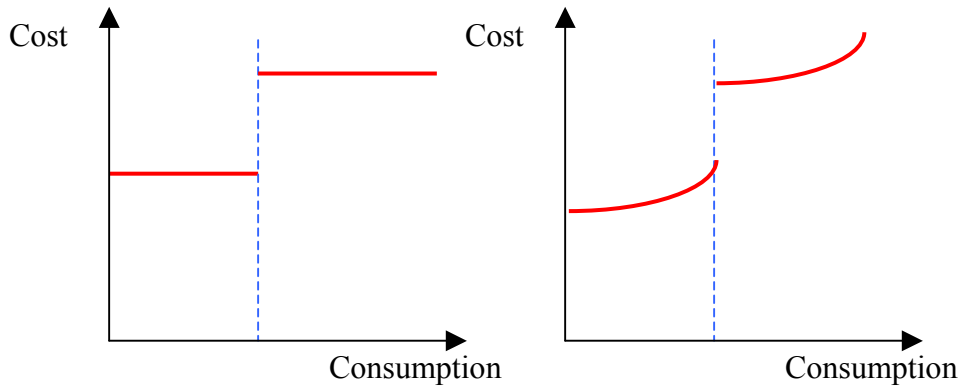
Figure 4.2: Cost Modeling

Cost function may be transformed into a mixed integer formulation using a binary integer variable (I). The following example illustrates such formulation. For a specific input (fresh, fuel, power) x to any manufacturing process with flowrate $Flow_x$, lets define a cost function, $Cost_x$. Also we define that the cost function $Cost_x$ is equal to a constant value $Cost\_A$ for the flowrate of the input $Cons_x$ less then certain flowrate $Cons_{switch}$, and the cost function $Cost_x$ is equal to another constant value $Cost\_B$ for the flowrate of the input $Cons_x$ greater then the switching flowrate $Cons_{switch}$. The situation is depicted with the following equations:

$$Cost_x = Cost\_A \quad \text{for } Cons_x \leq Cons_{switch} \qquad (4.3)$$

$$Cost_x = Cost\_B \ \ for \ \ Cons_x > Cons_{switch} \tag{4.4}$$

This situation can be handled with integer variable, $I = 0$ or 1 (*Binary Integer*)

$$Cost_x = (Cost\_A \bullet I + Cost\_B \bullet (1 - I)) \tag{4.5}$$

Here, When I = 1,

$$Cost_x = (Cost\_A \bullet I + Cost\_B \bullet (1 - I)) = Cost\_A \tag{4.6}$$

And, when I = 0

$$Cost_x = (Cost\_A \bullet I + Cost\_B \bullet (1 - I)) = Cost\_B \tag{4.7}$$

Now to model the conditions that assign the values of I to be 0(zero) or 1(one), we define another constraint:

$$(Cons_{switch} - Cons_x) \bullet (2 \bullet I - 1) \geq 0 \tag{4.8}$$

If, $Cons_x \leq Cons_{switch}$, I is forced to be 1(one), otherwise if it is zero the value of the function on the right hand side becomes negative, which is infeasible. On the other hand, if $Cons_x \geq Cons_{switch}$, the term $(Cons_{switch} - Cons_x)$ becomes negative and I is forced to be zero so that $(2 \bullet I - 1)$ becomes negative and the function on the right hand side becomes positive satisfying non-negative inequality constraint. For the mathematical programming solution of the current problem the cost function is assumed to be a continuous function of flowrate.

The overall formulation will be decomposed into multiple sections that can be integrated. The formulation for each section is given below. Next, the whole approach will be illustrated through a case study. Finally, the results of the solutions obtained

through the graphical approach and the mathematical programming formulation will be compared.

4.2 SIMULTANEOUS MASS AND HEAT INTEGRATION

The problem statement for this section becomes:

Given a process with:

- A set of process sources: SOURCES = $\{i \mid i = 1,2, \ldots, N_{sources}\}$ which can be recycled/reused in process sinks. The process sources can also be burned in boiler to recover thermal value. Each sources has a given flowrate, W_i, a given composition, y_i^{in}, and a given heating value H_i.

- A set of process sinks (units): SINKS = $\{j \mid j = 1,2, \ldots, N_{sinks}\}$. Each sinks requires a given flowrate, W_j, and a given composition, z_j^{in}, that satisfies the following constraint:

$$z_j^{min} \le z_j^{in} \le z_j^{max} \qquad \forall j \in \{1 \ldots N_{sinks}\} \qquad (4.9)$$

where z_j^{min} and z_j^{max} are given lower and upper bounds on acceptable compositions to unit j.

- A set of interception units: INTERCEPTORS = $\{k \mid k = 1,2, \ldots, N_{Int}\}$ that can be used to remove the targeted species from the sources.

- A process boiler q, which supplies the heating demand for the process. The boiler can be used to recover thermal value from the sources.

Available for service is a fresh (external) raw material that can be purchased to supplement the use of process sources in the process sinks. Also available fresh

(external) fuel which can be purchased and utilize in boiler in addition to any process source to supply the heating demand.

The objective is to develop an optimization method to determine the following:

- Minimum cost of the fresh (external) raw material and interception units that satisfy the process requirements

- Minimum cost of fresh (external) fuel to satisfy the process demand.

- Minimum cost of waste treatment

- Optimum allocation of sources to sinks

- Optimum selection of interception devices

- Optimum duties of source interception.

Gabriel and El-Halwagi (2005) solved the problem of simultaneous synthesis of waste interception and material reuse network by problem reformulation for global optimization. An analogous approach will be utilized in this section for approaching a global optimum solution for mass and heat integration. This work introduces an additional process sink to the problem. This sink is in the form of the process boiler which is utilized to recover thermal value from the process sources. The introduction of the boiler as a process sink provides opportunity for simultaneous comparison between the material value recovery and thermal value recovery from the process sources.

For the problem reformulation, the problem will be reformulated as a supply-demand (source-sink) optimization problem. Similar discretized data table for the interception network devices, as used by Gabriel and El-Halwagi (2005) will be utilized to eliminate the non-linearity in the formulation, and merge the performance of the

interceptor to the interceptor cost function in the objective equation. The assumptions defined by Gabriel and El-Halwagi (2005) also hold true for this current formulation.

4.2.1 PROBLEM PRESENTATION

The problem to be solved can be described through the figure 4.3. This is analogous to the work previously done by Gabriel and El-Halwagi (2005):
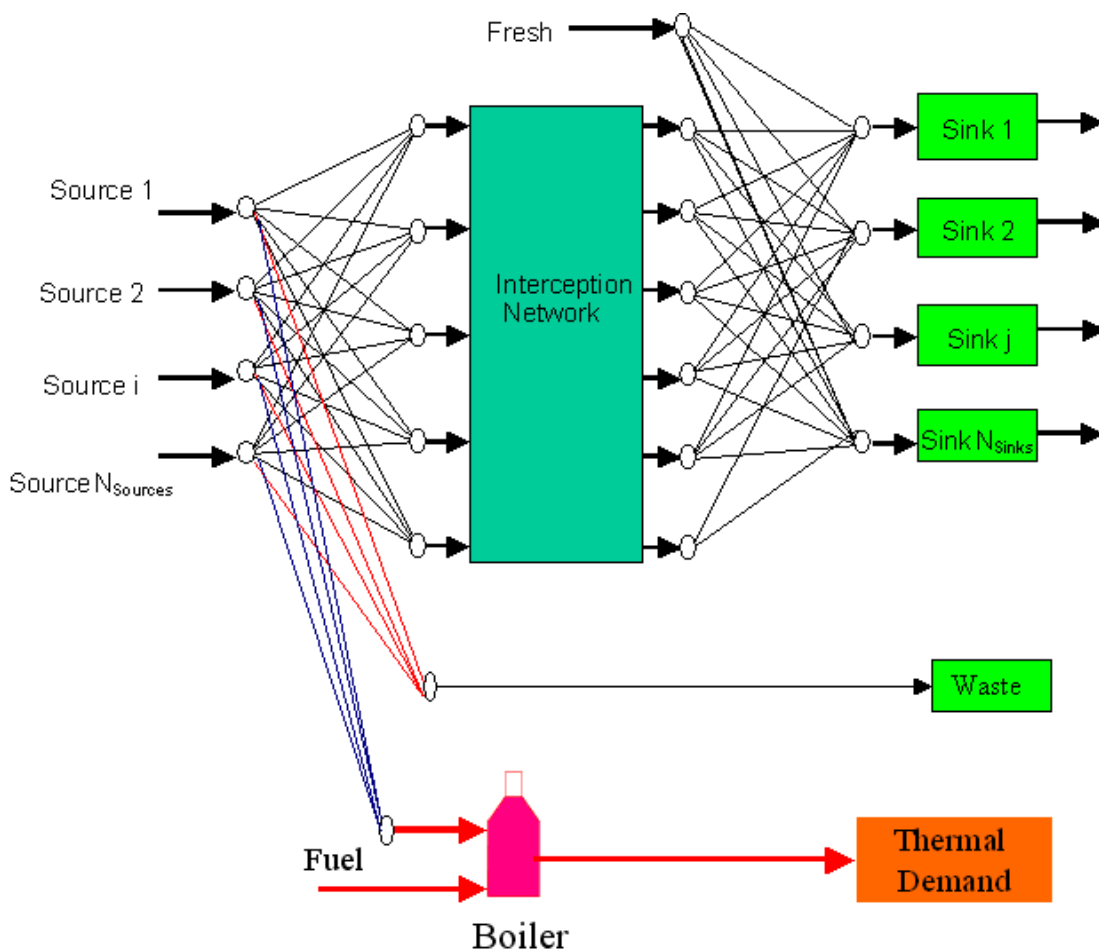
Figure 4.3: Simultaneous Mass and Heat Integration (Mathematical Programming)

The sources can be fed into the interception network for material value recovery, and the process boilers for thermal value recovery. Unused sources are fed into the waste stream for waste treatment facility. The sources are segregated into unknown flowrates (to be determined by optimization formulation) and fed into all these sinks. The sources that are fed into interception network get their impurity composition altered according to the process requirement if an interception device is used. A process stream may also pass through an interception network unchanged, indicating that no interception was utilized. This case is equivalent to a process source that is directly recycled to the sink. The sources leaving the interception devices are allowed to mix. The source streams that are fed to the boiler are assumed to undergo complete combustion. This recovers thermal value from the source and creates fuel substitution opportunities. The source streams that are not fed to interception networks or boiler (i.e., the unused source streams) are fed to the waste treatment facility for discharge.

4.2.2 MATHEMATICAL FORMULATION

The objective is to minimize the cost of the fresh resource, interception devices, and waste treatment. Hence, the objective functions can be expressed as:

Minimize total annualized cost =

$$C_{Fresh} \bullet \sum_{j=1}^{N_{sinks}} Fresh_j + \sum_{k=1}^{N_{int}} Interception\_Cost_k \bullet E_k + C_{waste} \bullet waste + C_{fuel} \bullet Fuel_q \qquad (4.10)$$

where C_{Fresh} is the cost of the fresh resource ($/amount of resource), $Fresh_j$ is the amount of fresh resource fed to the jth sink (mass per year). $Interception\_Cost_k$ is the total annualized fixed equipment cost associated with interception device k. This cost is

typically a nonconvex function of flowrate, inlet and outlet compositions, and design as well as operating parameters. E_k is a binary integer variable that has the value of 1 or 0 depending on whether or not unit k is used or not, respectively. C_{fuel} is the cost of fresh fuel (\$/MMBTU) fed into the boiler, $Fuel_q$ is the amount/flowrate (MMBTU/h) of fuel fed into the boiler. C_{waste} is the annual waste treatment cost and *waste* is the total amount of flow going to waste (tons/yr).

Subject to the following constraints: Splitting of the sources to all the interception devices, boiler and waste treatment facility:

$$F_i = \sum_{k=1}^{N_{int}} W_{i,k} + W_{i,z} + W_{i,waste} \qquad \forall\, i \in \{1...N_{Sources}\} \qquad (4.11)$$

where F_i is the flowrate of the ith source.

Mixing of sources before the interception devices:

$$W_k = \sum_{i=1}^{N_{sources}} w_{i,k} \qquad \forall\, k \in \{1...N_{int}\} \qquad (4.12)$$

Mixing of sources before bolier:

$$W_z = \sum_{i=1}^{N_{sources}} w_{i,z} \qquad (4.13)$$

Component material balance for the mixing before interception:

$$W_k \cdot Y_k^{in} = \sum_{i=1}^{N_{sources}} w_{i,k} \cdot y_i^{in} \qquad \forall\, k \in \{1...N_{int}\} \qquad (4.14)$$

Component material balance for mixing before boiler

$$W_z \bullet Y_z^{in} = \sum_{i=1}^{N_{Sources}} w_{i,z} \bullet y_i^{in} \qquad (4.15)$$

Performance function for the kth interceptor:

$$y_k^{out} = f_k(Y_k^{in}, W_k, D_k, P_k) \qquad \forall\, k \in \{1...N_{int}\} \tag{4.16}$$

where D_k and P_k are the design and operating variables of unit k.

Splitting of the sources after the interception devices:

$$W_k = \sum_{j=1}^{N_{sink}} g_{k,j} \qquad \forall\, k \in \{1...N_{int}\} \tag{4.17}$$

Mixing for the jth sink:

$$G_j = F_j + \sum_{k=1}^{N_{int}} g_{k,j} \qquad \forall\, j \in \{1...N_{sinks}\} \tag{4.18}$$

Now considering a fresh source, the following component material balance around the mixing point of the feed to the sink can be derived:

$$G_j \cdot z_j^{in} \geq F_j \cdot y_{fresh} + \sum_{k=1}^{N_{int}} g_{k,j} \cdot y_k^{out} \qquad \forall\, j \in \{1...N_{sinks}\} \tag{4.19}$$

$$z_j^{min} \leq z_j^{in} \leq z_j^{max} \qquad \forall\, j \in \{1...N_{sinks}\} \tag{4.20}$$

Fuel substitution:

$$Fuel_q \bullet H_{fuel} + \sum_{i=1}^{N_{sources}} W_{i,s} \bullet H_i \geq D_{Steam} \tag{4.21}$$

Additionally, performance equations are needed to describe the performance of each interceptor and the waste treatment facility and relate such performance to the cost objective function.

Non-negativity of each fraction of source allocated to a sink, to an interception device, for flow of fresh resources to a sink and amount of waste:

$$g_{k,j} \geq 0 \tag{4.22}$$

$$w_{i,k} \geq 0 \tag{4.23}$$

$$w_{i,z} \geq 0 \tag{4.24}$$

$$F_j \geq 0 \tag{4.25}$$

$$Fuel_q \geq 0 \tag{4.26}$$

$$waste \geq 0 \tag{4.27}$$

4.2.3 GLOBAL OPTIMIZATION REFORMULATION

The developed program is a mixed-integer nonlinear program (MINLP). Because of the nonconvexity of the objective function and the bilinearity of several constraints, a global solution cannot be guaranteed by commercial software. Hence, we develop a global optimization procedure which is based on reformulating the problem into a linear program.

In order to reformulate the problem, we consider the problem as a supply-demand (source-sink) problem where suppliers (sources) are the process sources, the fresh raw material source, and the fresh fuel source. All these sources are utilized to fulfill specific demands of the sinks. In the formulation, the sinks are process sinks, and boiler. Figure 4.4 is a supply-demand-interception representation of the problem. This representation eliminates the integer variable from the formulation.

In addition to this, we invoke the following simplifying assumptions:

1. No mixing of sources is allowed before interception; mixing is used primarily after interception and before entering the sinks.
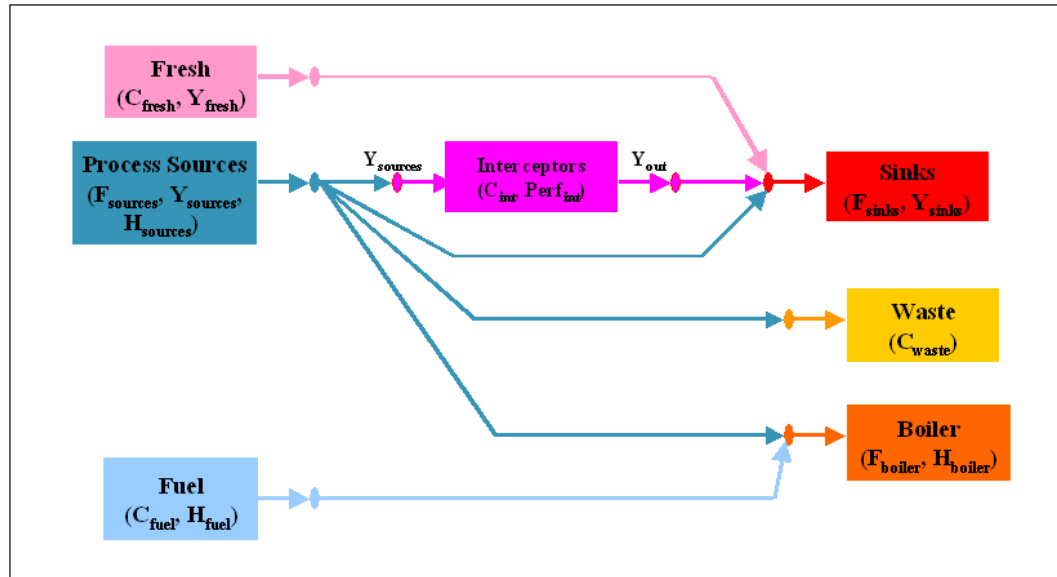
Figure 4.4: Supply-Demand (Source-Sink) Representation of Simultaneous Mass and Heat Integration Problem

2. Each interceptor is discretized into a number of interceptors with given removal efficiencies.

This is a mild assumption as it can still accurately capture the original performance of the original interceptor. To illustrate this assumption, consider figure 4.5 where we show the original interceptor while we show the discretized interceptors on the figure 4.6. We also adopt a decomposition scheme where each source is split into a number of substreams. Each substream is assigned to a discretized interceptor. The flowrate of the source i assigned to the kth interceptor is unknown (to be determined through optimization) and is designated as $w_{i,k}$. Additionally, the performance of each

interceptor is discretized in optimizing the kth interceptor, the outlet composition (y_k^{out})

is to be determined. Hence, the mathematical expression for the load to be removed from

source i is given by $F_i * (y_i^{in} - y_k^{out})$ which contains a bilinear term. Now, discretizing

the kth interceptor into multiple interceptors each having a given removal efficiency

(o_k). The flowrate exiting each discretized interceptors ($w_{o,k,j}$) is unknown and to be

determined by mathematical formulation. Therefore, the load to be removed from the

source using interceptor k can be matched by the load removed by the discretized

interceptors as follows:

$$F_i \bullet (y_i^{in} - y_k^{out}) = \sum_{k_i=1}^{NK_i} w_{o,k,j} \bullet (y_{i=k}^{in} - Y_k^{out}) \qquad (4.28)$$

where

$$Y_k^{out} = (1 - o_k) \bullet y_{k=i}^{in} \qquad (4.29)$$

Since the values of the various o_k 's are fixed, the right hand side of Eq. (4.29) is a linear

term and the remaining task is to determine the optimal value of each $w_{o,k,j}$. This way,

the performance of one interceptor with unknown removal efficiency can be exactly

matched by the performance of multiple interceptors each with a given removal

efficiency. Additionally, the modeling and costing of the interceptor can now be taken

outside the optimization formulation and transformed into a pre-synthesis task. For a

given source and removal efficiency, detailed simulation and costing can be carried out

ahead of synthesis thereby eliminating a significant source of nonconvexity. The

reformulated problem is schematically represented by figure 4.7.

Source, i   Interceptor, k   Sink, j
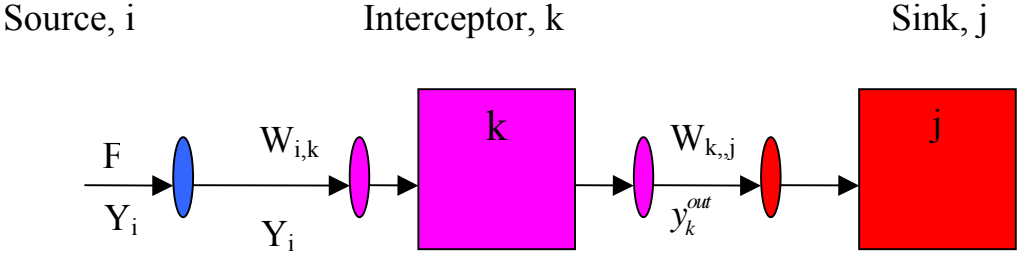
Fig 4.5: A Single Interceptor
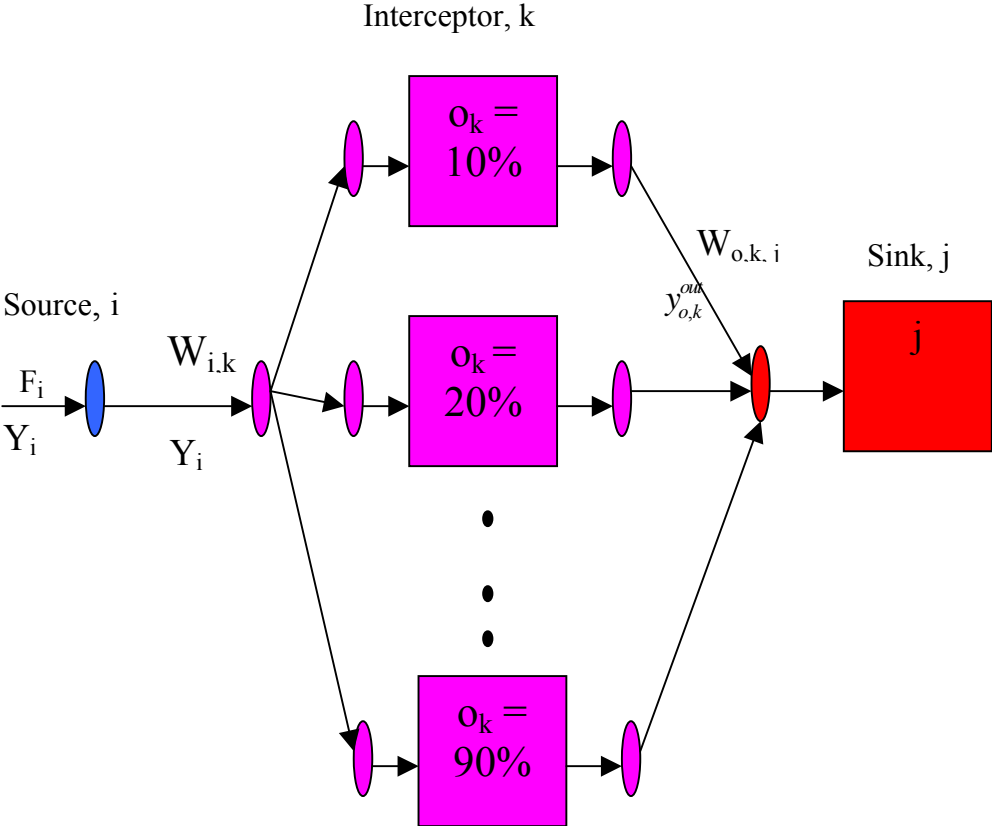
Interceptor, k

Fig 4.6: Discretizing the Interceptor
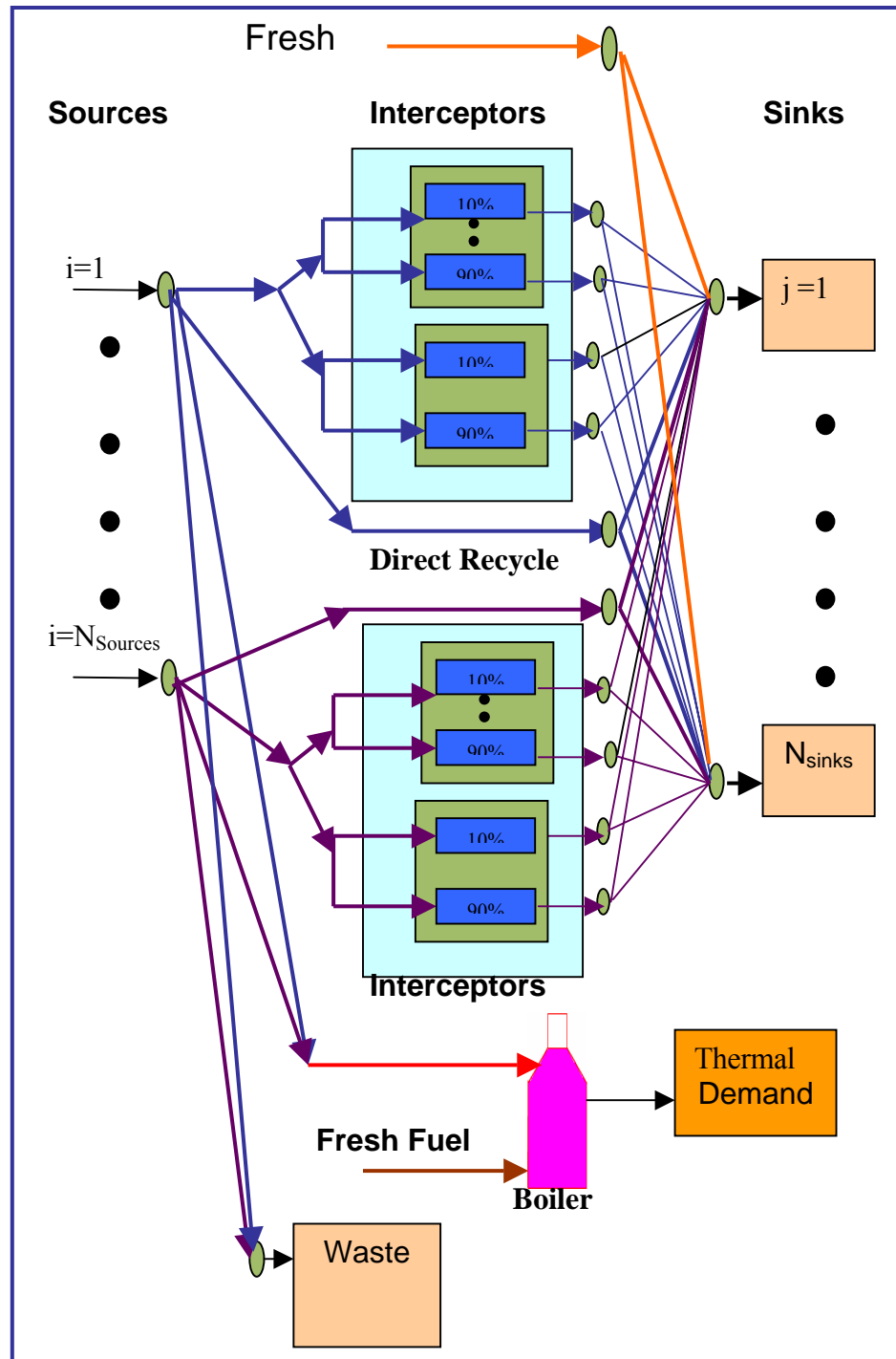
Fig 4.7: Structural Representation of the Reformulated Problem

3. As a result of pre-synthesis calculations, each interceptor k has a known removal efficiency o_k and cost Co,k (\$/Load removed). For a given source with known inlet composition, we assume that the total annualized cost of the interceptor is proportional to the removed load of the targeted species in the interceptor. Therefore, we can express the interception cost for the kth unit with o^{th} efficiency as:

$$Interception\_Cost_{k,o} = C_{o,k} \bullet W_{o,k,j} \bullet y^{in}_{k=i} \bullet o_k \tag{4.30}$$

This is a linear term with the only variable being $w_{o,k,j}$.

We are now in a position to express the reformulated mathematical formulation.

4.2.3 MATHEMATICAL REFORMULATION

Objective Function:

Minimize, total annual operating cost =

$$C_{Fresh} \cdot \sum_{j=1}^{N_{Sinks}} W_{m=fresh,j} + \sum_{k=1}^{N_{int}} C_{o,k} \cdot o_k \cdot W_{o,k,j} \cdot y^{in}_{k=i} + C_{waste} \cdot \sum_{i=1}^{N_{sources}} W_{i,l=waste} + C_{Fuel} \cdot W_{r=fuel,q=boiler} \tag{4.31}$$

Here, $W_{m=fresh,j}$ is flowrate of fresh raw material to the process sink

$C_{o,k}$ is cost of impurity removal with interceptor k of efficiency o

$W_{o,k,j}$ is flow to the sink j through interceptor k with efficiency o

$y^{in}_{k=i}$ is the inlet composition of source i into interceptor k

$W_{i,l=waste}$ is the flowrate from process sources to waste treatment

$W_{r=fuel,q=boiler}$ is the flowrate of fresh fuel to boiler

This equation defines the annual operating cost for the simultaneous mass and heat integration operation. The operating cost includes the cost of fresh, fuel, interception and waste treatment. This equation is also explained earlier.

Subject to following constraints:

Source (supply) constraint:

$$\sum_{i=1}^{N_{sources}} \sum_{j=1}^{N_{sinks}} W_{i,j} + \sum_{i=1}^{N_{sources}} \sum_{k=1}^{N_{int}} W_{i,k} + \sum_{i=1}^{N_{sources}} W_{i,l=waste} + \sum_{i=1}^{N_{sources}} W_{i,q=boiler} \leq F_i \qquad (4.32)$$

Here, $W_{i,j}$ is the flowrate from process sources to sinks (direct recycle)

$W_{i,k}$ is the flowrate from process sources i to the interceptors k

$W_{i,l=waste}$ is the flowrate from process sources i to the waste treatment l

$W_{i,q=boiler}$ is the flowrate from process sources i to the boiler q

F_i is the available flowrate at process source i

This equation defines that the total output flowrate from any process source should be less then or equal to the available flowrate at that process source.

Sink (demand) constraint:

$$\sum_{j=1}^{N_{sinks}} \sum_{i=1}^{N_{sources}} W_{i,j} + \sum_{j=1}^{N_{sinks}} W_{m=fresh,j} + \sum_{j=1}^{N_{sinks}} \sum_{k=1}^{N_{int}} \sum_{o=1}^{N_{eff}} W_{o,k,j} \geq F_j \qquad (4.33)$$

Here, $W_{i,j}$ is the flowrate from process sources to sinks (direct recycle)

$W_{m=fresh,j}$ is flow rate of fresh raw material to the process sink

$W_{o,k,j}$ is flow to the sink j through interceptor k with efficiency o

(interception)

F_j is the flowrate demand at process sink j

This equation defines that the total input (flowrate) to the sink should be greater or equal to the demand of the sink.

Pollutant removal at k^{th} interceptor:

$$y_{o,k}^{out} = (1 - o_k) \bullet y_{k=i}^{in} \qquad (4.34)$$

Here, $y_{o,k}^{out}$ is the outlet composition from interceptor k with efficiency o

o_k is the set of efficiency for interceptor k

$y_{k=i}^{in}$ is inlet composition of source i to interceptor k

Sink concentration constraint:

$$\sum_{j=1}^{N_{\sin ks}} \sum_{i=1}^{N_{sources}} W_{i,j} \bullet Y_i + \sum_{j=1}^{N_{\sin ks}} W_{m=fresh,j} \bullet Y_m + \sum_{j=1}^{N_{\sin ks}} \sum_{k=1}^{N_{int}} \sum_{o=1}^{N_{eff}} W_{o,k,j} \bullet Y_{o,k}^{out} \leq F_j \bullet Y_j \qquad (4.35)$$

Here, $W_{i,j}$ is the flowrate from process sources to sinks (direct recycle)

Y_i is the composition of source i

$W_{m=fresh,j}$ is flow rate of fresh raw material to the process sink j

Y_m is the composition if fresh raw material

$W_{o,k,j}$ is flow to the sink j through interceptor k with efficiency o

(interception)

$y_{o,k}^{out}$ is the outlet composition from interceptor k with efficiency o

This equation defines that the composition of the mixture at the inlet of sink must satisfy the concentration constraint of the sink.

Flow assignment equation:

$$\sum_{\substack{i=1 \\ i=k}}^{N_{sources}} \sum_{\substack{k=1 \\ i=k}}^{N_{int\,e}} W_{i,k} = \sum_{j=1}^{N_{sin\,ks}} \sum_{k=1}^{N_{int}} \sum_{o=1}^{N_{eff}} W_{o,k,j} \qquad (4.36)$$

Here, $W_{i,k}$ is the flowrate from process sources i to the interceptors k

$W_{o,k,j}$ is flow to the sink j through interceptor k with efficiency o

This equation defines that the all the flowrates coming from a process source to a particular interceptor is equal to the summation of all the flowrate to the sink from that interceptor with the set of efficiencies. This equation relates input to an interceptor to the output from the interceptor. The interceptor receives flow from the process sources and after intercepting impurities according to the removal efficiencies, send the same flow to the sinks.

Boiler demand equation:

$$\sum_{i=1}^{N_{sources}} W_{i,q=boiler} \bullet H_i + W_{r=fuel,q=boiler} \bullet H_{fuel} \geq H_{boiler} \qquad (4.37)$$

Here, $W_{i,q=boiler}$ is the flowrate from process sources i to the boiler q

H_i is the heating value of source i

$W_{r=fuel,q=boiler}$ is the flowrate of fresh fuel to boiler

H_{fuel} is the heating value of fresh fuel

This equation defines that the heating value generated by process sources in the boiler and the heating value generated by the fresh fuel in the boiler should satisfy the boiler demand.

Non-negativity constraints:

All $W's \geq 0$ (4.38)

All $F's \geq 0$ (4.39)

All $Y's \geq 0$ (4.40)

These equations define that, all the flowrates are non-negative, all the supplies and demands are non-negative and all the compositions are also non-negative.

4.3 STEAM HEADER BALANCE

Steam header balance is developed utilizing the result from simultaneous mass and heat integration study, and also from specific process situation and process data. Simultaneous mass and heat integration analysis provide information about the availability of the process sources for producing steam within the process utilizing the boiler. Also steam can be generated through waste heat boilers or heat recovery steam generators. Figure 4.8 illustrates generation of steam header balance.

Steam header balance can be generated from the following equations

Sources of Data for Steam Header Balance:

Steam for heating purposes (From GCC): $D_{Steam-Heating}$

Steam as an MSA (from previous analysis): $D_{Steam-MSA}$

Total Steam Demand, $Steam_{demand} = D_{Steam-Heating} + D_{Steam-MSA}$ (4.41)

Steam Supply from Waste (from previous analysis): $S_{Steam-Waste}$

Steam Supply from Hot Exhaust (Process Data): $S_{Steam-HT}$

Steam Supply from Process Fuel: $S_{Steam-Process}$

Total Steam Supply, $Steam_{Supply} = S_{Steam-Waste} + S_{Steam-HT} + S_{Steam-Process}$ (4.42)
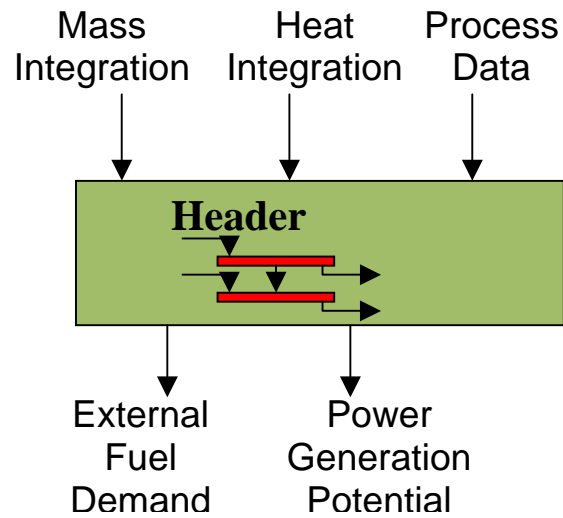
Fig 4.8: Steam Header Generation (Mathematical Programming)

4.4 COGENERATION

In this section we calculate the optimal cogeneration potential from a given set of steam headers. Steam headers were generated in the previous section. From the analysis so far, following information at each steam headers are available:

- Steam header pressure

- Steam header temperature

- Demand of steam at any given steam header

- Supply of steam at any given steam header

First the steam enthalpies at different header conditions are calculated from the available steam header data. For calculating enthalpies at different steam header levels, we utilize the enthalpy calculation equations and correlations developed by Irvine and Liley (1984).

Now from the steam header data surplus headers and deficit headers are identified. Surplus headers are the headers where the steam supply is greater then steam demand. Similarly deficit headers are the headers where the supply of steam is less then the demand of steam. Steam can only flow from a higher-pressure header to a lower pressure header according to the supply and demand of steam at any given header.

From the supply and demand data, and calculated steam enthalpy at any given header, mass flow rate of excess or deficit steam at a given header is calculated. Following equation is used to calculate the mass flowrate of steam at any header:

Steam Mass Flow = (Supply of steam- Demand of steam)/enthalpy of steam   (4.43)

For this study steam turbine efficiency has been assumed to be equal to 0.7. From all these data, extractable work method is utilized mathematically to determine the optimum cogeneration potential of the steam header system. Extractable work between two steam headers is defined by the multiplication between steam mass flow and turbine efficiency and enthalpy difference between the two headers. Optimization formulation is utilized to determine the optimum cogeneration potentials from a given set of steam headers.

4.4.1 PROBLEM STATEMENT

For this part of the mathematical programming the problem statement becomes as follows.

Given,

- A set of steam headers i with

  - Pressure P_i,

  - Temperature T_i,

  - Steam supply S_i, and

  - Steam demand D_i

- Determine optimum cogeneration potential P_{cogen} from this header set

- Determine availability of excess steam in the header set

- Determine optimum allocation of excess steam

- Determine condensing power generation possibility

For ensuring the global optimality of the solution, the problem is formulated as a linear programming problem. The result from this part is utilized to determine the electricity consumption of the plant from external sources. This solution is utilized to solve the last portion of the overall objective function, i.e.:

$$\text{Minimize } C_{power} \bullet Power \tag{4.44}$$

4.4.2 PROBLEM REPRESENTATION

Figure 4.9 shows schematically the cogeneration problem. Steam can be passed from any higher-pressure header to any lower pressure header according to the supply and demand condition of the headers. Steam is usually passed through a turbine while passing it from high pressure to lower pressure header. Steam is expanded in the turbine, and supplied to the lower pressure-header according to the process requirement. Turbine is utilized to cogenerate power, or turbine can also produce shaft work to run mechanical

devices. Any power cogeneration within the systems reduces the external power consumption cost for the process.
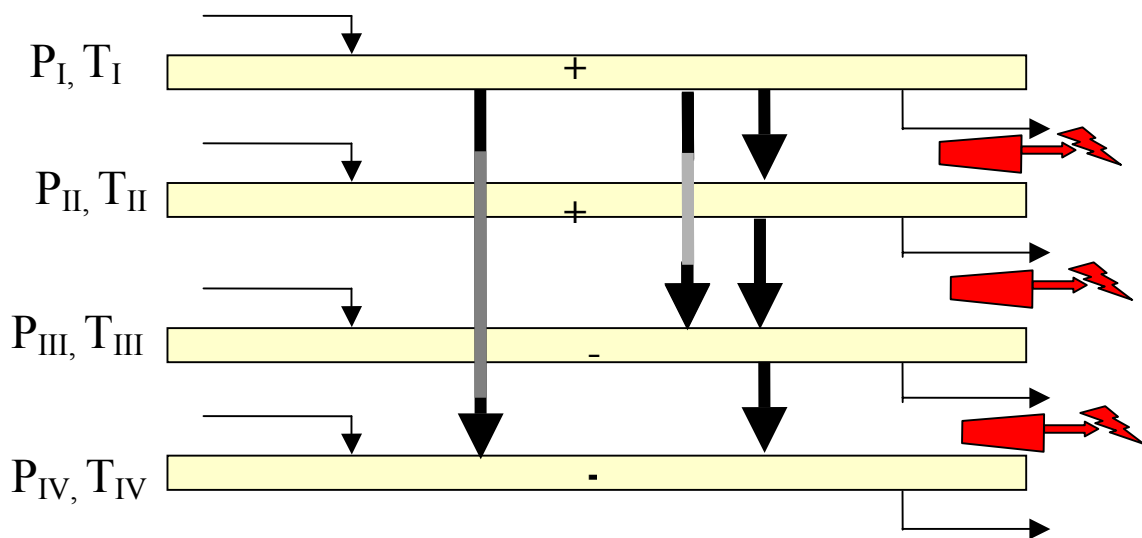
Fig 4.9: Cogeneration Potential (Mathematical Programming)

4.4.3 MATHEMATICAL FORMULATION

Objective Function: Maximize total power $P_{total} = P_{cogen} + P_{condenssation}$ (4.45)

Subject to the following constraints:

Defining cogeneration between two headers.

$$P_{cogeneration} = \sum_{i>j} \left[W_{ij} \times \eta_{eff} \times (H_i - H_j) \right]$$ (4.46)

Here, $W_{i,j}$ is flow of steam between steam header i and steam header j,

η_{eff} is the efficiency of the turbine,

H_i, H_j are the enthalpies of steam at header conditions i and j.

The equation is applicable to all headers where index of i (steam headers) is greater then that of j (steam headers). This condition ensures that the steam is only flowing from a higher-pressure header to a lower pressure header.

Defining condensing power generation:

$$P_{condensation} = \sum_i \left[W_{ik} \times \eta_{eff} \times (H_i - H_k) \right] \tag{4.47}$$

Here, $W_{i,k}$ is flow of steam between source header and condensing turbine

η_{eff} is the efficiency of the condensing turbine

H_i, H_k are the enthalpies of steam at header level and at

condensing turbine outlet

This equation defines the condensing power generation. Steam from any header level can flow to the condensing turbine, since the condensing turbine has lower pressure then any of existing steam headers. For steam to flow from any given header to condensing turbine, the header should have existing availability of excess steam.

Source (supply) constraint equation:

$$\sum_{\substack{i \in surplus \\ i > j}} W_{ij} + \sum_{i \in surplus} W_{ik} \leq S_i \tag{4.48}$$

Here, S_i is the availability of surplus steam at a header i

This equation explains that the total steam flow from any surplus header to a deficit header or condensing turbine must not exceed the availability of surplus steam in the surplus header. The condition that the index of i is greater then the index of j makes sure that the steam is only flowing from a higher-pressure header to a lower pressure header.

Demand (sink) constraint equation:

$$\sum_{\substack{j \in deficit \\ i > j}} W_{ij} \geq D_j \tag{4.49}$$

Here, D_j is the deficit of steam at header j

This equation explains that the steam flows coming to a deficit header from all the surplus headers above it, must meet the steam demand at the deficit header. Again, the condition that the index of i is greater then the index of j makes sure that the steam is only flowing from a higher-pressure header to a lower pressure header.

Flow from a deficit header:

$$\sum_{\substack{i \in deficit \\ j \in deficit \\ i > j}} W_{ij} + \sum_{i \in deficit} W_{ik} \leq \sum_{\substack{j \in deficit \\ i > j}} W_{ij} + D_{j\ =\ deficit} \tag{4.50}$$

Here, $W_{i,j}$ $\substack{i \in deficit \\ j \in deficit \\ i > j}$ is the steam flow between deficit headers

$W_{i,k}$ $\substack{i \in deficit}$ is the steam flow from a deficit header to condensing turbine

$W_{i,j}$ $\substack{j \in deficit \\ i > j}$ is the steam flow to a deficit header from any upper level header

$D_{j=deficit}$ is the demand of steam at a deficit header

This equation defines the possibility of steam flow from any deficit headers to any lower level deficit headers or to the condensing turbine. The equation defines that for the deficit header to become a surplus header (header with existing excess steam), the flow to the deficit header should be greater then it's initial demand i.e. output from a deficit header should be less than or equal to it's input plus it's demand. It is assumed

that a deficit header can only pass steam to another deficit header or to the condensing turbine.

Excess steam at surplus header:

$$X_{i=surplus} = S_i - \sum_{\substack{i>j \\ i\in surplus}} W_{ij} + \sum_{\substack{j>i \\ i\in surplus \\ j\in surplus}} W_{ji} \qquad (4.51)$$

Here, $X_{i=surplus}$ is the excess steam at surplus header

S_i is the surplus steam at header i

$W_{\substack{i,j \\ i>j \\ i\in surplus}}$ is the steam flowrate from surplus header to headers at lower levels

$W_{\substack{j,i \\ j>i \\ i\in surplus \\ j\in surplus}}$ is the steam flowrate to a surplus header from other surplus headers

         at higher levels

This equation calculates the amount of excess steam available at any surplus steam header. The equation defines the excess steam to be equivalent to the existing surplus steam minus output of steam from surplus header plus any input of steam to the surplus header. It is assumed that a surplus header can only get input steam from another surplus header at higher level.

Excess steam at deficit header:

$$X_{j\in deficit} = \sum_{i>j} W_{ij} - D_{j=deficit} - \sum_{\substack{j>i \\ i\in deficit \\ j\in deficit}} W_{ji} \qquad (4.52)$$

Here, $X_{j=deficit}$ is the excess steam at deficit header

$D_{j-deficit}$ deficit of steam at header j

$W_{i,j}$ is the steam flowrate from upper level headers to deficit header at
$\underset{\substack{i>j \\ j \in deficit}}{}$

lower level

$W_{j,i}$ is the steam flowrate from a deficit header to other deficit headers
$\underset{\substack{j>i \\ i \in dificit \\ j \in deficit}}{}$

at lower level

This equation calculates the amount of excess steam available at deficit headers. The equation defines excess steam at deficit headers to be equivalent to the supply of steam to a deficit header from upper level headers minus the deficit (demand) of steam at that steam header and minus the flow of any steam from this header to any deficit header at the lower level. It is assumed that a deficit steam header can only pass steam to another deficit header at lower level.

The excess steam, if available, are utilized either for supplying the cooling load of the process or for condensing power generation. If supplying the cooling load of the process, then the header balance is updated by adding the steam required to supply the cooling in the steam header at appropriate header level. So the cooling load is added as:

Steam for cooling purposes : $D_{Steam-Cooling}$

Total Steam Demand, $Steam_{demand} = D_{Steam-Heating} + D_{Steam-MSA} + D_{Steam-Cooling}$ (4.53)

Total Steam Supply, $Steam_{Supply} = S_{Steam-Waste} + S_{Steam-HT} + S_{Steam-Process}$ (4.54)

So, from this steam header balance is updated, and utilizing this updated header balance, new cogeneration potential and availability of any further excess steam is determined. The cogeneration potential is enhanced due to adding the cooling demand in the steam header. If further excess steam is available, then this steam is passed through a

condensing turbine to enhance the total power output from the steam header system. This result is utilized to determine the amount of external power that is required for the plant. For determining the external power consumption, the objective function is:

$$\text{Minimize } C_{power} \bullet P_{external} \qquad (4.55)$$

Here, C_{power} is the cost of power purchase from external sources

$P_{external}$ is the amount of external power purchase.

Subject to the following constraint:

$$P_{cogen} + P_{condensation} + P_{external} \geq D_{power} \qquad (4.56)$$

Here, P_{cogen} is the power produced by cogeneration

P_{conden} is the power produced by condensing turbine

$P_{external}$ is the amount of external power purchase

D_{power} is the total power demand of the plant

It is assumed that the power by cogeneration and power by condensing turbine incur a small operating cost, as they are produced from the steam (excess) already available within the plant (virtually for free). It is worth noting that the plant is not producing steam for the power generation purpose only. It is producing steam for supplying the heating load of the plant and for fulfilling other non-heating demand of the plant. Hence, the power generated by cogeneration incurs a marginal cost of steam. Another noteworthy point is that if the plant produces steam for power generation purposes only, it is unlikely that the cost of this power will be competitive with the cost of power provided by a utility company, which has the benefits of expertise, the

contracts, the know-how, and the economy of scale. So, in the event of no excess steam, cogeneration is always a better option then the condensing power generation.

4.5 CASE STUDY

We show the applicability of the mathematical programming approach by a detailed case study. Here we choose the same system that was utilized for the graphical technique case study. The process flowsheet is shown in Fig. 4.10. Tables 4.1-4.7 summarize the data for the heating, cooling, process sinks, process sources, and external resources for the case study.
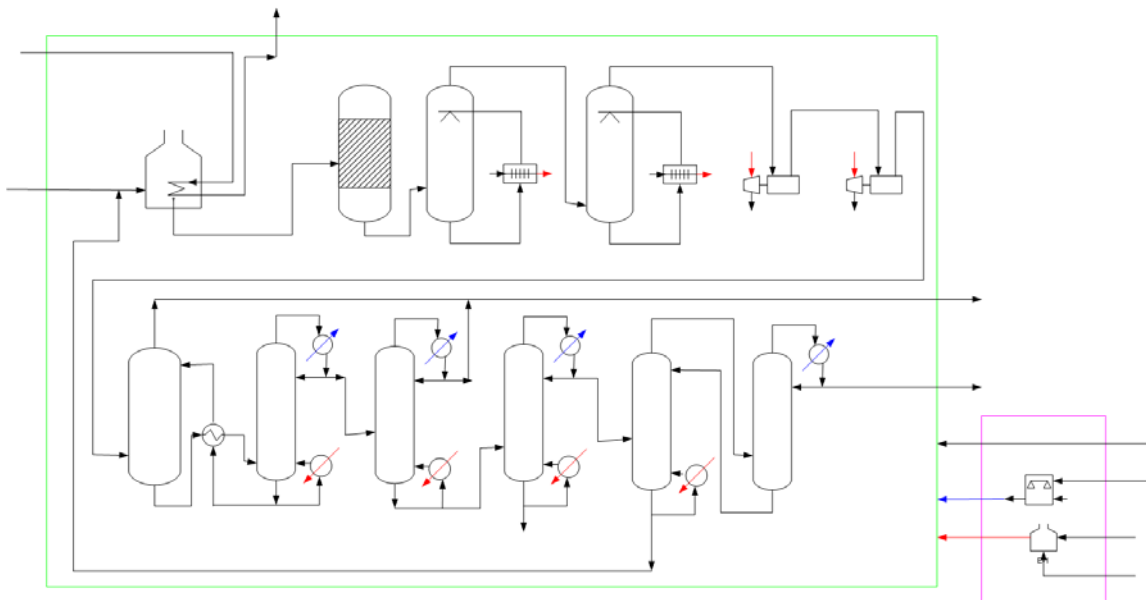
Fig 4.10: Propylene by Catalytic De-Hydrogenation of Propane

For comparison, the same process of propylene by catalytic hydrogenation of propane is utilized. After solving the case study, the results from both the approaches will be compared.

Table 4.1: Process Heating/Cooling Data

Stream	T_{Supply} (^{o}F)	T_{Target} (^{o}F)
H1	170	130
H2	150	100
H3	190	100
H4	160	80
C1	90	170
C2	100	140
C3	130	190
C4	150	170

Table 4.2: Raw Material Data

Raw Material	Cost ($/lb)	Mass Fraction (Impurities)
Fresh	0.11	0.00

Table 4.3: Process Sink Data

Sink	Flowrate (lb/hr)	Maximum Inlet Mass Fraction (Impurities)	Maximum Inlet Load (kg/hr)
VA Process Reactor	34000	0.20	6800

Table 4.4: Process Source Data

Source	Flowrate (lb/hr)	Outlet Mass Fraction (Impurities)	Maximum Inlet Load (kg/hr)	Heating Values (Btu/lb)
De-ethanizer	10000	0.48	4800	1000
Absorption Column	20000	0.65	13000	1500

Table 4.5: Fresh Fuel Data

Fresh Fuel	Cost ($/MMBTU)	Heating Value (Btu/lb)
Fuel	2.6	13400

Table 4.6: Process Fuel Data

Process Fuel	Flow (lb/hr)	Heating Value (Btu/lb)
Depropanizer Bottom	20000	13400

Table 4.7: Electricity Consumption Data

	Demand (MW)	Cost ($/kwhr)
Electricity	12	0.06

Cost of waste treatment is assumed to be equivalent to $0.0022/lb of waste. It was determined in the previous chapter (through the grand composite analysis) that the heating demand for the plant is 182 MMBTU/h and the cooling demand is 166 MMBTU/h.

As mentioned earlier, the plant purchases all its required raw material and utilities and electricity from external sources, without considering recycle, reuse opportunity and also without considering recovery of thermal or material value from waste and producing electricity through cogeneration. Furthermore, the plant has an existing annual operating cost of around $40 million. Figure 4.11 shows the existing situation and operating cost of the plant. The objective here is to utilize the mathematical programming approach to optimize the process.
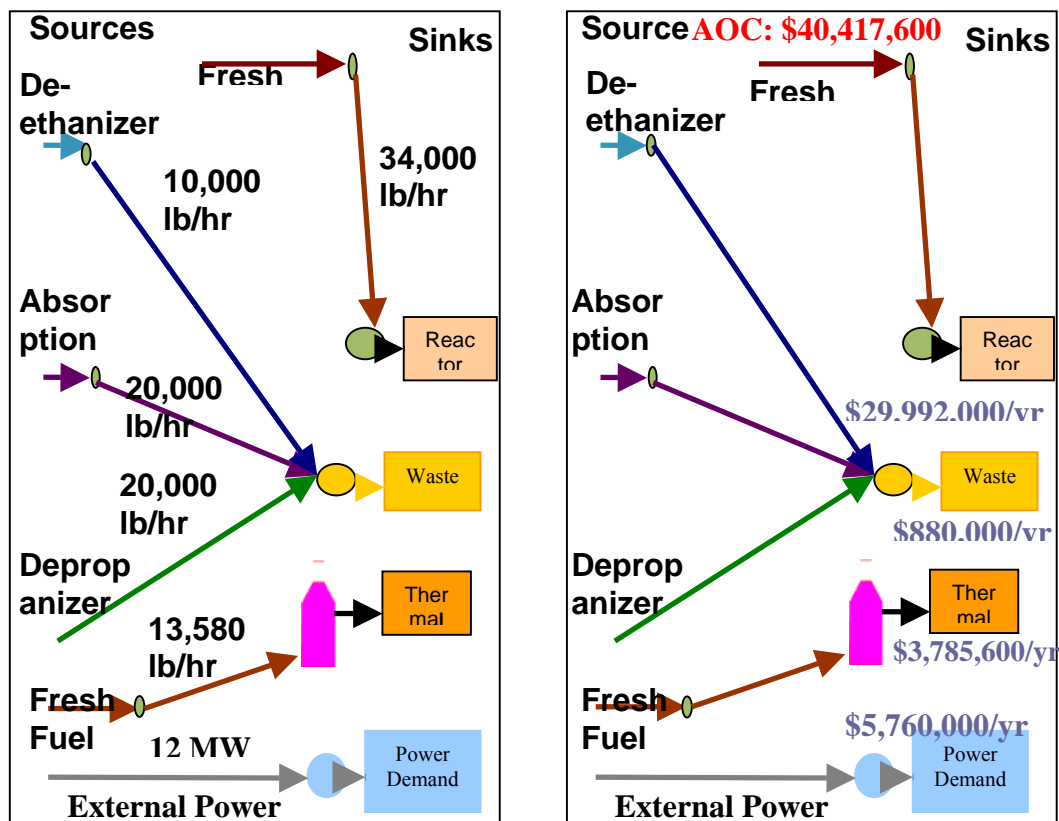
Fig 4.11: Existing Flow and Operating Cost

The mathematical approach enables the consideration of interception for the removal of contaminants. For the waste interception from the two process sources, two different technologies can be utilized. The technologies are steam stripping and ion exchange. The cost data tables for these two techniques at different removal efficiencies for both the process sources are summarized by Tables 4.8 and 4.9. Optimization utilizes these tables to determine the optimum removal efficiencies according to the cost and process requirements.

Table 4.8: Cost at Different Contaminant-Removal Efficiencies for De-Ethanizer

Technology	Removal Efficiency (%)	Cost ($/lb removed)
Stripping	10	0.068
	20	0.083
	30	0.102
	40	0.125
	50	0.146
	60	0.164
	70	0.188
	80	0.224
	90	0.296
Ion Exchange	10	0.081
	20	0.099
	30	0.122
	40	0.149
	50	0.175
	60	0.196
	70	0.225
	80	0.268
	90	0.355

Table 4.9: Cost at Different Contaminant-Removal Efficiencies for Absorption

Technology	Removal Efficiency (%)	Cost ($/kg removed)
Stripping	10	0.068
	20	0.083
	30	0.102
	40	0.125
	50	0.146
	60	0.164
	70	0.188
	80	0.224
	90	0.296
Ion Exchange	10	0.081
	20	0.099
	30	0.122
	40	0.149
	50	0.175
	60	0.196
	70	0.225
	80	0.268
	90	0.355

Now, the objective of the case study is to develop a revised process configuration, which optimizes:

- Heating and cooling utilities

- Waste recycle

- Waste interception for material values

- Waste conversion to thermal energy

- External fuel consumption

- External power consumption

- Power cogeneration

For achieving all those purposes, the proposed hierarchical mathematical programming technique is utilized. In the following section, we describe the approach in successive sections.

4.5.1 SIMULTANEOUS MASS AND HEAT INTEGRATION

There are two process sources with recoverable material value and thermal value. We also have a sink for material value and a sink for thermal value. We also have one fresh raw material source for material sink and a fresh fuel source for thermal sink. There is a process source with only recoverable thermal value. Furthermore, there exists one waste treatment facility where all unused process sources are treated. Our objective here is to determine the optimum allocation of process sources to the process sinks for reducing annual operating cost. Material and thermal recoveries are competing and the mathematical approach will reconcile the extent of each one. The problem is depicted in the figure 4.12.

99
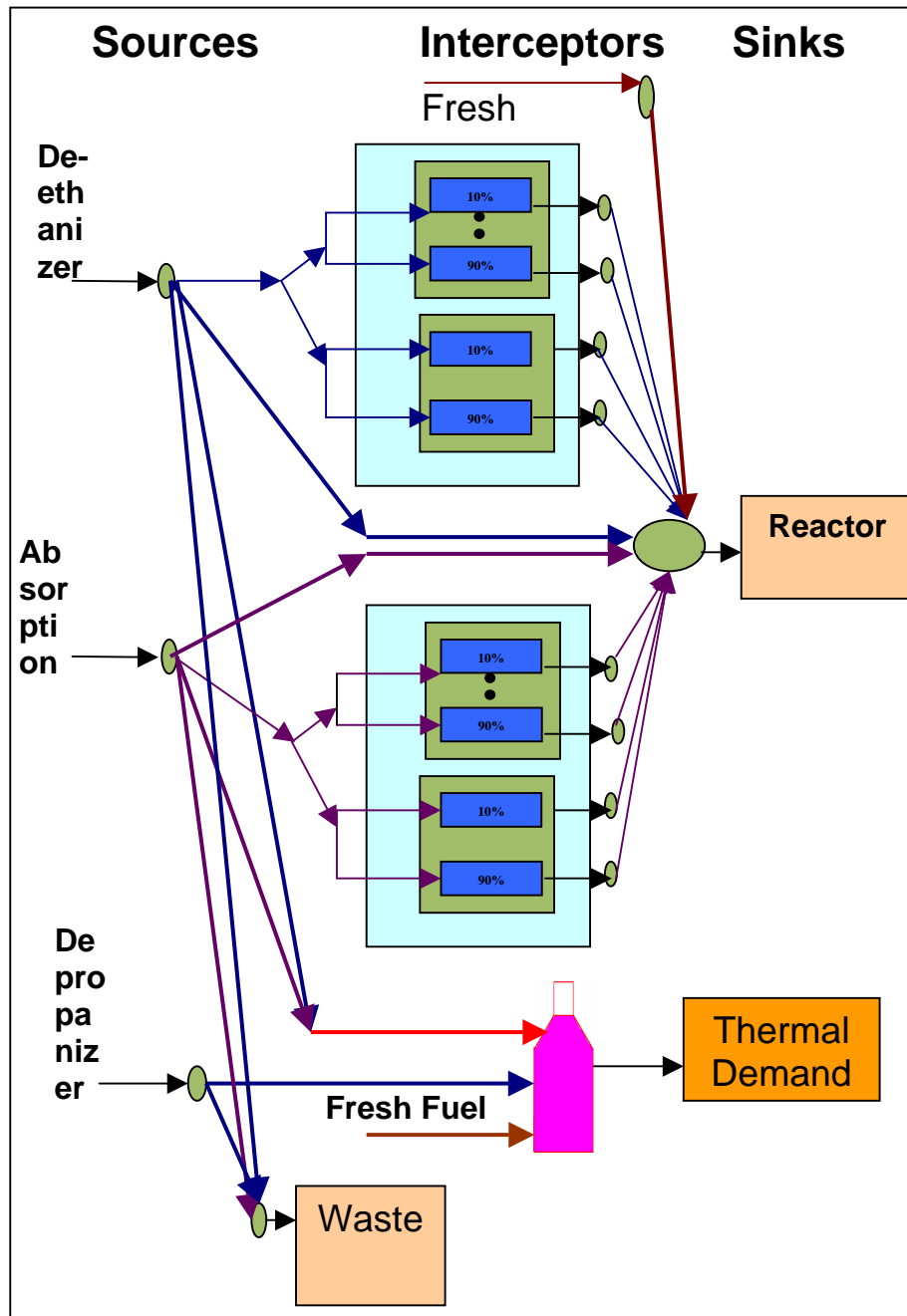
Fig 4.12: Schematic Representation of Case Study

Unlike the graphical technique where for mass integration, direct recycle preceded the interception network, here in mathematical programming those two options with thermal recovery options are analyzed simultaneously. Figure 4.13 shows the result of the simultaneous mass and heat integration analysis using optimization formulation.
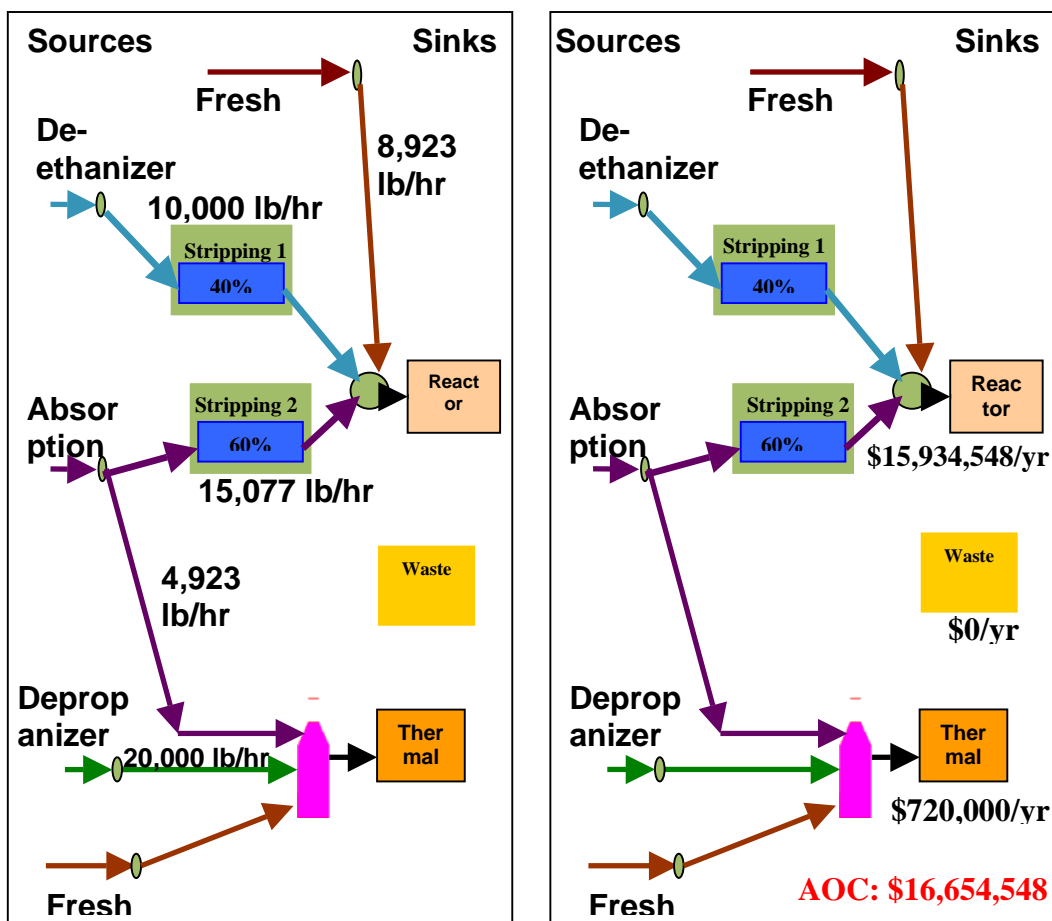
Fig 4.13: Simultaneous Mass and Heat integration (Math Programming)

Here from the result, we can see that, all the process sources are utilized either for material value recovery purpose or thermal value recovery purpose. The waste

treatment cost is eliminated. Also in the material value recovery, no direct recycle is utilized. The product from the de-ethanizer is sent through a stripping column with 40% removal efficiency. Also at least 75% of the product from absorption column is sent through stripping column with 60% removal efficiency, and the rest is sent for thermal value recovery. The fresh consumption is reduced drastically for these operations. The comparison of these results with the existing situation are shown by Figure 4.14.
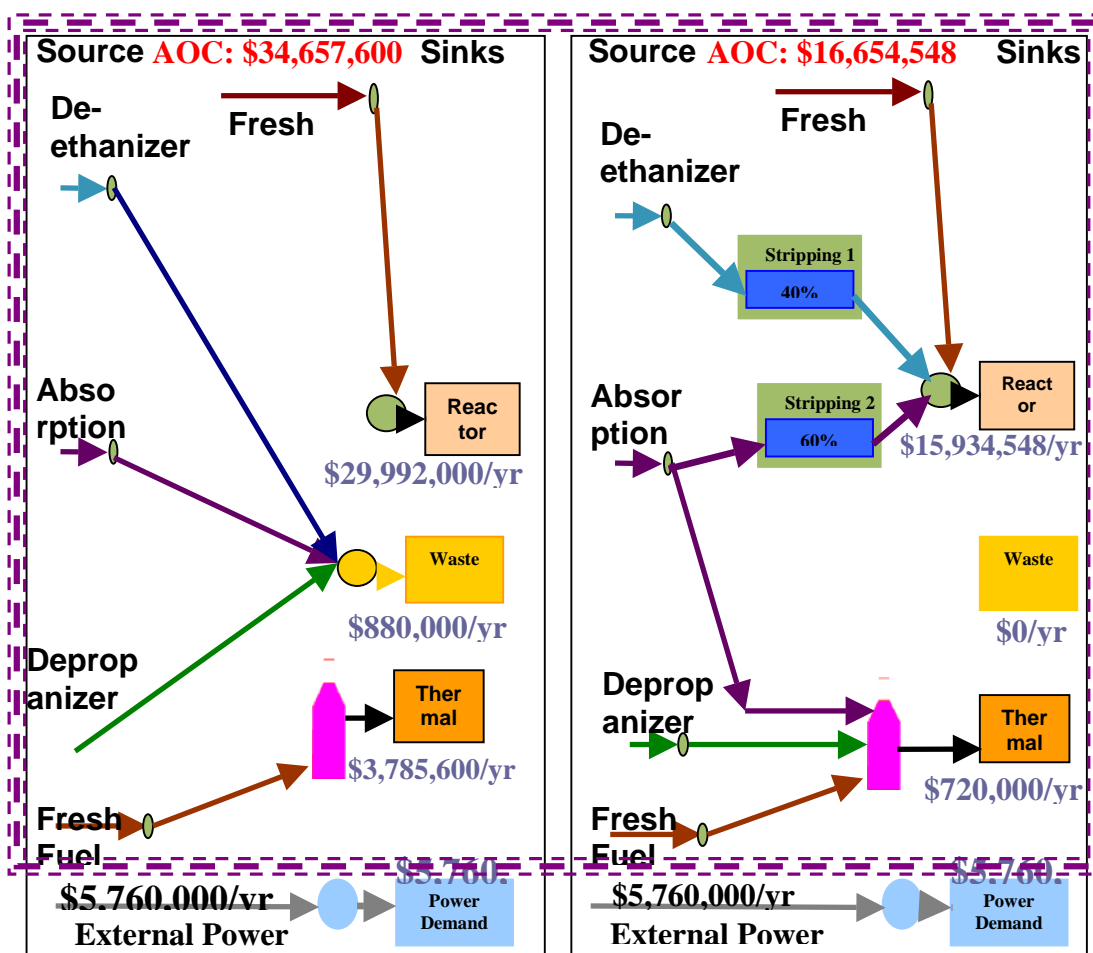
Fig 4.14: Comparison of Existing Situation and Math Programming Result

As can be seen from Figure 4.14, the consumption of raw materials and fresh fuel is reduced as a result of the material and thermal value recovery from the process sources. Additionally, the cost of waste treatment is eliminated. Consequently, the annual operating cost is reduced from around $35 million to around $16.5 million. The raw material cost is reduced by almost 50% and the fresh fuel consumption cost is reduced by almost 80% of the initial value.

4.5.2 STEAM HEADER BALANCE

From the simultaneous mass and heat integration analysis, it is found that around 147.5 MMBTU/hr of steam can be generated using the process sources. So another 35 MMBTU/hr of required steam is generated by external fuel. In addition to this, the result from simultaneous mass and heat integration analysis shows that steam-stripping operation is chosen for the interception. For steam stripping operation we require 10 MMBTU/hr steam at MP level. This is added in the steam header as a steam demand at MP level.

Further process data analysis reveals that, more steam can be generated from the process utilizing waste heat boiler and heat recovery steam generators. Following is the steam production data from process utilizing WHB and HRSG.

HP: 90.5 MMBTU/hr (by heat recovery from hot exhaust)

MP: 10 MMBTU/hr (HRSG)

LP: 20 MMBTU/hr (HRSG)

Therefore, the process has excess steam and the external fuel consumption for steam production is totally eliminated. Also, the process has some non-heating steam

demand for the stripping column operation and steam driven equipments. Due to these additional availabilities of steam from the process it eliminates the external fuel consumption for the plant. This reduces the annual operating cost of the plant further. The updated annual operating cost is around $16 million. Figure 4.15 shows the updated annual operating cost.
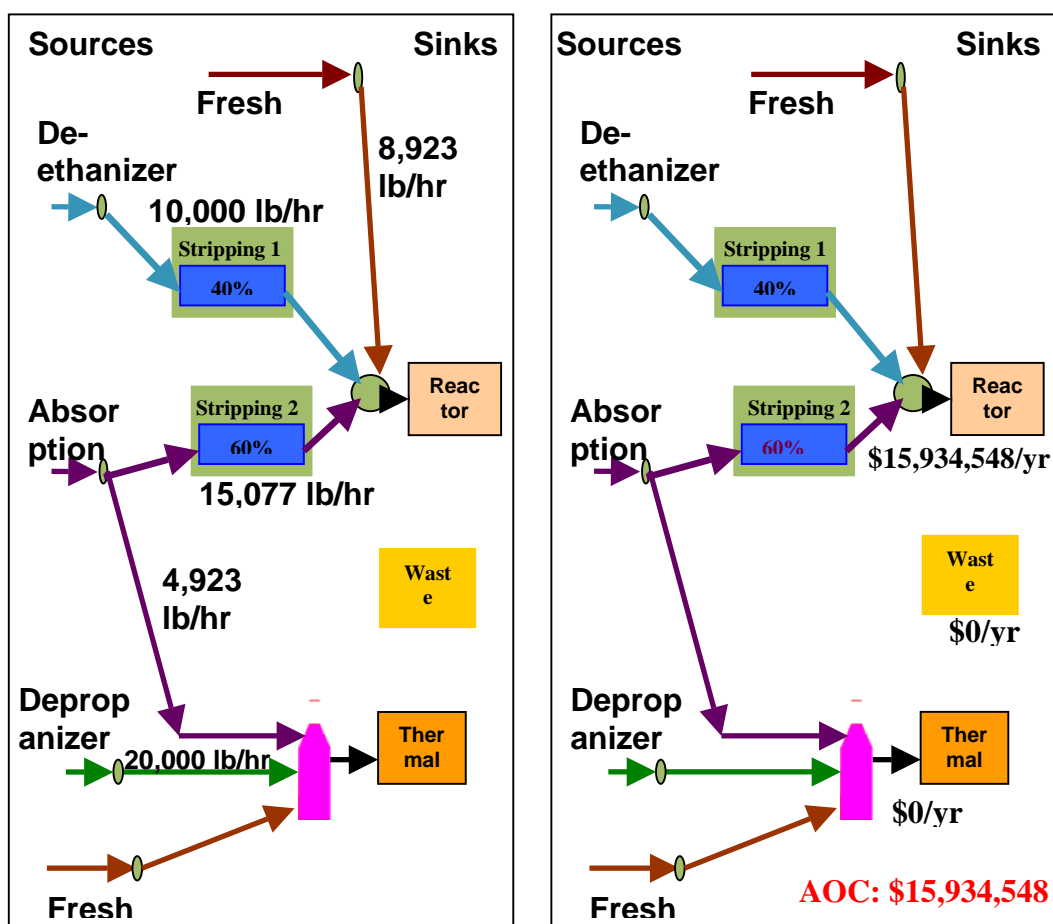
Fig 4.15: Elimination of External Fuel and Reduction in Operating Cost

From all these data, the steam balance is generated as shown by Table 4.10.

Table 4.10: Steam Header Balance (Mathematical Technique)

Steam	Pressure (psia)	Temperature (^{o}F)	Supply (MMBTU/h)	Demand (MMBtu/h)
HP	600	800	238	0
MP	130	350	10	25
LP	40	270	20	192

4.5.3 COGENERATION

The steam header balance is now utilized to determine the cogeneration potential from the given steam header. Any cogeneration within the system will result in reduction of external power purchase cost, which will also reduce the annual operating cost of the plant. Any excess steam available within the steam header is utilized to supply part of the required cooling load and the steam header balance is updated as shown by Table 4.11.

Table 4.11: Updated Steam Header Balance (Mathematical Technique)

Steam	Pressure (psia)	Temperature (^{o}F)	Supply (MMBTU/h)	Demand (MMBtu/h)
HP	600	800	238	0
MP	130	350	10	36
LP	40	270	20	192

Figure. 4.16 shows the result from the cogeneration and power section. This diagram shows the total result from the hierarchical mathematical programming approach. It is seen that from the given steam headers, 8.24 MW of power can be generated annually. This reduces the external power consumption from 12.00 MW to

3.76 MW, Also 12.90 MMBTU/hr of cooling load can be supplied by the steam, utilizing absorption refrigeration system.
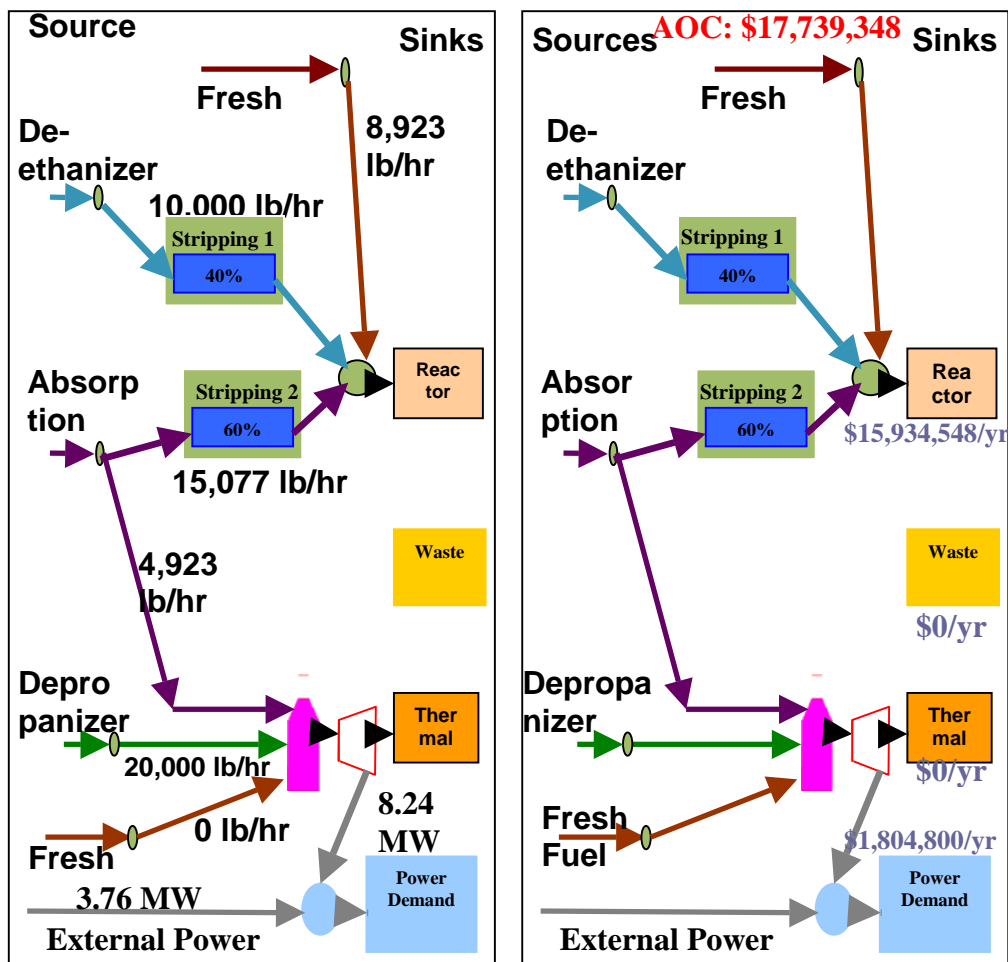
Fig 4.16: Mathematical Programming Approach Solution

4.5.4 UPDATED ANNUAL OPERATING COST

Now the updated annual operating cost will be calculated for the system. We already observed that, mathematical approach resulted in a better solution. Annual operating cost calculation is shown in table 4.12.

Table 4.12: Updated Operating Cost (Mathematical Programming)

Fuel Cost

$$\frac{0 \text{ MMBTU}}{\text{HR}} \left| \frac{\$2.6}{\text{MMBTU}} \right| \frac{8000 \text{ HR}}{\text{YR}} = \$0/\text{YR}$$

Raw Material Cost

$$\frac{8,923 \text{ LB}}{\text{HR}} \left| \frac{\$0.11}{\text{LB}} \right| \frac{8000 \text{ HR}}{\text{YR}} = \$15,934,548/\text{YR}$$

Electricity Cost

$$\frac{3.76 \text{ MW}}{} \left| \frac{\$0.06}{\text{KWHR}} \right| \frac{1000 \text{ KW}}{1 \text{ MW}} \left| \frac{8000 \text{ HR}}{\text{YR}} \right. = \$1,804,800/\text{YR}$$

Waste Treatment Cost

$$\frac{0 \text{ LB}}{\text{HR}} \left| \frac{\$0.0022}{\text{LB}} \right| \frac{8000 \text{ HR}}{\text{YR}} = \$0/\text{YR}$$

Therefore, total annual operating cost for the existing situation is:

$0 + \$15,934,548 + \$1,804,800 + \$0 = \$17,739,348 \approx \$ 18$ Million

4.5.5 ECONOMIC COMPARISON

Now we compare the result from hierarchical mathematical programming approach with the existing situation. The comparison is shown schematically by figure 4.17.
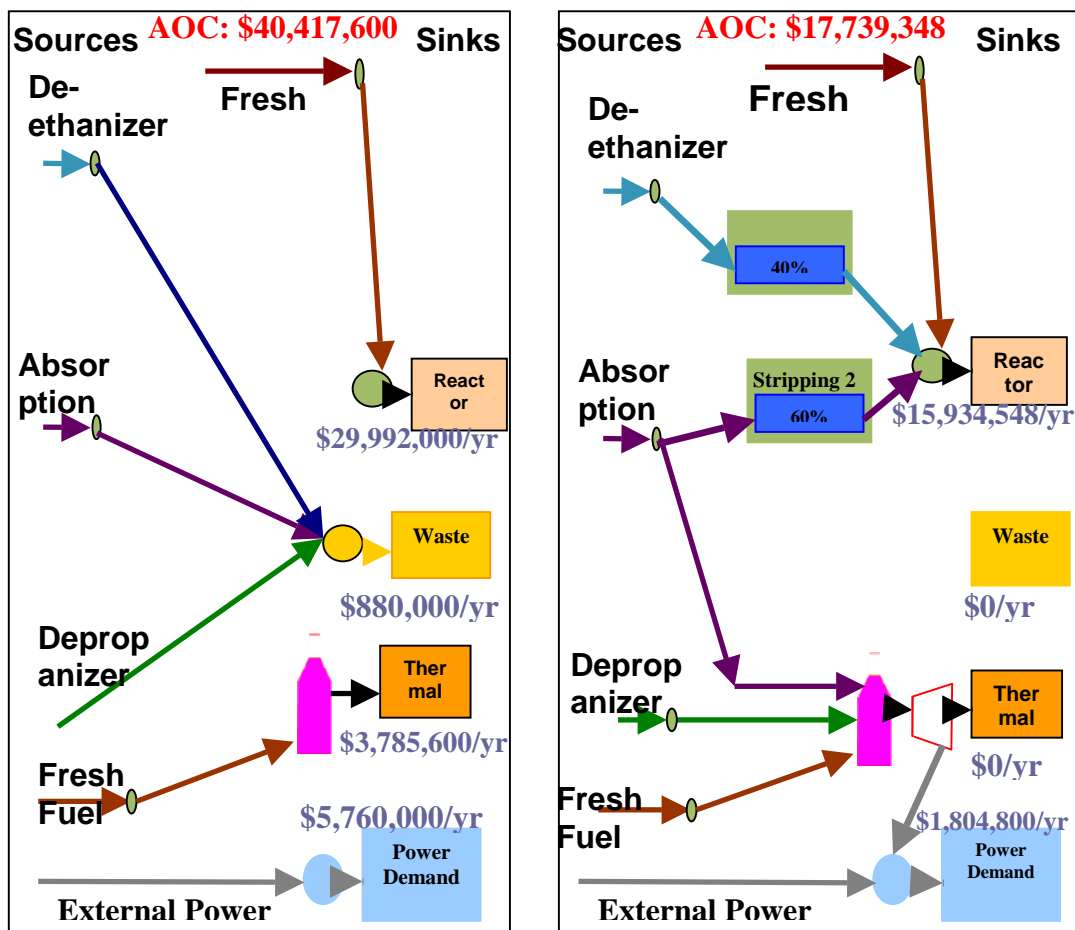
Fig 4.17: Overall Economic Comparison between Existing Situation and Mathematical Programming Solution

Figure 4.17 illustrates the difference in annual operating cost between the existing system and the reconfigured system. Figure 4.18 shows the reconfigured process flow diagram, which results in the annual operating cost saving of approximately $ 22 Million. In the reconfigured diagram it is shown that all the waste stream has been utilized either for material value or thermal value recovery. Part of the absorption column and de-ethanizer top product is utilized for material value recovery by recovering ethylene for the adjacent VAC process, and the rest is utilized as fuel in the

boiler. Also depropanizer bottom product is utilized as fuel in the boiler. So the process does not consume any external fuel. Also power is cogenerated within the system reducing external power consumption.
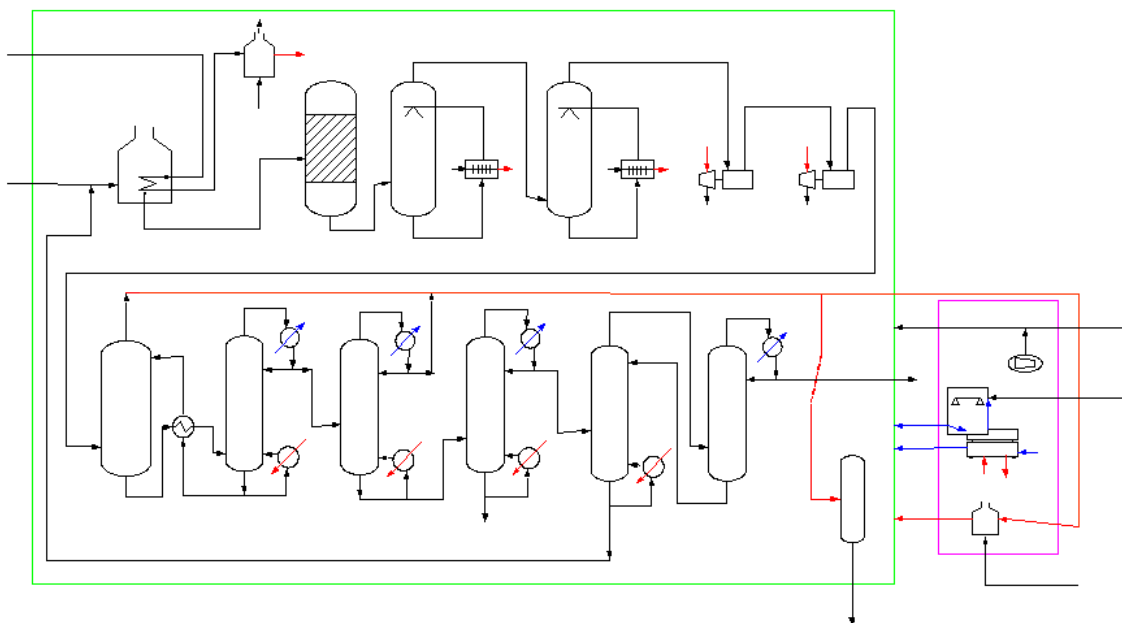
Fig 4.18: Reconfigured Process Flow Diagram (Math Programming)

4.6 CONCLUSIONS

This chapter has introduced a new mathematical formulation, which addresses the optimization of material and energy utilities as well as the reconciliation of heating, cooling, cogeneration, and waste management. Sections of the mathematical formulation were solved simultaneously while other sections were solved hierarchically. A global-optimization technique was developed based on discretization. The simultaneous mass

and heat integration analysis resulted in optimum allocation of process source streams towards material and thermal value recovery and the reduction of waste stream. Material value was recovered after economically comparing with the thermal value recovery. The methodology also identified the fuel substitution opportunities. The whole approach resulted in reduction of consumption of fresh resources like fresh raw material and fresh fuel. Also, it was demonstrated how the result from mass and heat integration analysis affects the steam header balance and eventually the power cogeneration and cooling load management. The developed mathematical approach can provide significant information regarding the complex interaction between material utilities and energy and can result in achievement of multiple interrelated goals for whole process plant.

The case study has described how the hierarchical mathematical programming technique has been utilized in reducing the annual operating cost of a given plant. The strong interaction between the material utilities and energy has also been demonstrated.

CHAPTER V

COMPARISON BETWEEN THE GRAPHICAL AND THE

MATHEMATICAL APPROACHES

Graphical tools provide valuable insights to the designer. Nonetheless, they are limited a "static" set of data and values of variables. Therefore, a decision has to be made on the values of variables to be used and the sequence of calculations. For instance, when waste recycle and interception options were considered, the graphical approach adopted a sequential approach where direct recycle is carried out first followed by interception. The rationale for this approach is that direct recycle is a no or low cost solution that should take precedence over capital-demanding solutions such as the installation of new interception devices. While this rationale may hold true in many cases, it is not always guaranteed to work. In such cases, it is important to simultaneously solve the recycle and the interception problems. The same is true when the material value of the waste is compared with its thermal value. In this cases, the mathematical approach offers a unique advantage by simultaneously addressing theses issues.

The propylene case study was solved in Chapters III and IV using the graphical and the mathematical approaches. Figure 5.1 shows the difference between the results of graphical and the mathematical approaches. While both the approaches have resulted in substantial reduction in total annual operating cost, some of the solution details were different.
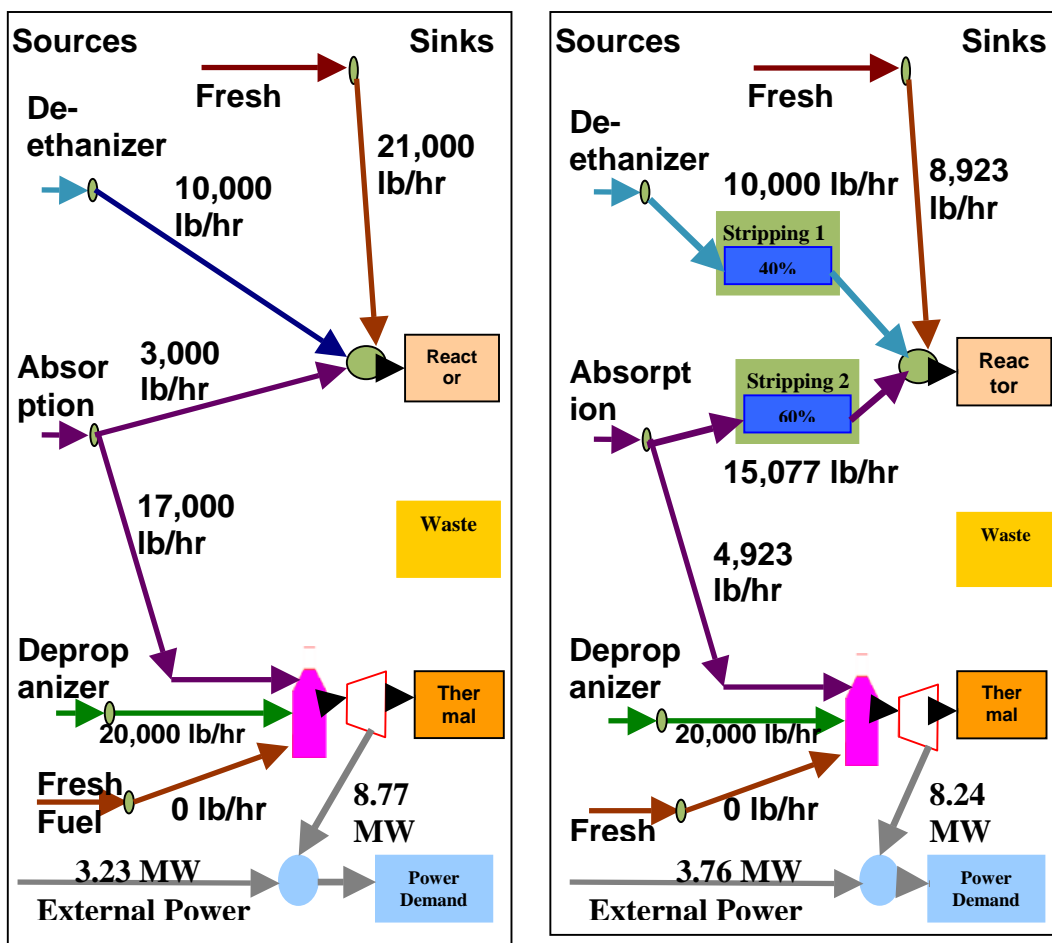
Fig 5.1: Flow Comparison between Graphical and Mathematical Solutions

For instance, both approaches have resulted in the elimination of waste treatment and the elimination of external fuel consumption, but they vary in their fresh resource consumptions such as the consumption of raw materials and the consumption of external power. Also, the graphical technique has resulted in direct recycle without any interception, while the mathematical programming resulted in only interception without direct recycle. Also, the amounts of process steam going for thermal value recovery are

different in the two methods. Figure 5.2 shows the cost difference between the graphical and mathematical programming approaches. It is seen that the mathematical programming approach has generated a solution whose annual operating cost is approximately $2.5 MM less than the solution obtained by the graphical approach.
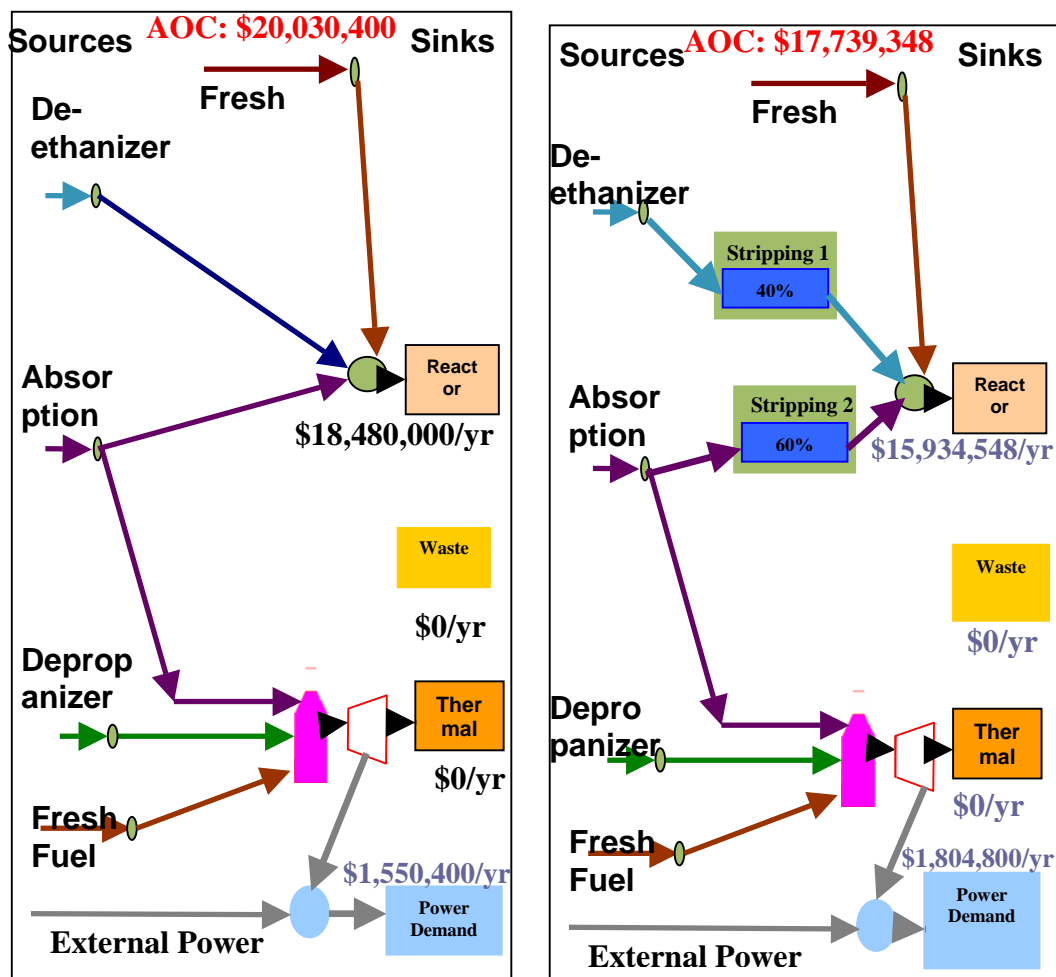
Fig 5.2: Cost Comparison between Graphical and Mathematical Solutions

Here we provide an explanation for the difference in the solution results. In the graphical approach, direct recycle preceded the interception network option because in the graphical approach it was not possible simultaneously analyze direct recycle and interception. This sequential approach has resulted in a direct recycle solution and no interception for the given data set of the problem. Although direct recycle is the cheapest technique of material value recovery from the process sources, from an overall perspective it might not be the optimal choice for a given system. The value of the intercepted stream may exceed the cost of interception thereby making it superior to direct recycle.

For mathematical programming approach, although the cheapest material value recovery technique i.e. direct recycle was not chosen, but from the overall point of view, it resulted in a cheaper raw material consumption cost. It is due to the fact that, by choosing the right and optimal removal efficiency from the given efficiency table, and by choosing optimal interception technology, the approach could reduce fresh raw material consumption drastically. Here the fresh raw material consumption was only 8923 lb/hr compared to 21000 lb/hr for the graphical method. So the total annual raw material consumption cost was only around $ 16.0 million compared to $18.5 million in graphical technique. Again, this is attributed to the ability of simultaneously comparing direct recycle, interception, and fresh consumption and also the ability of comparing the thermal recovery with all these resulted in a very optimal solution.

In further analyzing the results it was found that, the graphical technique used more process sources for thermal value recovery then the mathematical technique

utilized. Again although not optimum, this happened for the graphical technique due to the fact that, it could recover less material value from the process sources, so the process sources were directed towards thermal value recovery. This resulted in more excess steam for the graphical technique than the mathematical technique. Therefore, for the graphical technique the cogeneration potential was higher than that of the mathematical programming technique, which resulted in less external power consumption for the graphical technique than the mathematical programming technique. For the graphical technique, the annual cogeneration potential was 8.77 MW while for the mathematical programming technique the value of cogeneration potential was 8.24 MW. Consequently, for the graphical technique the external power consumption was only 3.23 MW compared to external power consumption of 3.76 MW for the mathematical programming approach. Therefore, the external power consumption cost for the graphical technique is $1.5 Million compared to $1.8 Million for the mathematical approach. So, the graphical technique could offset some cost differences with the mathematical programming technique by the lower external power consumption cost. But this was not enough to surpass the optimum result from the mathematical programming approach based on interception and recycle. In this case study, the raw material cost was higher than the utility cost and also the difference in outside power consumption costs was not large enough. Hence, for the overall solution, the mathematical programming technique provided better optimum solution then the graphical technique.

Another advantage of the mathematical programming approach is the capability of performing sensitivity analysis. The prices of inputs and outputs can varied easily to see the effect on the solution and reconfiguration and the overall optimum result.

Figure 5.3 shows an example of the sensitivity analysis that can be achieved by the mathematical programming approach.
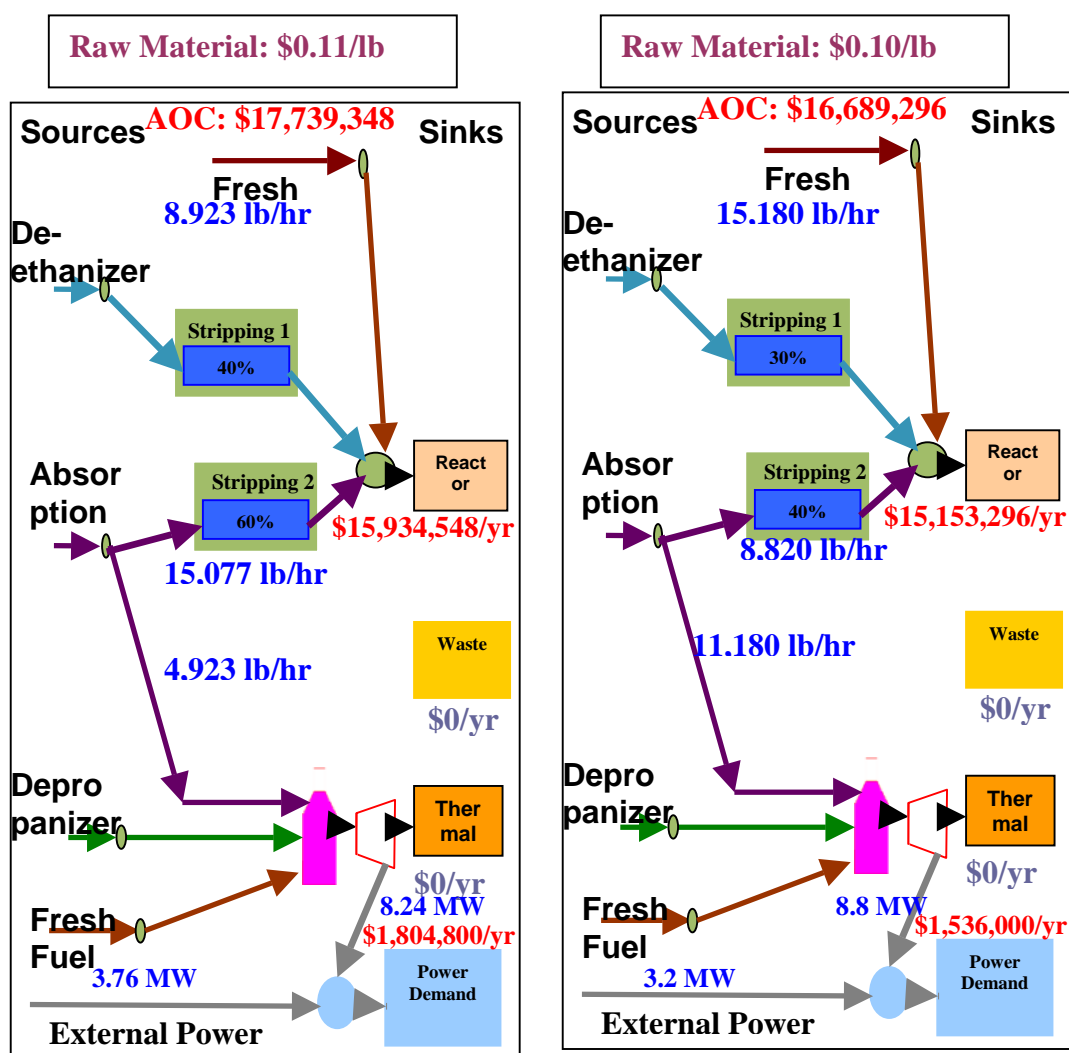
Fig 5.3: Sensitivity Analysis (Mathematical Programming Approach)

Here we see that for only a change of $0.01 in the price of raw material, the mathematical approach generates a new solution. The $0.01 price change not only affects the selection of interception technology with different removal efficiencies, but it also changes the optimum values of process sources that are targeted for material value recovery and thermal value recovery. Meanwhile, the solution increases the purchase of fresh raw material. The total annual raw material cost is decreased to approximately $15 million from approximately $16 million from the previous analysis. The reduction in price of the raw material results in an increase in process sources going for thermal value recovery. This also has a positive effect on the overall solution. Additionally, the decrease in raw material price increases the cogeneration potential (since more process source is directed towards thermal value recovery, it results in additional excess steam, which is utilized for more cogeneration). Hence, the decrease in raw material price also results in decrease in external power consumption price. The result shows the strong interaction between mass and energy integration. This observation highlights the value of this work, which provides a systematic way to integrate and reconcile mass and energy integration.

Finally, it is worth mentioning that notwithstanding the superior results of the mathematical approach to the graphical approach and its ability to readily conduct sensitivity analysis, there are advantages for the graphical approach. The graphical approach provides many insights that are not obtained by the numerical results of the mathematical approach. Also, the graphical approach is less sensitive to the non-convexity of the problem whereas the quality of the mathematical programming solution

is strongly tied to the convexity of the formulation. More engineers are inclined to use visualization techniques than mathematical-programming techniques. Finally, the graphical solution may be used as a starting point for the mathematical approach. As such, both approaches, graphical and mathematical programming, play very important roles and complement each other.

CHAPTER VI

CONCLUSIONS AND FUTURE WORK

6.1 CONCLUSIONS

This work has introduced a systematic methodology for simultaneously targeting and optimizing heating, cooling, power cogeneration, and waste management for any processing facility. This is the first work that systematically addresses the integration of material and energy utilities, power cogeneration, and waste management along with their interactions with the core processing units. Two approaches were developed: graphical and mathematical. In both approaches, a hierarchical procedure was developed to decompose the problem into successive stages that are globally solvable then. The solution fragments were then merged into overall process solutions and targets. Because of the nonlinear, non-convex nature of the developed mathematical-optimization formulation, a problem discretization and reformulation scheme was devised to yield global solutions. Heating and cooling requirements were optimized using a heat-exchange network approach. Waste streams were considered for material recovery, thermal value, and disposal. Economic criteria were used in determining the extent of utilizing the wastes in each alternative. Mass integration was used to determine the optimal direct recycle and interception tasks for the process streams. The results of the heat and mass integration analyses were used to construct steam-header balances. The notion of extractable work was coupled with steam-header balances to determine the power cogeneration targets. Absorption refrigeration cycles were also considered for

cooling duties and their impact on steam balance and cogeneration was studied. The developed methodology is holistic and is expected to aid the process engineers in problem decomposition, targeting, and optimization of problems involving the management of energy and material utilities along with waste management. In particular, the methodology provides process engineers with a useful tool that can help in the following applications:

- Determination of optimal pathways for waste recycle, utilization, and discharge. These pathways include direct reuse for material value, recovery through interception devices, and combustion for thermal value.

- Reconciling the use of steam for various demands of the process (heating, absorptive refrigeration, non-heating, power generation, etc.)

- Identifying the conditions under which power cogeneration becomes economically feasible for the process.

- Selecting optimal types and magnitudes of utilities for the whole process (e.g., steam levels and flows, cooling water versus absorptive refrigeration, etc.)

- Studying the effects of the varying costs of fuel, energy, and feedstocks on the optimal design and operation of the utility systems.

- Conducting a sensitivity analysis on the effect of process and cost data on the process design and operation

A case study was solved using the graphical and the mathematical approaches. The following observations were concluded based on the devised methodology and the results of the case study:

- Substantial improvement can be made over conventional approaches of designing and operating utilities and wastes separately. By integrating utilities, wastes, and core processing units, the graphical and the mathematical approaches developed in this dissertation yielded significant savings over conventional (un-integrated) methods as assessed by the cost metrics (e.g., ~50% reduction in annual operating cost of utilities)

- Conventional wisdom of maximizing the direct recycle/reuse of the waste "recycle your best then treat the rest" is not always true. The rationale for this conventional wisdom is that direct recycle is a no or low cost solution that should take precedence over capital-demanding solutions such as the installation of new interception devices. While this rationale may hold true in some cases, it does not always hold true. Since the mathematical approach developed in this dissertation is the first work to enable the systematic screening and integration of plantwide utilities and wastes, it was shown that depending on the process demands, flowrates and qualities of process sources and wastes, and relative cost data that it in some cases partial recycle coupled with recovery and thermal utilization is the optimal solution.

- On-site power generation is not economically competitive with purchasing external power when steam is generated using an external fuel and power is generated in a condensing turbine without steam usage for heating purposes. On the other hand, power cogeneration may be economically attractive under one or more of the following conditions: (a) Process wastes are burned for steam

generation (b) Combined heat and power is used to distribute the cost over heating and power demands (c) Absorptive refrigeration is needed and power is generated in a non-condensing turbine. In each of these cases, the developed methodology enables a quick determination of whether or not cogeneration is an attractive option.

- Absorption refrigeration is inferior to cooling water at the same operating temperatures if the thermal value of the wastes is less than the material value of the wastes.

- The extent of combined heat and power, steam balance, and waste utilization and discharge is highly sensitive to relative costs of energy, feedstocks, and interception.

- Using the concept of extractable work and using problem reformulation and discretization are effective techniques in getting a global solution to the mathematical-optimization formulation without compromising the accuracy of the original formulation.

Graphical tools, while easy to apply and valuable in providing useful insights to the designer, are limited a "static" set of data and values of variables. Therefore, a decision has to be made on the values of variables to be used and the sequence of calculations. For instance, as mentioned earlier when waste recycle and interception options were considered, the graphical approach adopted a sequential approach where direct recycle is carried out first followed by interception. In such cases, it is important to simultaneously solve the recycle and the interception problems. The same is true

when the material value of the waste is compared with its thermal value. In this cases, the mathematical approach offers a unique advantage by simultaneously addressing theses issues.

6.2 FUTURE WORK

There are several aspects of this dissertation that are recommended for future work. These recommendations include:

- The graphical and mathematical techniques described in the dissertation are hierarchical approaches. In implementing the hierarchical approach, a certain sequence of computations was proposed to decompose the problem into tractable tasks. Future work may be aimed at reducing the hierarchical nature of the approach towards a simultaneous approach. Clearly, this will entail formulating highly nonlinear and nonconvex optimization programs. Substantial effort will have to be exerted to develop global optimization techniques for these programs

- The whole problem was solved assuming steady state. In future work, expansion to unsteady state operation could be explored in each of the decomposed subsection. This future work will involve dynamic modeling of the process as well as scheduling studies.

- The present problem does not directly address the relationship between managing the cooling water and managing the wastewater discharge. Future study can be directed towards linking these two problems by relating energy

integration and waste recycle/reuse with water serving as a common interface

between the two problems.

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APPENDIX A

SIMULTANEOUS MASS AND HEAT INTEGRATION CODE

(GAMS)

$ONTEXT
FOLLOWING PROGRAM DETERMINES THE OPTIMUM UTILIZATION OF
RESOURCES FOR SIMULTANEOUS MASS AND HEAT INTEGRATION
ANALYSIS. THE PROGRAM DETERMINES THE OPTIMUM ALLOCATION
OF COMBUSTIBLE WASTES FOR MATERIAL VALUE RECOVERY, THERMAL
VALUE RECOVERY AND WASTE TREATMENT. ALSO THE PROGRAM
ANALYZES DIFFERENT TECHNOLOGIES FOR MATERIAL VALUE
RECOVERY. THE OPTION BETWEEN THE DIRECT RECYCLE AND
UTILIZATION OF INTERCEPTION NETWORK ARE ALSO COMPARED. THE
MODEL IS SOLVED AS A SUPPLY-DEMAND (SOURCE-SINK) MODEL. THE
SOURCES ARE THE AVAILABLE PROCESS SOURCES, FROM WHICH
MATERIAL, OR THERMAL VALUE CAN BE RECOVERED. ALSO THERE ARE
FRESH SOURCES LIKE RAW MATERIAL FOR MATERIAL SINKS AND FRESH
FUEL FOR THERMAL SINK. AS SINK THE MATERIAL SINK AND THERMAL
SINKS ARE UTILIZED.
$OFFTEXT

* DEFINING AVAILABLE SOURCES
SET   I SOURCES
     /DE_ETH DE_ETHANIZER,
      ABS ABSORPTION/

* DEFINING AVAILABLE SINKS
     J SINKS
      /REACT VA PROCESS REACTOR/

* DEFINING SET OF STRIPPING INTERCEPTOR AVAILABLE FOR EACH
* SOURCE
     K INTERCEPTOR FOR SOURCES
      /STRIP1 INTERCEPTOR,
       STRIP2 INTERCEPTOR/

* DEFINING SET OF ION EXCHANGE INTERCEPTORS AVAILABLE FOR EACH
* SOURCE
     N INTERCEPTORS FOR SOURCES
     /ION1 ION EXCHANGE,

ION2 ION EXCHANGE/

* DEFINING SETS OF EFFICIENCIES FOR THE INTERCEPTORS. THE
* INTERCEPTORS ARE DISCRETIZED IN DIFFERENT REMOVAL EFFICIENCY
* LEVELS
     O EFFICIENCIES
     /EFF1*EFF9/

* DEFINING COST OF INTERCEPTION
     P COST
     /COST1/
* DEFINING A SET OF WASTE TREATMENT FACILITY
     L WASTE TREATMENT
      /WST_TR WASTE TREATMENT/

* DEFINING BOILER FOR THERMAL VALUE RECOVERY FROM THE WASTE
     Q BOILER
     /BL BOILER/

* DEFINING FRESH FUEL CONSUMPTION FOR SUPPLYING THE THERMAL
* DEMAND
     R FRESH FUEL
     /FUEL FRESH FUEL/

* DEFINING FRESH RAW MATERIAL
     M FRESH
      /RAW FRESH RAW MATERIAL/

* DIFFERENT ATTRIBUTE OF THE PROCESS SOURCES
     U ATTRIBUTES TO SOURCES
      /FL_SRC SOURCES FLOW,
       CON_SRC CONC OF SOURCE,
       HEAT_SRC HEATING VALUE OF THE SOURCE IN BTU PER LB/

* DIFFERENT ATTRIBUTES OF THE PROCESS SINKS
     V ATTRIBUTES TO SINKS
      /FL_SNK FLOW OF SINK,
       CON_SNK CONC OF SINK/

* DIFFERENT ATTRIBUTES OF FRESH FUEL
     W ATTRIBUTES OF FUEL
      /COST_F COST BER MILLION BTU,
       HEAT_F HEATING VALUE OF FUEL BTU PER LB/

* DIFFERENT ATTRIBUTES OF RAW MATERIAL
      X ATTRIBUTES OF RAW MATERIAL
      /COST_RAW COST OF RAW MATERIAL,
       COMP_RAW COMPOSITION OF RAW MATERIAL/

* DIFFERENT ATTRIBUTES OF WASTE TREATMENT
      Y ATTRIBUTES FOR WASTE TREATMENT
      /COST_WST COST OF WASTE TREATMENT/

* DIFFERENT ATTRIBUTES OF BOILER
      Z BOILER DEMAND
      /BOIL_D DEMAND OF THE BOILER/

* STREAMS THAT HAVE THERMAL VALUE BUT NO RECOVERABLE
* MATERIAL VALUE
      S DEPROPANIZER BOTTOM
      /D_PROP BOTTOM PRODUCT OF DEPROPANIZER/

* ATTRIBUTES OF THE STREAM WITH RECOVERABLE THERMAL VALUE
* ONLY
      T ATTRIBUTE OF DEPROPANIZER BOTTOM
      /HEAT_DPR HEATING VALUE OF DEPROPANIZER BOTTOM
       PRODUCT,
       FL_DPR AVAILABLE FLOWRATE OF DEPROPANIZER BOTTOM/

$ONTEXT
TABLE FOR SOURCE DATA REPRESENTATION. HERE,
FL_SRC IS THE FLOW OF THE SOURCE (LB/HR)
CON_SRC IMPURITY CONCENTRATION MASS FRACTION IN SOURCE (FR)
HEAT_SRC HEATING VALUE OF SOURCE STREAM (BTU/LB)
$OFFTEXT
TABLE A(I,U) SOURCE DATA

	FL_SRC	CON_SRC	HEAT_SRC
DE_ETH	10000	0.48	1000
ABS	20000	0.65	1500;

$ONTEXT
TABLE FOR SINK DATA REPRESENTATION. HERE,
FL_SNK IS THE FLOW OF THE SINK (LB/HR)
CON_SNK MAXIMIUM INLET IMPURITY CONCENTRATION (MASS
FRACTION) IN SINK (FR)
$OFFTEXT
TABLE B(J,V) SINK DATA

	FL_SNK	CON_SNK
REACT	34000	0.20;

TABLE FOR THE FRESH FUEL DATA. HERE,
 COST_F IS THE COST OF FUEL ($/MMBTU)
 HEAT_F IS THE HEATING VALUE OF FUEL (BTU/LB)
TABLE C(R,W) FUEL DATA

	COST_F	HEAT_F
FUEL	2.6	13400;

TABLE FOR RAW MATERIAL. HERE,
COST_RAW IS THE COST OF RAW MATERIAL ($/LB)
COMP_RAW IS THE IMPURITY COMPOSITION IN RAW MATERIAL (MASS FR.)
TABLE D(M,X) RAW MATERIAL DATA

	COST_RAW	COMP_RAW
RAW	0.11	0.0;

* WASTE TREATMENT COST ($/LB OF WASTE)
TABLE E(L,Y) WASTE TREATMENT DATA

	COST_WST
WST_TR	0.0022;

* BOILER DEMAND (MMBTU)
TABLE F(Q,Z) BOILER DATA

	BOIL_D
BL	182;

TABLE FOR THE DATA OF STREAM WITH ONLY RECOVERABLE THERMAL
VALUE. HEAR THE TABLE FOR DE-PROPANIZER BOTTOM. HEATIN VALUE
IS PROVIDED IN (BTU/LB)
TABLE G(S,T) DEPROPANIZER BOTTOM DATA

	HEAT_DPR	FL_DPR
D_PROP	7000	20000;

TABLE FOR STRIPPING COST. IT IS ASSUMED THAT THE SOURCES WILL NOT BE ALLOWED TO MIX BEFORE THE INTERCEPTION DEVICES. SO COST IS PROVIDED FOR THREE STRIPPERS FOR CORRES- PONDING SOURCES AND AT DIFFERENT REMOVAL EFFICIENCY LEVEL. HERE THE COST IS PRESENTED AS INTERCEPTION COST PER LB OF WASTE REMOVED. ($/LB OF WASTE REMOVED)

$OFFTEXT

TABLE STRIP_COST1(K,O,P) STRIPPING COST DATA

	COST1
STRIP1.EFF1	0.068
STRIP1.EFF2	0.083
STRIP1.EFF3	0.102
STRIP1.EFF4	0.125
STRIP1.EFF5	0.146
STRIP1.EFF6	0.164
STRIP1.EFF7	0.188
STRIP1.EFF8	0.224
STRIP1.EFF9	0.296
STRIP2.EFF1	0.054
STRIP2.EFF2	0.066
STRIP2.EFF3	0.082
STRIP2.EFF4	0.100
STRIP2.EFF5	0.116
STRIP2.EFF6	0.131
STRIP2.EFF7	0.150
STRIP2.EFF8	0.179
STRIP2.EFF9	0.236;

$ONTEXT

TABLE FOR ION EXCHANGE COST. IT IS ASSUMED THAT THE SOURCES WILL NOT BE ALLOWED TO MIX BEFORE THE INTERCEPTION DEVICES. SO COST IS PROVIDED FOR THREE STRIPPERS FOR CORRESPONDING SOURCES AND AT DIFFERENT REMOVAL EFFICIENCY LEVEL. HERE THE COST IS PRESENTED AS INTERCEPTION COST PER LB OF WASTE REMOVED. ($/LB OF WASTE REMOVED)

$OFFTEXT

TABLE ION_COST1(N,O,P) TABLE OF ALL THE DATA

	COST1
ION1.EFF1	0.081
ION1.EFF2	0.099
ION1.EFF3	0.122
ION1.EFF4	0.149
ION1.EFF5	0.179
ION1.EFF6	0.196

ION1.EFF7	0.225
ION1.EFF8	0.268
ION1.EFF9	0.355
ION2.EFF1	0.065
ION2.EFF2	0.079
ION2.EFF3	0.098
ION2.EFF4	0.120
ION2.EFF5	0.140
ION2.EFF6	0.157
ION2.EFF7	0.180
ION2.EFF8	0.215
ION2.EFF9	0.284;

$ONTEXT
DISCRETIZED EFFICIENCIES FOR THE STRIPPING INTERCEPTOR FOR
DIFFERENT SOURCES.
$OFFTEXT
TABLE STRIP_EFF(K,O)

	EFF1	EFF2	EFF3	EFF4	EFF5	EFF6	EFF7	EFF8	EFF9
STRIP1	0.1	0.2	0.3	0.4	0.5	0.6	0.7	0.8	0.9
STRIP2	0.1	0.2	0.3	0.4	0.5	0.6	0.7	0.8	0.9;

$ONTEXT
DISCRETIZED EFFICIENCIES FOR THE ION EXCHANGE INTERCEPTOR FOR
DIFFERENT SOURCES.
$OFFTEXT
TABLE ION_EFF(N,O)

	EFF1	EFF2	EFF3	EFF4	EFF5	EFF6	EFF7	EFF8	EFF9
STRIP1	0.1	0.2	0.3	0.4	0.5	0.6	0.7	0.8	0.9
STRIP2	0.1	0.2	0.3	0.4	0.5	0.6	0.7	0.8	0.9;

* VARIABLE DECLARATION
VARIABLES
    MASS1(I,K) SOURCE TO INTERCEPTION,
    MASS2(M,J) FRESH TO SINK,
    MASS3(I,L) SOURCE TO WASTE,
    MASS4(I,J) SOURCE TO SINK,
    MASS5(K,O,J) INTERCEPTOR TO SINK,
    MASS6(I,N) SOURCE TO INTERCEPTOR,
    MASS7(N,O,J) INTERCEPTOR TO SINK,
    MASS8(I,Q) SOURCE TO BOILER,
    MASS9(R,Q) FRESH FUEL TO BOILER,
    MASS10(S,Q) DEPROPANIZER BOTTOM TO BOILER,
    MASS11(S,L) DEPROPANIZER BOTTOM TO WASTE,

COF COST OF FRESH,
COINT1 COST OF INTERCEPTION STRIP,
COINT2 COST OF INTERCEPTION ION,
CWASTE COST OF WASTE TREATMENT,
AOC ANNUAL OPERATING COST
COFUEL COST OF FUEL;

* NON-NEGATIVE VARIABLES
POSITIVE VARIABLES MASS1, MASS2, MASS3, MASS4, MASS5, MASS6,
MASS7, MASS8, MASS9, MASS10, MASS11, COF, COINT1, COINT2, CWASTE,
COFUEL;

* EQUIATIONS DECLARATION
EQUATIONS
  EQ1 SOURCE CONSTRAINT EQUATION,
  EQ2 SINK CONSTRAINT EQUATION FLOW,
  EQ3 SINK CONSTRAINT EQUATION CONCENTRATION,
  EQ4 FLOW SIMILARITY STRIP,
  EQ5 FLOW SIMILARITY ION,
  EQ6 BOILER DEMAND EQUATION,
  EQ7 SOURCE CONSTRAINT EQUATION FOR THE
        DEPROPANIZER BOTTOM,
  OBJ1 FRESH COST,
  OBJ2 INTERCEPTION COST STRIP,
  OBJ3 WASTE COST,
  OBJ4 INTERCEPTION COST ION,
  OBJ5 COST OF FRESH FUEL,
  OBJ OBJECTIVE FUNCTION;

$ONTEXT
SOURCE CONSTRAINT EQUATION. THE EQUATION SHOWS:
MASS(SOURCE-STRIPPER)+MASS(SOURCE-ION)+MASS(SOURCE-
SINK)+MASS(SOURCE-BOILER)+ MASS(SOURCES-WASTE) <=
SUPPLY(SOURCE);
$OFFTEXT
EQ1(I).. SUM(K$(ORD(K) EQ ORD(I)), MASS1(I,K))+
  SUM(N$(ORD(N) EQ ORD(I)), MASS6(I,N))+SUM(J, MASS4(I,J))+
  SUM(L,MASS3(I,L))+SUM(Q, MASS8(I,Q)) =L= A(I,'FL_SRC');

$ONTEXT
SINK CONSTRAINT EQUATION. THE EQUATION SHOWS:
MASS(STRIPPER-SINK)+MASS(ION-SINK)+MASS(SOURCE-
SINK)+MASS(FRESH-SINK)>= DEMAND(SINK);

$OFFTEXT
EQ2(J).. SUM((K,O), MASS5(K,O,J))+ SUM((N,O), MASS7(N,O,J))+
     SUM(I, MASS4(I,J))+SUM(M, MASS2(M,J)) =G= B(J,'FL_SNK');

$ONTEXT
SINK COCENTRATION CONSTRAINT EQUATION. THE EQUATION SHOWS:
MASS(FRESH-SINK)*CONC(FRESH)+MASS(SOURCE-
SINK)*CONC(SOURCE)+MASS(STRIP-SINK)*(1-
EFFICIENCY(STRIP))*CONC(SOURCE-STRIP)+ MASS(STRIP-ION)*(1-
EFFICIENCY(ION))*CONC(SOURCE-STRIP)
$OFFTEXT
EQ3(J).. SUM(M,MASS2(M,J)*D(M,'COMP_RAW'))+
      SUM(I, MASS4(I,J)*A(I,'CON_SRC')) +
      SUM((K,O), MASS5(K,O,J)*SUM(I$(ORD(I) EQ
      ORD(K)),A(I,'CON_SRC')*(1-STRIP_EFF(K,O))))) +
      SUM((N,O), MASS7(N,O,J)*SUM(I$(ORD(I) EQ
      ORD(N)),A(I,'CON_SRC')*(1-ION_EFF(N,O))))=L=
      B(J,'CON_SNK')*B(J,'FL_SNK');

$ONTEXT
FLOW ASSIGNMENT EQUATION. THIS EQUATION ASSIGNS THE FLOW
FROM A PARTICULAR SOURCE, TO A PARTICULUR INTERCEPTOR TO THE
FLOW FROM THAT INTERCEPTOR TO ALL THE SINKS. FOR STRIP1, THE
EQUATION HERE SAYS:
MASS(SOURCE1-STRIP1) = MASS(STRIP1(EFF1)-
SINK1)+.......+MASS(STRIP1(EFF9)-SINK1)+
MASS(STRIP1(EFF1)-SINK2)+.........+MASS(STRIP1(EFF9)-SINK2)
$OFFTEXT
EQ4(I,K)$(ORD(I) EQ ORD(K)).. SUM((J,O),MASS5(K,O,J)) =E= MASS1(I,K);

$ONTEXT
FLOW ASSIGNMENT EQUATION. THIS EQUATION ASSIGNS THE FLOW
FROM A PARTICULAR SOURCE, TO A PARTICULUR INTERCEPTOR TO THE
FLOW FROM THAT INTERCEPTOR TO ALL THE SINKS.
FOR ION1, THE EQUATION HERE SAYS:
MASS(SOURCE1-ION11) = MASS(ION1(EFF1)-SINK1)+.......+MASS(ION1(EFF9)-
SINK1)+ MASS(ION1(EFF1)-SINK2)+.........+MASS(ION1(EFF9)-SINK2)
$OFFTEXT
EQ5(I,N)$(ORD(I) EQ ORD(N)).. SUM((J,O),MASS7(N,O,J)) =E= MASS6(I,N);

$ONTEXT
BOLIER DEMAND EQUATION. THE EQUATION SHOWS:
MASS(SOURCES-BOILER)*HEAT(SOURCEC)+MASS(FUEL-
BOILER)*HEAT(FUEL) <=DEMAND(BOILER);

$OFFTEXT
EQ6(Q).. SUM(I,MASS8(I,Q)*A(I,'HEAT_SRC'))+
   SUM(R,MASS9(R,Q)*C(R,'HEAT_F'))+
   SUM(S,MASS10(S,Q)*G(S,'HEAT_DPR'))=G=
   F(Q,'BOIL_D')*(10**6);

$ONTEXT
SOURCE CONSTRAINT EQUATION FOR DEPROPANIZER BOTTOM PRODUCT.
THE EQUATION SHOWS:
MASS(DPR-BOILER) + MASS(DPR-WASTE) <= SUPPLY(DPR)
$OFFTEXT
EQ7(S).. SUM(Q, MASS10(S,Q)) +SUM(L, MASS11(S,L)) =L= G(S,'FL_DPR');

* OBJECTIVE FUNCTION
OBJ.. AOC =E= COF + COINT1+ COINT2+ CWASTE +COFUEL;

* DETERMINIMG COST OF FRESH RAW MATERIAL
OBJ1.. COF =E= SUM((M,J), MASS2(M,J)*D(M,'COST_RAW'))*8000;

* DETERMINING COST OF INTERCEPTION BY STEAM STRIPPING
OBJ2.. COINT1 =E= SUM((K,J,O,P), MASS5(K,O,J)*SUM(I$(ORD(I) EQ
   ORD(K)), A(I,'CON_SRC')*STRIP_EFF(K,O)*
   STRIP_COST1(K,O,P)))*8000;

* DETERMINING COST OF WASTE TREATMENT
OBJ3.. CWASTE =E= SUM((I,L), MASS3(I,L)*E(L,'COST_WST'))*8000+
   SUM((S,L), MASS11(S,L)*E(L,'COST_WST'))*8000;

* DETERMINING COST OF INTERCEPTION BY ION-EXCHANGE
OBJ4.. COINT2 =E= SUM((N,J,O,P), MASS7(N,O,J)*
   SUM (I$(ORD(I) EQ ORD(N)), A(I,'CON_SRC')*
   ION_EFF(N,O) * ION_COST1(N,O,P)))*8000;

* DETERMINING COST OF FRESH FUEL
OBJ5.. COFUEL =E= SUM((R,Q),MASS9(R,Q)*
   C(R,'HEAT_F')*C(R,'COST_F'))*8000*(10**(-6));

* MODELING EQUATION
MODEL ANNUALCOST /ALL/;

* SOLVE STATEMENT
SOLVE ANNUALCOST USING LP MINIMIZAING AOC;

* **DISPLAY VALUES OF DIFFERENT VARIABLES**

DISPLAY MASS1.L;
DISPLAY MASS2.L;
DISPLAY MASS3.L;
DISPLAY MASS4.L;
DISPLAY MASS5.L;
DISPLAY MASS6.L;
DISPLAY MASS7.L;
DISPLAY MASS8.L;
DISPLAY MASS9.L;
DISPLAY MASS10.L;
DISPLAY MASS11.L;
DISPLAY COF.L;
DISPLAY COINT1.L;
DISPLAY COINT2.L
DISPLAY CWASTE.L;
DISPLAY COFUEL.L;
DISPLAY AOC.L;

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G E N E R A L   A L G E B R A I C   M O D E L I N G   S Y S T E M
EQUATION LISTING   SOLVE ANNUALCOST USING LP FROM LINE 360

---- EQ1 =L= SOURCE CONSTRAINT EQUATION

EQ1(DE_ETH).. MASS1(DE_ETH,STRIP1) + MASS3(DE_ETH,WST_TR) +
MASS4(DE_ETH,REACT)

   + MASS6(DE_ETH,ION1) + MASS8(DE_ETH,BL) =L= 10000 ; (LHS = 0)

EQ1(ABS).. MASS1(ABS,STRIP2) + MASS3(ABS,WST_TR) +
MASS4(ABS,REACT)

   + MASS6(ABS,ION2) + MASS8(ABS,BL) =L= 20000 ; (LHS = 0)

---- EQ2 =G= SINK CONSTRAINT EQUATION FLOW

EQ2(REACT).. MASS2(RAW,REACT) + MASS4(DE_ETH,REACT) + MASS4(ABS,REACT)

+ MASS5(STRIP1,EFF1,REACT) + MASS5(STRIP1,EFF2,REACT)

+ MASS5(STRIP1,EFF3,REACT) + MASS5(STRIP1,EFF4,REACT)

+ MASS5(STRIP1,EFF5,REACT) + MASS5(STRIP1,EFF6,REACT)

+ MASS5(STRIP1,EFF7,REACT) + MASS5(STRIP1,EFF8,REACT)

+ MASS5(STRIP1,EFF9,REACT) + MASS5(STRIP2,EFF1,REACT)

+ MASS5(STRIP2,EFF2,REACT) + MASS5(STRIP2,EFF3,REACT)

+ MASS5(STRIP2,EFF4,REACT) + MASS5(STRIP2,EFF5,REACT)

+ MASS5(STRIP2,EFF6,REACT) + MASS5(STRIP2,EFF7,REACT)

+ MASS5(STRIP2,EFF8,REACT) + MASS5(STRIP2,EFF9,REACT)

+ MASS7(ION1,EFF1,REACT) + MASS7(ION1,EFF2,REACT) + MASS7(ION1,EFF3,REACT)

+ MASS7(ION1,EFF4,REACT) + MASS7(ION1,EFF5,REACT) + MASS7(ION1,EFF6,REACT)

+ MASS7(ION1,EFF7,REACT) + MASS7(ION1,EFF8,REACT) + MASS7(ION1,EFF9,REACT)

+ MASS7(ION2,EFF1,REACT) + MASS7(ION2,EFF2,REACT) + MASS7(ION2,EFF3,REACT)

+ MASS7(ION2,EFF4,REACT) + MASS7(ION2,EFF5,REACT) + MASS7(ION2,EFF6,REACT)

+ MASS7(ION2,EFF7,REACT) + MASS7(ION2,EFF8,REACT) + MASS7(ION2,EFF9,REACT)
   =G= 34000 ; (LHS = 0, INFES = 34000 ***)

---- EQ3 =L= SINK CONSTRAINT EQUATION CONCENTRATION

EQ3(REACT).. 0.48*MASS4(DE_ETH,REACT) + 0.65*MASS4(ABS,REACT)

+ 0.432*MASS5(STRIP1,EFF1,REACT) + 0.384*MASS5(STRIP1,EFF2,REACT)

+ 0.336*MASS5(STRIP1,EFF3,REACT) + 0.288*MASS5(STRIP1,EFF4,REACT)

+ 0.24*MASS5(STRIP1,EFF5,REACT) + 0.192*MASS5(STRIP1,EFF6,REACT)

+ 0.144*MASS5(STRIP1,EFF7,REACT) + 0.096*MASS5(STRIP1,EFF8,REACT)

+ 0.048*MASS5(STRIP1,EFF9,REACT) + 0.585*MASS5(STRIP2,EFF1,REACT)

+ 0.52*MASS5(STRIP2,EFF2,REACT) + 0.455*MASS5(STRIP2,EFF3,REACT)

+ 0.39*MASS5(STRIP2,EFF4,REACT) + 0.325*MASS5(STRIP2,EFF5,REACT)

+ 0.26*MASS5(STRIP2,EFF6,REACT) + 0.195*MASS5(STRIP2,EFF7,REACT)

+ 0.13*MASS5(STRIP2,EFF8,REACT) + 0.065*MASS5(STRIP2,EFF9,REACT)

+ 0.432*MASS7(ION1,EFF1,REACT) + 0.384*MASS7(ION1,EFF2,REACT)

+ 0.336*MASS7(ION1,EFF3,REACT) + 0.288*MASS7(ION1,EFF4,REACT)

+ 0.24*MASS7(ION1,EFF5,REACT) + 0.192*MASS7(ION1,EFF6,REACT)

+ 0.144*MASS7(ION1,EFF7,REACT) + 0.096*MASS7(ION1,EFF8,REACT)

+ 0.048*MASS7(ION1,EFF9,REACT) + 0.585*MASS7(ION2,EFF1,REACT)

+ 0.52*MASS7(ION2,EFF2,REACT) + 0.455*MASS7(ION2,EFF3,REACT)

+ 0.39*MASS7(ION2,EFF4,REACT) + 0.325*MASS7(ION2,EFF5,REACT)

+ 0.26*MASS7(ION2,EFF6,REACT) + 0.195*MASS7(ION2,EFF7,REACT)

+ 0.13*MASS7(ION2,EFF8,REACT) + 0.065*MASS7(ION2,EFF9,REACT) =L= 6800 ;

(LHS = 0)

---- EQ4 =E= FLOW SIMILARITY STRIP

EQ4(DE_ETH,STRIP1).. - MASS1(DE_ETH,STRIP1) + MASS5(STRIP1,EFF1,REACT)

+ MASS5(STRIP1,EFF2,REACT) + MASS5(STRIP1,EFF3,REACT)

+ MASS5(STRIP1,EFF4,REACT) + MASS5(STRIP1,EFF5,REACT)

+ MASS5(STRIP1,EFF6,REACT) + MASS5(STRIP1,EFF7,REACT)

+ MASS5(STRIP1,EFF8,REACT) + MASS5(STRIP1,EFF9,REACT) =E= 0 ; (LHS = 0)

EQ4(ABS,STRIP2).. - MASS1(ABS,STRIP2) + MASS5(STRIP2,EFF1,REACT)

+ MASS5(STRIP2,EFF2,REACT) + MASS5(STRIP2,EFF3,REACT)

+ MASS5(STRIP2,EFF4,REACT) + MASS5(STRIP2,EFF5,REACT)

+ MASS5(STRIP2,EFF6,REACT) + MASS5(STRIP2,EFF7,REACT)

+ MASS5(STRIP2,EFF8,REACT) + MASS5(STRIP2,EFF9,REACT) =E= 0 ; (LHS = 0)

---- EQ5 =E= FLOW SIMILARITY ION

EQ5(DE_ETH,ION1).. - MASS6(DE_ETH,ION1) + MASS7(ION1,EFF1,REACT)

+ MASS7(ION1,EFF2,REACT) + MASS7(ION1,EFF3,REACT) + MASS7(ION1,EFF4,REACT)

+ MASS7(ION1,EFF5,REACT) + MASS7(ION1,EFF6,REACT) + MASS7(ION1,EFF7,REACT)

+ MASS7(ION1,EFF8,REACT) + MASS7(ION1,EFF9,REACT) =E= 0 ; (LHS = 0)

EQ5(ABS,ION2).. - MASS6(ABS,ION2) + MASS7(ION2,EFF1,REACT)

+ MASS7(ION2,EFF2,REACT) + MASS7(ION2,EFF3,REACT) + MASS7(ION2,EFF4,REACT)

+ MASS7(ION2,EFF5,REACT) + MASS7(ION2,EFF6,REACT) + MASS7(ION2,EFF7,REACT)

+ MASS7(ION2,EFF8,REACT) + MASS7(ION2,EFF9,REACT) =E= 0 ; (LHS = 0)

---- EQ6 =G= BOILER DEMAND EQUATION

EQ6(BL).. 1000*MASS8(DE_ETH,BL) + 1500*MASS8(ABS,BL) + 13400*MASS9(FUEL,BL)

+ 7000*MASS10(D_PROP,BL) =G= 182000000 ; (LHS = 0, INFES = 182000000 *)**

---- EQ7 =L= SOURCE CONSTRAINT EQUATION FOR THE DEPROPANIZER BOTTOM

EQ7(D_PROP).. MASS10(D_PROP,BL) + MASS11(D_PROP,WST_TR) =L= 20000 ; (LHS = 0)

---- OBJ1 =E= FRESH COST

OBJ1.. - 880*MASS2(RAW,REACT) + COF =E= 0 ; (LHS = 0)

---- OBJ2 =E= INTERCEPTION COST STRIP

OBJ2.. - 26.112*MASS5(STRIP1,EFF1,REACT) - 63.744*MASS5(STRIP1,EFF2,REACT)

- 117.504*MASS5(STRIP1,EFF3,REACT) - 192*MASS5(STRIP1,EFF4,REACT)

- 280.32*MASS5(STRIP1,EFF5,REACT) - 377.856*MASS5(STRIP1,EFF6,REACT)

- 505.344*MASS5(STRIP1,EFF7,REACT) - 688.128*MASS5(STRIP1,EFF8,REACT)

- 1022.976*MASS5(STRIP1,EFF9,REACT) - 28.08*MASS5(STRIP2,EFF1,REACT)

- 68.64*MASS5(STRIP2,EFF2,REACT) - 127.92*MASS5(STRIP2,EFF3,REACT)

- 208*MASS5(STRIP2,EFF4,REACT) - 301.6*MASS5(STRIP2,EFF5,REACT)

- 408.72*MASS5(STRIP2,EFF6,REACT) - 546*MASS5(STRIP2,EFF7,REACT)

- 744.64*MASS5(STRIP2,EFF8,REACT) - 1104.48*MASS5(STRIP2,EFF9,REACT)

+ COINT1 =E= 0 ; (LHS = 0)

---- OBJ3 =E= WASTE COST

OBJ3.. - 17.6*MASS3(DE_ETH,WST_TR) - 17.6*MASS3(ABS,WST_TR)

- 17.6*MASS11(D_PROP,WST_TR) + CWASTE =E= 0 ; (LHS = 0)

---- OBJ4 =E= INTERCEPTION COST ION

OBJ4.. - 31.104*MASS7(ION1,EFF1,REACT) - 76.032*MASS7(ION1,EFF2,REACT)

- 140.544*MASS7(ION1,EFF3,REACT) - 228.864*MASS7(ION1,EFF4,REACT)

- 343.68*MASS7(ION1,EFF5,REACT) - 451.584*MASS7(ION1,EFF6,REACT)

- 604.8*MASS7(ION1,EFF7,REACT) - 823.296*MASS7(ION1,EFF8,REACT)

- 1226.88*MASS7(ION1,EFF9,REACT) - 33.8*MASS7(ION2,EFF1,REACT)

- 82.16*MASS7(ION2,EFF2,REACT) - 152.88*MASS7(ION2,EFF3,REACT)

- 249.6*MASS7(ION2,EFF4,REACT) - 364*MASS7(ION2,EFF5,REACT)

- 489.84*MASS7(ION2,EFF6,REACT) - 655.2*MASS7(ION2,EFF7,REACT)

- 894.4*MASS7(ION2,EFF8,REACT) - 1329.12*MASS7(ION2,EFF9,REACT) + COINT2
=E= 0 ; (LHS = 0)

---- OBJ5 =E= COST OF FRESH FUEL

OBJ5.. - 278.72*MASS9(FUEL,BL) + COFUEL =E= 0 ; (LHS = 0)

---- OBJ =E= OBJECTIVE FUNCTION

OBJ.. - COF - COINT1 - COINT2 - CWASTE + AOC - COFUEL =E= 0 ; (LHS = 0)

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G E N E R A L   A L G E B R A I C   M O D E L I N G   S Y S T E M
COLUMN LISTING   SOLVE ANNUALCOST USING LP FROM LINE 360

---- MASS1 SOURCE TO INTERCEPTION

MASS1(DE_ETH,STRIP1)
        (.LO, .L, .UP = 0, 0, +INF)
     1   EQ1(DE_ETH)
    -1   EQ4(DE_ETH,STRIP1)

MASS1(ABS,STRIP2)
        (.LO, .L, .UP = 0, 0, +INF)
     1   EQ1(ABS)
    -1   EQ4(ABS,STRIP2)

---- MASS2 FRESH TO SINK

MASS2(RAW,REACT)
        (.LO, .L, .UP = 0, 0, +INF)
     1   EQ2(REACT)
  -880   OBJ1

---- MASS3 SOURCE TO WASTE

MASS3(DE_ETH,WST_TR)
        (.LO, .L, .UP = 0, 0, +INF)
     1   EQ1(DE_ETH)
  -17.6   OBJ3

MASS3(ABS,WST_TR)
        (.LO, .L, .UP = 0, 0, +INF)
     1   EQ1(ABS)
  -17.6   OBJ3

---- MASS4 SOURCE TO SINK

MASS4(DE_ETH,REACT)
        (.LO, .L, .UP = 0, 0, +INF)
     1   EQ1(DE_ETH)

    1   EQ2(REACT)
    0.48   EQ3(REACT)

MASS4(ABS,REACT)
       (.LO, .L, .UP = 0, 0, +INF)
    1   EQ1(ABS)
    1   EQ2(REACT)
    0.65   EQ3(REACT)

---- MASS5 INTERCEPTOR TO SINK

MASS5(STRIP1,EFF1,REACT)
       (.LO, .L, .UP = 0, 0, +INF)
    1   EQ2(REACT)
   0.432   EQ3(REACT)
    1   EQ4(DE_ETH,STRIP1)
  -26.112   OBJ2

MASS5(STRIP1,EFF2,REACT)
       (.LO, .L, .UP = 0, 0, +INF)
    1   EQ2(REACT)
   0.384   EQ3(REACT)
    1   EQ4(DE_ETH,STRIP1)
  -63.744   OBJ2

MASS5(STRIP1,EFF3,REACT)
       (.LO, .L, .UP = 0, 0, +INF)
    1   EQ2(REACT)
   0.336   EQ3(REACT)
    1   EQ4(DE_ETH,STRIP1)
  -117.504   OBJ2

REMAINING 15 ENTRIES SKIPPED

---- MASS6 SOURCE TO INTERCEPTOR

MASS6(DE_ETH,ION1)
       (.LO, .L, .UP = 0, 0, +INF)
    1   EQ1(DE_ETH)
   -1   EQ5(DE_ETH,ION1)

MASS6(ABS,ION2)
       (.LO, .L, .UP = 0, 0, +INF)

        1   EQ1(ABS)
       -1   EQ5(ABS,ION2)

---- MASS7 INTERCEPTOR TO SINK

MASS7(ION1,EFF1,REACT)
        (.LO, .L, .UP = 0, 0, +INF)
     1   EQ2(REACT)
    0.432   EQ3(REACT)
     1   EQ5(DE_ETH,ION1)
   -31.104 OBJ4

MASS7(ION1,EFF2,REACT)
        (.LO, .L, .UP = 0, 0, +INF)
     1   EQ2(REACT)
    0.384   EQ3(REACT)
     1   EQ5(DE_ETH,ION1)
   -76.032 OBJ4

MASS7(ION1,EFF3,REACT)
        (.LO, .L, .UP = 0, 0, +INF)
     1   EQ2(REACT)
    0.336   EQ3(REACT)
     1   EQ5(DE_ETH,ION1)
   -140.544 OBJ4

REMAINING 15 ENTRIES SKIPPED

---- MASS8 SOURCE TO BOILER

MASS8(DE_ETH,BL)
        (.LO, .L, .UP = 0, 0, +INF)
     1   EQ1(DE_ETH)
   1000   EQ6(BL)

MASS8(ABS,BL)
        (.LO, .L, .UP = 0, 0, +INF)
     1   EQ1(ABS)
   1500   EQ6(BL)

---- MASS9 FRESH FUEL TO BOILER

MASS9(FUEL,BL)

      (.LO, .L, .UP = 0, 0, +INF)

  13400   **EQ6(BL)**

  -278.72 **OBJ5**

---- MASS10 DEPROPANIZER BOTTOM TO BOILER

MASS10(D_PROP,BL)

      (.LO, .L, .UP = 0, 0, +INF)

  7000   **EQ6(BL)**

    1   **EQ7(D_PROP)**

---- MASS11 DEPROPANIZER BOTTOM TO WASTE

MASS11(D_PROP,WST_TR)

      (.LO, .L, .UP = 0, 0, +INF)

    1   **EQ7(D_PROP)**

  -17.6   **OBJ3**

---- COF COST OF FRESH

COF

      (.LO, .L, .UP = 0, 0, +INF)

    1   **OBJ1**

    -1   **OBJ**

---- COINT1 COST OF INTERCEPTION STRIP

COINT1

      (.LO, .L, .UP = 0, 0, +INF)

    1   **OBJ2**

    -1   **OBJ**

---- COINT2 COST OF INTERCEPTION ION

COINT2

      (.LO, .L, .UP = 0, 0, +INF)

    1   **OBJ4**

    -1   **OBJ**

---- CWASTE COST OF WASTE TREATMENT

CWASTE
        (.LO, .L, .UP = 0, 0, +INF)
     1   OBJ3
     -1   OBJ

---- AOC ANNUAL OPERATING COST

AOC
        (.LO, .L, .UP = -INF, 0, +INF)
     1   OBJ

---- COFUEL COST OF FUEL

COFUEL
        (.LO, .L, .UP = 0, 0, +INF)
     1   OBJ5
     -1   OBJ

GAMS REV 134 WINDOWS NT/95/98   07/17/05 12:21:54 PAGE 4
G E N E R A L   A L G E B R A I C   M O D E L I N G   S Y S T E M
MODEL STATISTICS   SOLVE ANNUALCOST USING LP FROM LINE 360

MODEL STATISTICS

BLOCKS OF EQUATIONS   13   SINGLE EQUATIONS   16
BLOCKS OF VARIABLES   17   SINGLE VARIABLES   56
NON ZERO ELEMENTS   185

GENERATION TIME   =   0.090 SECONDS   1.7 MB   WIN210-134

EXECUTION TIME   =   0.110 SECONDS   1.7 MB   WIN210-134

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G E N E R A L   A L G E B R A I C   M O D E L I N G   S Y S T E M
SOLUTION REPORT   SOLVE ANNUALCOST USING LP FROM LINE 360

S O L V E   S U M M A R Y

     MODEL   ANNUALCOST   OBJECTIVE AOC
     TYPE   LP   DIRECTION MINIMIZE
     SOLVER CPLEX   FROM LINE 360

**** SOLVER STATUS   1 NORMAL COMPLETION
**** MODEL STATUS   1 OPTIMAL
**** OBJECTIVE VALUE   16654547.6923

 RESOURCE USAGE, LIMIT   0.110   1000.000
 ITERATION COUNT, LIMIT   14   10000

GAMS/CPLEX   MAY 15, 2003 WIN.CP.CP 21.0 023.025.041.VIS FOR CPLEX
8.1
CPLEX 8.1.0, GAMS LINK 23

OPTIMAL SOLUTION FOUND.
OBJECTIVE :   16654547.692308

---- EQU EQ1 SOURCE CONSTRAINT EQUATION

      LOWER   LEVEL   UPPER   MARGINAL

DE_ETH   -INF 10000.000 10000.000 -200.527
ABS   -INF 20000.000 20000.000   -31.200

---- EQU EQ2 SINK CONSTRAINT EQUATION FLOW

      LOWER   LEVEL   UPPER   MARGINAL

REACT 34000.000 34000.000   +INF   880.000

---- EQU EQ3 SINK CONSTRAINT EQUATION CONCENTRATION

      LOWER   LEVEL   UPPER   MARGINAL

REACT   -INF   6800.000 6800.000 -1692.615

---- EQU EQ4 FLOW SIMILARITY STRIP

	LOWER	LEVEL	UPPER	MARGINAL
DE_ETH.STRIP1	.	.	.	-200.527
ABS .STRIP2	.	.	.	-31.200

---- EQU EQ5 FLOW SIMILARITY ION

	LOWER	LEVEL	UPPER	MARGINAL
DE_ETH.ION1	.	.	.	-200.527
ABS .ION2	.	.	.	-31.200

---- EQU EQ6 BOILER DEMAND EQUATION

	LOWER	LEVEL	UPPER	MARGINAL
BL	1.8200E+8	1.8200E+8	+INF	0.021

---- EQU EQ7 SOURCE CONSTRAINT EQUATION FOR THE DEPROPANIZER BOTTOM

	LOWER	LEVEL	UPPER	MARGINAL
D_PROP	-INF	20000.000	20000.000	-145.600

	LOWER	LEVEL	UPPER	MARGINAL
---- EQU OBJ1	.	.	.	1.000
---- EQU OBJ2	.	.	.	1.000
---- EQU OBJ3	.	.	.	1.000
---- EQU OBJ4	.	.	.	1.000
---- EQU OBJ5	.	.	.	1.000
---- EQU OBJ	.	.	.	1.000

OBJ1 FRESH COST
OBJ2 INTERCEPTION COST STRIP
OBJ3 WASTE COST
OBJ4 INTERCEPTION COST ION
OBJ5 COST OF FRESH FUEL
OBJ OBJECTIVE FUNCTION

---- VAR MASS1 SOURCE TO INTERCEPTION

	LOWER	LEVEL	UPPER	MARGINAL
DE_ETH.STRIP1	.	10000.000	+INF	.
ABS .STRIP2	.	15076.923	+INF	.

---- VAR MASS2 FRESH TO SINK

	LOWER	LEVEL	UPPER	MARGINAL
RAW.REACT	.	8923.077	+INF	.

---- VAR MASS3 SOURCE TO WASTE

	LOWER	LEVEL	UPPER	MARGINAL
DE_ETH.WST_TR	.	.	+INF	218.127
ABS .WST_TR	.	.	+INF	48.800

---- VAR MASS4 SOURCE TO SINK

	LOWER	LEVEL	UPPER	MARGINAL
DE_ETH.REACT	.	.	+INF	132.982
ABS .REACT	.	.	+INF	251.400

---- VAR MASS5 INTERCEPTOR TO SINK

	LOWER	LEVEL	UPPER	MARGINAL
STRIP1.EFF1.REACT	.	.	+INF	77.849
STRIP1.EFF2.REACT	.	.	+INF	34.235
STRIP1.EFF3.REACT	.	.	+INF	6.750
STRIP1.EFF4.REACT	.	10000.000	+INF	.
STRIP1.EFF5.REACT	.	.	+INF	7.074
STRIP1.EFF6.REACT	.	.	+INF	23.365
STRIP1.EFF7.REACT	.	.	+INF	69.607
STRIP1.EFF8.REACT	.	.	+INF	171.146
STRIP1.EFF9.REACT	.	.	+INF	424.748
STRIP2.EFF1.REACT	.	.	+INF	169.460
STRIP2.EFF2.REACT	.	.	+INF	100.000
STRIP2.EFF3.REACT	.	.	+INF	49.260
STRIP2.EFF4.REACT	.	.	+INF	19.320

```
STRIP2.EFF5.REACT    .    .    +INF    2.900
STRIP2.EFF6.REACT    .  15076.923  +INF    .
STRIP2.EFF7.REACT    .    .    +INF   27.260
STRIP2.EFF8.REACT    .    .    +INF  115.880
STRIP2.EFF9.REACT    .    .    +INF  365.700
```

---- VAR MASS6 SOURCE TO INTERCEPTOR

```
          LOWER   LEVEL   UPPER   MARGINAL

DE_ETH.ION1    .    .    +INF    .
ABS  .ION2    .    .    +INF    .
```

---- VAR MASS7 INTERCEPTOR TO SINK

```
          LOWER   LEVEL   UPPER   MARGINAL

ION1.EFF1.REACT    .    .    +INF   82.841
ION1.EFF2.REACT    .    .    +INF   46.523
ION1.EFF3.REACT    .    .    +INF   29.790
ION1.EFF4.REACT    .    .    +INF   36.864
ION1.EFF5.REACT    .    .    +INF   70.434
ION1.EFF6.REACT    .    .    +INF   97.093
ION1.EFF7.REACT    .    .    +INF  169.063
ION1.EFF8.REACT    .    .    +INF  306.314
ION1.EFF9.REACT    .    .    +INF  628.652
ION2.EFF1.REACT    .    .    +INF  175.180
ION2.EFF2.REACT    .    .    +INF  113.520
ION2.EFF3.REACT    .    .    +INF   74.220
ION2.EFF4.REACT    .    .    +INF   60.920
ION2.EFF5.REACT    .    .    +INF   65.300
ION2.EFF6.REACT    .    .    +INF   81.120
ION2.EFF7.REACT    .    .    +INF  136.460
ION2.EFF8.REACT    .    .    +INF  265.640
ION2.EFF9.REACT    .    .    +INF  590.340
```

---- VAR MASS8 SOURCE TO BOILER

```
          LOWER   LEVEL   UPPER   MARGINAL

DE_ETH.BL    .    .    +INF  179.727
ABS  .BL    .  4923.077  +INF    .
```

---- VAR MASS9 FRESH FUEL TO BOILER

```
                LOWER    LEVEL    UPPER   MARGINAL

FUEL.BL     .    2583.238    +INF      .
```

---- VAR MASS10 DEPROPANIZER BOTTOM TO BOILER

```
                  LOWER    LEVEL    UPPER   MARGINAL

D_PROP.BL     .    20000.000    +INF      .
```

---- VAR MASS11 DEPROPANIZER BOTTOM TO WASTE

```
                    LOWER    LEVEL    UPPER   MARGINAL

D_PROP.WST_TR    .      .        +INF   163.200
```

```
                      LOWER    LEVEL    UPPER   MARGINAL

---- VAR COF          .    7.8523E+6    +INF      .
---- VAR COINT1       .    8.0822E+6    +INF      .
---- VAR COINT2       .       .         +INF      .
---- VAR CWASTE       .       .         +INF      .
---- VAR AOC        -INF   1.6655E+7    +INF      .
---- VAR COFUEL       .    7.2000E+5    +INF      .
```

```
 COF  COST OF FRESH
 COINT1  COST OF INTERCEPTION STRIP
 COINT2  COST OF INTERCEPTION ION
 CWASTE  COST OF WASTE TREATMENT
 AOC  ANNUAL OPERATING COST
 COFUEL  COST OF FUEL
```

```
**** REPORT SUMMARY :     0   NONOPT
                 0 INFEASIBLE
                 0  UNBOUNDED
```

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G E N E R A L   A L G E B R A I C   M O D E L I N G   S Y S T E M
E X E C U T I O N

----   362 VARIABLE MASS1.L SOURCE TO INTERCEPTION

        STRIP1   STRIP2

DE_ETH   10000.000
ABS   15076.923

----   363 VARIABLE MASS2.L FRESH TO SINK

        REACT

RAW   8923.077

----   364 VARIABLE MASS3.L SOURCE TO WASTE

        (ALL   0.000)

----   365 VARIABLE MASS4.L SOURCE TO SINK

        (ALL   0.000)

----   366 VARIABLE MASS5.L INTERCEPTOR TO SINK

        REACT

STRIP1.EFF4   10000.000
STRIP2.EFF6   15076.923

----   367 VARIABLE MASS6.L SOURCE TO INTERCEPTOR

        (ALL   0.000)

----   368 VARIABLE MASS7.L INTERCEPTOR TO SINK

(ALL   0.000)

---- 369 VARIABLE MASS8.L SOURCE TO BOILER

BL

ABS   4923.077

---- 370 VARIABLE MASS9.L FRESH FUEL TO BOILER

BL

FUEL   2583.238

---- 371 VARIABLE MASS10.L DEPROPANIZER BOTTOM TO BOILER

BL

D_PROP 20000.000

---- 372 VARIABLE MASS11.L DEPROPANIZER BOTTOM TO WASTE

(ALL   0.000)

---- 373 VARIABLE COF.L   = 7852307.692 COST OF FRESH

---- 374 VARIABLE COINT1.L   = 8082240.000 COST OF
INTERCEPTION

STRIP

---- 375 VARIABLE COINT2.L   =   0.000 COST OF INTERCEPTION

ION

---- 376 VARIABLE CWASTE.L   =   0.000 COST OF WASTE
TREATME

NT

---- 377 VARIABLE COFUEL.L   = 720000.000 COST OF FUEL

----   378 VARIABLE AOC.L   = 1.665455E+7 ANNUAL OPERATING COST

EXECUTION TIME   =   0.000 SECONDS   1.5 MB   WIN210-134

USER: ADVANCED GAMS CLASS   G040108:1749AJ-WIN
    JAN., 2004 COLLEGE STATION, TEXAS   DC4545

**** FILE SUMMARY

INPUT   C:\DOCUMENTS AND SETTINGS\RUBAYAT MAHMUD\DESKTOP\THESIS FILES- AUG 14
        TH\THESIS-M-H-RAW11-AUG10.GMS
OUTPUT   C:\WINDOWS\GAMSDIR\THESIS-M-H-RAW11-AUG10.LST

APPENDIX B

COGENERATION POTENTIAL CODE (GAMS)

$TITLE **COGENERATION POTENTIAL**
$ONTEXT
FOLLOWING PROGRAM CALCULATES OPTIMUM COGENERATION
POTENTIAL FROM A GIVEN SET OF STEAM HEADERS. INFORMATIONS
AVAILABLE FOR STEAM HEADERS ARE: STEAM HEADER PRESSURE,
STEAM HEADER TEMPERATURE, PROCESS STEAM DEMAND AT ANY GIVEN
HEADER AND ALSO AVAILABLE STEAM SUPPLY AT ANY GIVEN HEADER.
PROCEDURES FOLLOWED FOR THE PROGRAM ARE AS FOLLOWED:

- FROM THE PRESSURE AND TEMPERATURE DATA AT HEADER LEVEL,
ENTHALPY OF THE STEAM AT THAT HEADER CONDITION IS CALCULATED.
FOR THE ENTHALPY CALCULATION, EQUATIONS AND CORRELATIONS
DEVELOPED BY "THOMAS F. IRVINE, Jr." AND "PETER E. LILEY" HAVE BEEN
UTILIZED. REFERENCE FOR THE ENTHALPY CALCULATION: "STEAM AND
GAS TABLES WITH COMPUTER EQUATION", THOMAS F. IRVINE, Jr. AND
PETER E. LILEY, ACADEMIC PRESS, INC., 1984.

- FROM THE SUPPLY AND DEMAND INFORMATION THE PROGRAM
DETERMINES THE SUPPLY HEADER AND DEFICIT HEADER.

- FROM THE SUPPLY AND DEMAND DATA AND STEAM ENTHALPY AT A
GIVEN HEADER, MASS FLOW RATE OF EXCESS OR DEFICIT STEAM AT ANY
GIVEN HEADER IS CALCULATED BY FOLLOWING EQUATION:
     STEAM MASS FLOW = (SUPPLY-DEMAND)/ENTHALPY

- TURBINE EFFICIENCY HAS BEEN ASSUMED TO BE EQUAL TO 0.7

- FROM ALL THESE DATA THE EXTRACTABLE ENERGY METHODOLOGY IS
UTILIZED TO CALCULATE THE EXTRACTABLE POWER FROM ANY TWO
HEADER COMBINATION. EXTRACTABLE POWER IS DEFINED BY:
EXTRACTABLE POWER   = STEAM MASS FLOW * TURBINE
EFFICINECY*ENTHALPY DIFFERENCE (BETWEEN STEAM HEADER)

- THEN OPTIMIZATION FORMULATION IS UTILIZED TO DETERMINE THE
OPTIMUM COGENERATION POTENTIAL FROM THE GIVEN SET OF STEAM
HEADERS.
$OFFTEXT

* DEFINING THE AVAILABLE STEAM HEADERS
SET I STEAM HEADERS
  /HP HIGH PRESSURE,
   MP MEDIUM PRESSURE,
   LP LOW PRESSURE/

* ATTRIBUTES FOR EACH STEAM HEADERS
  J ATTRIBUTES OF THE HEADRS
   /PRES PRESSURE OF THE HEADER (PSIA),
   TEMP TEMPERATURE OF THE HEADER (DEG F),
   SUPPLY SUPPLY FOR THE HEADER (MMBTU PER HOUR),
   DEMAND DEMAND OF THE HEADER (MMBTU PER HOUR)/

* CONDENSING TURBINE
  L CONDENSING TURBINE /COND/

* DEFINING SURPLUS AND DEFICIT HEADERS AS SUBSETS OF STEAM *
HEADER SET. MEMEBERS FOR SURPLUS AND DEFICIT HEADERS ARE *
ASSIGEND DYNAMICALLY.
  SU(I) SURPLUS HEADER
  DF(I) DEFICIT HEADER

$ONTEXT
STEAM HEADER INFORMATIONS. DATA INPUT IN TABLE FORMAT. HERE
UNITS ARE: PRESSURE: PSIA (PRESSURE OF THE HEADER)
TEMPERATURE: DEGREE FARENHIET (TEMEPRATURE OF THE HEADER)
SUPPLY: MMBTU/HR (STEAM SUPPLY AT GIVEN HEADER)
DEMAND: MMBTU/HR (PROCESS STEAM DEMAND AT GIVEN HEADER)
$OFFTEXT
TABLE D (I,J) INLET TEMPERATURE AND PRESSUR DATA TABLE

	PRES	**TEMP**	**SUPPLY**	**DEMAND**
HP	**600**	**800**	**238**	**0**
MP	**130**	**350**	**10**	**25**
LP	**40**	**270**	**20**	**192;**

* CONDENSING TURBINE ENTHALPY
PARAMETER HCD(L) ENTHALPY OF CONDENSER EXHAUST /COND 1000/;

$ONTEXT
ASSIGNING MEMEBERS FOR THE SURPLUS SUBSET. THE CODE ASSIGNS
SUCH HEADERS TO THE SURPLUS SUBSET FOR WHOM SUPPLY IS GREATER
THEN THE DEMAND
$OFFTEXT
SU(I) = YES$(D(I,'SUPPLY') GT D(I,'DEMAND'));

$ONTEXT
ASSIGNING MEMEBERS FOR THE DEFICIT SUBSET. THE CODE ASSIGNS
SUCH HEADERS TO THE DEFICIT SUBSET FOR WHOM SUPPLY IS LESS THEN
THE DEMAND
$OFFTEXT
DF(I) = YES$(D(I,'SUPPLY') LT D(I,'DEMAND'));

$ONTEXT
ENTHALPY CALCULATION: "STEAM AND GAS TABLES WITH COMPUTER
EQUATIONS", THOMAS F. IRVINE, JR., PETER E. LILEY, ACADEMIC PRESS
INC., 1984.
$OFFTEXT

* PARAMETERS ARE DEFINED ON THE HEADER SET
PARAMETER A0(I) PARAMETER FOR ENTHALPY CALCULATION,
A1(I) PARAMETER FOR ENTHALPY CALCULATION,
A2(I) PARAMETER FOR ENTHALPY CALCULATION,
A3(I) PARAMETER FOR ENTHALPY CALCULATION,
TS(I) SATURATION TEMPERATURE (DEG K),
P(I) HEADER PRESSURE (MPA),
T(I) HEADER TEMPERATURE (DEG K),
ENTH(I) ENTHALPY AT HEADER CONDITION (KJ PER KG),
H(I) ENTHALPY AT HEADER CONDITION (BTU PER LB),
M(I) STEAM MASS FLOW AT HEADER (AVAILABLE OR
        REQUIRED)(LB PER HOUR);

$ONTEXT
PRESSURE FOR EACH HEADER. PRESSURE IN THE DATA TABLE IS GIVEN IN
THE PSIA UNIT. FOR THE FORMULATION OF IRVINE AND LILEY, THE
PRESSURE IS DEFINED IN THE MPA UNIT. SO THE NUMBER 0.00689476 IS
UTILIZED AS A CONVERSION FACTOR TO CONVERT
PSIA TO MPA. HERE, PRESSURE IN MPA = PRESSURE IN PSIA*0.00689476
$OFFTEXT
P(I) = D(I,'PRES')*0.00689476;

$ONTEXT
TEMEPRATURE FOR EACH HEADER. TEMPERATURE IN THE DATA TABLE IS
GIVEN IN DEG F. FOR THE FORMULATION OF IRVINE AND LILEY, THE
TEMPERATURE IS DEFINED IN DEG K. SO CONVERSION FORMULA IS USED
TO CONVERT DEG F TO DEG K. HERE,
TEMP IN DEG K = (TEMP IN DEG F + 459.67)/1.8
$OFFTEXT
T(I) = (D(I,'TEMP')+459.67)/1.8;

$ONTEXT
CALCULATION OF SATURATION TEMPERATURE (IRVINE & LILEY: PP21).
SCALAR VALUES ARE TAKEN FROM THE CORRELATIONS DEFINED BY
IRVINE & LILEY SCALAR AT, BT, CT SCALARS FOR CALCUALTING
STAURATION TEMPERATURE;
$OFFTEXT
AT = 0.426776E+02;
BT = -0.389270E+04;
CT = -0.948654E+01;

TS(I) = AT + (BT/(LOG(P(I))+CT));

*CALCULATION OF ENTHALPY (IRVINE & LILEY: PP51)
* SCALAR VALUES ARE TAKEN FROM THE CORRELATIONS DEFINED BY *
IRVINE & LILEY
SCALAR B11, B12, B13, B21, B22, B23, B31, B32, B33, B41, B42, B43,
   B44, B45, MM SCALARS FOR CALCULATING ENTHALPY;

B11 = 2.04121E+03;
B12 = -4.040021E+01;
B13 = -4.8095E-01;
B21 = 1.610693;
B22 = 5.472051E-02;
B23 = 7.517537E-04;
B31 = 3.383117E-04;
B32 = -1.975736E-05;
B33 = -2.87409E-07;
B41 = 1.70782E+03;
B42 = -1.699419E+01;
B43 = 6.2746295E-02;
B44 = -1.0284259E-04;
B45 = 6.4561298E-08;

MM = 4.5E+01;

A0(I) = B11+B12*P(I)+B13*(P(I)2);**
A1(I) = B21+B22*P(I)+B23*(P(I)2);**
A2(I) = B31+B32*P(I)+B33*(P(I)2);**
A3(I) = B41+B42*TS(I)+B43*(TS(I)2)+B44*(TS(I)**3)+B45*(TS(I)**4);**

* ENTHALPY (KJ / KG)
ENTH(I) = A0(I)*(T(I)0)+A1(I)*(T(I)**1)+A2(I)*(T(I)**2)-A3(I)***
   EXP((TS(I)-T(I))/MM);

$ONTEXT
ENTHALPY (BTU / LB). HERE 0.429923 IS CONVERSION FACTOR FOR
CONVERTING (KJ/KG) TO (BTU/LB). SO,
ENTHALPY (BTU/LB) = 0.429923*ENTHALPY (KJ/KG)
$OFFTEXT
H(I) = 0.429923*ENTH(I);

$ONTEXT
STEAM MASS FLOW RATES AT EACH HEADER LEVEL HERE STEAM MASS
FLOW (LB/HR) =((SUPPLLY-DEMAND)(MMBTU/HR))/(ENTHALPY (BTU/LB))
$OFFTXT
M(I) = (D(I,'SUPPLY')-D(I,'DEMAND'))*(106)/(H(I));**

* TURBINE EFFICIENCY ASSUMED CONSTANT AND EQUAL TO 0.7
SCALAR NU EFFICIENCY OF THE TURBINE /0.7/;

$ONTEXT
DEFINING SET K TO HAVE THE SAME MEMBERS OF SET I. THIS IS
ACTUALLY PROVIDING AN ALIAS K TO SET I. THIS IS DONE TO ENABLE
INTERACTION BETWEEN THE OWN MEMBERS OF A SET.
$OFFTEXT
ALIAS (I,K);

* DECLARING VARIABLES
**VARIABLES MASS(I,K) STEAM MASS FLOW BETWEEN DIFFERENT
          HEADERS (LB PER HOUR),
          EALL TOTAL COGENERATION POTENTIAL (MMBTU PER
          HOUR),
          EIND(I,K) COGENERATION BETWEEN DIFFERENT HEADERS
          (MMBTU PER HOUR),
          EXCESSRC(I) EXCESS STEAM AT DIFFERENT SURPLUS
          HEADERS (LB PER HOUR),
          EXCESSNK(K) EXCESS STEAM AT DIFFERENT DEFICIT
          HEADERS (LB PER HOUR)
          ECONDF(I,L) CONDENSATION TURBINE FLOW
          MASSCD(I,L) FLOW TO CONDENSING TURBINE
          ECOGEN POWER BY COGENERATION;**

* DECLARING POSITIVE VARIABLES
**POSITIVE VARIABLES MASS, EIND, EXCESSSRC, EXCESSSNK, MASSCD,
ECONDF, ECONDEN, ECOGEN;**

* DECLARING EQUATIONS

EQUATIONS EALLEQ TOTAL COGENERATION POTENTIAL,
ECOG POWER BY COGENERATION,
EINDEQ(I,K) COGENERATION BETWEEN DIFFERENT
HEADERS,
SRCONSTRAINT(I) SOURCE CONSTRAINT EQUATION,
SKCONSTRAINT(K) SINK CONSTRAINT EQUATION,
FLOWCONSTRAINT FLOW CONSTRAINT EQUATION,
EXSTEAMSRC(I) EXCESS STEAM AT SURPLUS HEADER,
EXSTEAMSNK(K) EXCESS STEAM AT DEFICIT HEADER;

* DEFINING EQUATIONS
$ONTEXT
FOLLWOING EQUATION DEFINES THAT TOTAL POWER IS COMBINATION
OF POWER BY COGENERATION AND POWER BY CONDENSATION.
$OFFTEXT
EALLEQ.. EALL =E= ECOGEN+ECONDEN;

$ONTEXT
THIS EQUATION DEFINES:
ECOGEN = EIND12+EIND13+EIND14+EINDE23+EIND24+EIND34;
THIS EQUATION DEFINES THAT POWER BY COGENERATION IS EQUAL TO
THE SUM OF ALL THE INDIVIDUAL COGENERATION POTENTIAL. LATER
FROM THE CONSTRAINTS IT IS DEFINED WHICH INDIVIDUAL
COGENERATION SHOULD EXISTS TO PROVIDE THE MAXIMUM VALUE OF
EALL. HERE THE STATEMENT SUM((I,K)$(ORD(K) GT ORD(I)),EIND(I,K))
DEFINES THAT THE SUMMATION SHOULD BE ON I AND K WHERE ORDER
OF K IS GREATER THEN ORDER OF I. IN GAMS ORD(I) OR ORDER OF I
RESULTS IN SPECIFIC POSITION OF THE MEMBER IN THE SET. SO FOR I = 1
i.e. FOR VHP, ORDER OF K IS EITHER 2, 3 OR 4, i.e. EITHER HP, MP OR LP.
THAT IS HOW IT IS GETTING EIND12+EIND13+EIND14. AND FOR I=2 IT IS
GETTING
EIND23+EIND24. AND FOR I=3 WE GET E34. THIS IS REQUIRED BECAUSE
FLOW CAN ONLY BE FROM HIGH PRESSURE TO LOW PRESSURE. THE
REVERSE IS NOT POSSIBLE. SO E21, E31, E32, E41, E42, E43 ARE NOT
POSSIBLE. ALSO FLOW CANNOT BE FROM THE HEAEDR TO ITSELF. SO E11,
E22, E33 ARE NOT POSSIBLE. HERE $(ORD(K) GT ORD(I)) IS CALLED
DOLLAR OPERATOR. SYNTAX IS $STATEMENT
$OFFTEXT
ECOG.. ECOGEN =E= SUM((I,K)$(ORD(K) GT ORD(I)),EIND(I,K));

$ONTEXT
THE FOLLOWING EQUATION IS CALCULATING INDIVIDUAL COMBINATION
OF HEADER AND CALCUALTING CORRESPONDING COGENERATION

POTENTIAL. SO THIS IS CALULATING E12, E13, E14, E23, E24, E34
SEPARATELY. THE FORMULA FOR E12 IS:
E12 = MASS12*TURBINE EFFICIENCY*(ENTHALPY(1) - NTHALPY(2))/10^6.
HERE 10^6 IS TO CHANGE THE UNIT FROM BTU/HR TO MMBTU/HR. ONE
OBSERVABLE THING IN THIS EQUATION IS THE USE OF DOLLAR
OPERATOR ON THE LEFT OF TWO DOTS (..). BY DOING THIS WE ARE
DFINING THE DOMAIN OF THE EQUATIONS TO BE GENERATED.
HERE EINDEQ(I,K)$(ORD(K) GT ORD(I)) IS IMPLYING THAT ALL THE
POSSIBLE EQUATIONS OF I AND K WHERE ORDER OF K IS GREATER THEN
ORDER OF I. ORDER IS ALREADY DISCUSSED IN PREVIOUS EQUATION.
HERE I WOULD LIKE TO DISCUSS THE IMPLICATION OF
THE DOLLAR OPERATOR ON THE EQUATIONS. BY THIS DOLLAR
OPERATOR I AM DEFINING THE SET OF EQUATIONS WHERE ORD(K) IS
GREATER THEN ORD(I). SO RATHER THEN PRODUCING 16 EQUATIONS
BETWEEN 4x4 MATRIX, ONLY FOLLOWING 6(SIX) EQUATIONS WILL BE
GENERATED FOR E12, E13, E14, E23, E24 AND E34.
$OFFTEXT
EINDEQ(I,K)$(ORD(K) GT ORD(I)).. EIND(I,K) =E= MASS(I,K)*NU*((H(I)-H(K))/(10**6));
$ONTEXT
THE FOLLOWING EQUATION SHOWS THE FLOW OF STEAM FROM
DIFFERENT HEADERS TO THE CONDENSING TURBINE.
$OFFTEXT
CONDENSE(I,L).. ECONDF(I,L) =E= MASSCD(I,L)*NU*(H(I)-CD(L))/(10**6);

$ONTEXT
THE FOLLOWING EQUATION IS SOURCE CONSTRAINT EQUATION. FOR
EXAMPLE FOR I=1, THE EQUATION IS:
MASS12+MASS13+MASS13+MASSCD(1,CONDENSE) <= M(1)
HERE THE DOLLAR OPERATOR ON THE LEFT OF THE DOTS(..) DEFINES
THAT THE EQUATION SHOULD EXISTS ONLY FOR THOSE MEMBERS OF I
THAT ARE ALSO MEMBER OF THE SURPLUS SUBSET. i.e. THE QUATIONS
ARE DEFINED ONLY OVER THE SURPLUS HEADERS. HERE ASSUMPTION IS
THAT ONLY THE SURPLUS HEADERS CAN BECOME THE SOURCE. THE
DOLLAR OPERATOR ON THE RIGHT SIDE IS SIMILAR TO THE PREVIOUS
ONES EXPLAINED EARLIER EQUATIONS. THE EQUATIONS ARE ALREADY
DEFINED ON I(BY DOMAIN MANIPULATION $ ON THE LEFT) SO THE SUM
CAN ONLY OCCUR ON K. HERE I IS FIXED BY EQUATION DOMAIN SO IN
SUM, ONLY K CAN BE VARIED. IT IS EASILY SHOWN BY SUM(K$(ORD(K) GT
ORD(I)),MASS(I,K)) ALSO SUM(L, MASSCD(I,L)) REPRESENTS THE FLOW
FROM I TO CONDENSING TURBINE.
$OFFTEXT
SRCONSTRAINT(I)$SU(I).. SUM(K$(ORD(K) GT ORD(I)),MASS(I,K))+
SUM(L,MASSCD(I,L)) =L= ABS(M(I));

$ONTEXT
THE FOLLOWING EQUATION IS SINK CONSTRAINT EQUATION AND
DEFINED OVER DEFICIT HEADERS ONLY BY THE DOLLAR OPERATOR ON
THE LEFT OF THE DOTS(..). FOR EXAMPLE FOR K=4, THE EQUATION IS:
MASS14+MASS23+MASS34 <= ABS(M(4))
HERE THE DOLLAR OPERATOR ON THE LEFT OF THE DOTS(..) DEFINES
THAT THE EQUATION SHOULD EXISTS ONLY FOR THOSE MEMBERS OF K
THAT ARE ALSO MEMBERS OF THE DEFICIT SUBSET. i.e. THE EQUATIONS
ARE DEFINED ONLY OVER THE DEFICIT HEADERS. HERE
ASSUMPTION IS THAT ONLY THE DEFICIT HEADERS CAN BECOME THE
SINK. THE DOLLAR OPERATOR ON THE RIGHT SIDE IS SIMILAR TO THE
PREVIOUS ONES EXPLAINED EARLIER EQUATIONS. THE EQUATIONS ARE
ALREADY DEFINED ON K(BY DOMAIN MANIPULATION $ ON THE LEFT) SO
THE SUM CAN ONLY OCCUR ON I. HERE K IS FIXED BY EQUATION DOMAIN
SO IN SUM, ONLY I CAN BE VARIED. IT IS EASILY SHOWN BY
SUM(I$(ORD(K) GT ORD(I)),MASS(I,K))
$OFFTEXT
**SKCONSTRAINT(K)$DF(K).. SUM(I$(ORD(K) GT ORD(I)),MASS(I,K)) =E=
ABS(M(K));**

$ONTEXT
THE FOLLOWING EQUATION IS REQUIRED TO CONTROL THE FLOW
BETWEEN DEFICIT HEADERS. FOR EXAMPLE, IF THERE ARE TWO DEFICIT
HEADERS MP(K=3) AND LP(K=4) AND IF THERE EXISTS A STEAM FLOW
FROM MP TO LP i.e. MASS34, THEN THE EQUATION SAYS
MASS34 <= MASS13+MASS23-M(3)(i.e. EXISITING DEMAND AT 3), WHERE 1,2
ARE SURPLUS HEADERS i.e. MASS34 IS ONLY POSSIBLE IF
INPUTS(MASS13+MASS23) AT 3 IS GREATER
THEN THE DEMAND AT 3; SO THE CONSTRAINT IS A DEFICIT HEADER CAN
ONLY BECOME SOURCE, IF AND ONLY IF, IT'S FLOW REQUIREMENT IS
FULLFILLED AND IT HAS EXCESS. ALSO ANOTHER ASSUMPTION HERE IS,
A DEFICIT HEADER CAN NEVER PASS STEAM TO A SURPLUS HEADER, AND
STEAM CAN GO ONLY FROM HIGHER PRESSURE LEVEL TO A LOWER
PRESSURE LEVEL.

THIS CONSTRAINT IS ONLY APPLICABLE FOR THE "GREATER THEN OR
EQUAL" CONSTRAINT AT THE SINK CONSTRAINT EQUATION. FOR "LESS
THEN OR EQUAL" CONSTRAINT IT IS NOT APPLICABLE, DUE TO THE FACT
THAT, IN THAT CASE THERE IS NO POSSIBILITY OF EXTRA STEAM AT
DEFICIT HEADER.

NOW THE FORMULATION. IN THE FOLLOWING EQUATION THE DOLLAR
OPERATOR ON THE LEFT OF THE DOT(..) RESTRICT THE DOMAIN OF K IN

THE DEFICIT SUBSET. SO IN EQUATIONS THAT WILL BE GENERATED, K WILL BE FIXED AND IT WILL ONLY BELONG TO DEFICIT HEADERS. SO ONLY I CAN BE MANIPULATED.

FIRST SUMMATION:SUM(I$(DF(I)),MASS(K,I)$(ORD(K) LT ORD(I))) ON THE LEFT OF =L=, DEFINES THE OUTPUT FLOW FROM A DEFICIT HEADER TO ANOTHER DEFICIT HEADER. AS DISCUSSED EARLIER, THE DOMAIN OF K IS ALREADY DEFINED IN DEFICIT HEADERS. BUT I IS NOT DEFINED, SO IT CAN SPAN BOTH DEFICIT AND SURPLUS HEADERS. BUT THE CONSTRAINTS ARE:
-THE FLOW CAN ONLY GO TO A DEFICIT HEADER FROM ANOTHER DEFICIT HEADER AND
- THE FLOW HAS TO BE FROM HIGHER PRESSURE (LOWER ORDER) TO LOWER PRESSURE (HIGHER ORDER).
THE FIRST CONSTRAINT IS SATISFIED BY RESTRICTING I ALSO TO ONLY DEFICIT HEADER, BY THE DOLLAR OPERATOR I$DF(I). AGAIN HERE K IS FIXED AND WE WANT TO KNOW THE OUTPUTS FROM K TO LOWER PRESSURE DEFICIT HEADERS. SO THE INDEX OF MASS IS SWITCHED FROM (I,K) TO (K,I) AND THE DOLLAR OPERATOR $(ORD(K) LT ORD(I)) DEFINES THE ORDER OF K LESS THEN ORDER OF I. SO FLOWS ARE DEFINED FROM HIGHER PRESSURE TO LOWER PRESSURE AND ONLY WITHIN DEFICIT HEADERS.

SECOND SUMMATION: SUM(L, MASSCD(K,L)) SHOWS THE FLOWS FROM THE DEFICIT HEADER TO CONDENSING TURBINE. SO IT IS ALSO PART OF OUTPUT.

THIRD SUMMATION: SUM(I$(ORD(K) GT ORD(I)), MASS(I,K)) IS SIMILAR TO THE ONES USED IN EARLIER EQUATIONS. IT DEFINES ALL THE INPUTS IN K. THE I IS NOT RESTRICTED HERE. SINCE THE INPUT CAN COME FROM ANY HIGHER PRESSURE LEVEL AND FROM BOTH SURPLUS AND DEFICIT HEADERS. HERE $(ORD(K) GT ORD(I)) DEFINES THAT
THE I HAS TO BE AT HIGHER PRESSURE (LOWER ORDER) THEN K.

THE M(K) IS DEMAND AT HEADER K. IT IS BY ITSELF A NEGATIVE NUMBER, SO IT IS JUST SHOWN AS AN ADDITION.
$OFFTEXT
FLOWCONSTRAINT(K)$DF(K)..   SUM(I$(DF(I)),MASS(K,I)$(ORD(K)   LT ORD(I)))+   SUM(L,   MASSCD(K,L))=L=   SUM(I$(ORD(K)   GT   ORD(I)), MASS(I,K))+M(K);

$ONTEXT
THE FOLLWOING EQUATION DETERMINES THE EXISTENCE OF EXCESS STEAM AT SOURCE HEADER BY DOLLAR OPERATOR ON THE LEFT OF

DOTS(..) THE DOMAIN IS DEFINED OVER SURPLUS HEADERS ONLY. THE QUATION FOR EXCESS STEAM FOR HP (I = 2, AS EXAMPLE) HEADER IS:
EXCESS(HP) = M(HP)-MASS23-MASS24+MASS12, i.e.
EXCESS = EXISTING SURPLUS-OUTPUT+INPUT.
HERE DOMAIN IS DEFINED OVER I FOR THE SURPLUS SUBSET. SO ONLY K CAN BE VARIED.

THE FIRST SUMMATION: SUM(K$(ORD(K) GT ORD(I)), MASS(I,K)) DEFINES THE OUTPUT. I IS FIXED AND K VARIED BY SUM AND $(ORD(K) GT ORD(I)) DEFINES THAT THE OUTPUTS ARE (FOR EXAMPLE FOR HP HEADER, I=2) MASS23 + MASS24.

THE SECOND SUMMATION: SUM(K$SU(K), MASS(K,I)$(ORD(K) LT ORD(I))) DEFINES THE INPUTS. HERE THE ASSUMPTION IS INPUT CAN COME FROM ANOTHER HEADER WHICH IS AT A HIGHER PRESSURE THAT IS AT A LOWER ORDER. ALSO I IS ALREADY DEFINED IN SURPLUS
BUT K IS NOT, SO HERE K IS ALSO DEFINED IN SURPLUS SUBSET. IT IS REQUIRED BECAUSE IT IS ASSUMED THAT STEAM CAN ONLY COME FROM SURPLUS HEADER. AGAIN, FOR ALL THE EQUATIONS OF THIS DOMAIN I IS FIXED. SO THE INDEX OF MASS IS SWITCHED FROM (I,K) TO (K,I), AND THE ORDER OF K IS DEFINED TO BE LESS THEN THE ORDER OF I SINCE WE WANT TO SEE THE INPUTS IN I FROM DIFFERENT K HEADERS AT HIGHER PRESSURE AND LOWER ORDER LEVEL. FLOW TO CONDENSER IS OMMITTED DELIBERATELY.
$OFFTEXT
EXSTEAMSRC(I)$SU(I).. EXCESSRC(I) =E= M(I)-SUM(K$(ORD(K) GT ORD(I)), MASS(I,K))+SUM(K$SU(K), MASS(K,I)$(ORD(K) LT ORD(I)));

$ONTEXT
THE FOLLWOING EQUATION DETERMINES THE EXISTENCE OF EXCESS STEAM AT SINK HEADER BY DOLLAR OPERATOR ON THE LEFT OF DOTS(..) THE DOMAIN IS DEFINED OVER DEFICIT HEADERS ONLY. THE QUATION FOR EXCESS STEAM FOR MP (K = 3, AS EXAMPLE) HEADER IS:
EXCESS(MP) = MASS13+MASS23-M(MP)-MASS34, i.e.
EXCESS = INPUT-EXISTING DEMAND-OUTPUT.
HERE DOMAIN IS DEFINED OVER K FOR THE SURPLUS SUBSET. SO ONLY I CAN BE VARIED.

THE FIRST SUMMATION: SUM(I$(ORD(K) GT ORD(I)), MASS(I,K)) DEFINES THE INPUT. K IS FIXED AND I VARIED BY SUM AND $(ORD(K) GT ORD(I)) DEFINES THAT THE INPUTS ARE (FOR EXAMPLE FOR MP HEADER, K=3) MASS13 + MASS23.

THE SECOND SUMMATION: SUM(I$DF(I), MASS(K,I)$(ORD(K) LT ORD(I))) DEFINES THE OUTPUTS. HERE THE ASSUMPTION IS OUTPUT CAN GO TO ANOTHER HEADER WHICH IS AT A LOWER PRESSURE THAT IS AT A HIGHER ORDER. ALSO K IS ALREADY DEFINED IN DEFICIT BUT I IS NOT, SO HERE I IS ALSO DEFINED IN DEFICIT SUBSET. IT IS REQUIRED BECAUSE IT IS ASSUMED THAT STEAM CAN ONLY GO TO A DEFICIT HEADER. AGAIN, FOR ALL THE EQUATIONS OF THIS DOMAIN K IS FIXED. SO THE INDEX OF MASS IS SWITCHED FROM (I,K) TO (K,I), AND THE ORDER OF K IS DEFINED TO BE LESS THEN THE ORDER OF I, SINCE WE WANT TO SEE THE OUTPUTS FROM K TO DIFFERENT I HEADERS AT LOWER PRESSURE AND HIGHER ORDER LEVEL. FLOW TO CONDENSER IS OMMITTED DELIBERATELY.
$OFFTEXT
EXSTEAMSNK(K)$DF(K).. EXCESSNK(K) =E= SUM(I$(ORD(K) GT ORD(I)), MASS(I,K))-ABS(M(K))-SUM(I$DF(I), MASS(K,I)$(ORD(K) LT ORD(I)));

* MODEL STATEMENT, DEFINES THE MODEL INCLUDES ALL THE   *
EQUATIONS DECLARED
MODEL STEAM /ALL/;

* SOLVE STATEMENT TO SOLVE THE MODEL, THE MODEL IS LINEAR   *
AND MAXIMIZES EALL
SOLVE STEAM USING LP MAXIMIZING EALL;

* DISPLAY STATMENTS TO DISPLAY DIFFERENT VARIABLES AND   *
VALUES
DISPLAY EALL.L;
DISPLAY ECOGEN.L;
DISPLAY EIND.L;
DISPLAY MASS.L;
DISPLAY EXCESSRC.L;
DISPLAY EXCESSNK.L;
DISPLAY MASSCD.L;

$ONTEXT
ADDING THE EXCESS STEAM AT THE MP HEADER LEVEL, FOR SUPPLYING THE COOLING LOAD FOR THE PROCESS UTLIZING ABSORPTION REFRIGERATION.
$OFFTEXT
D('MP','DEMAND') = D('MP', 'DEMAND')+10.96;

*THIS CHANGES THE FLOWRATE
M(I) = (D(I,'SUPPLY')-D(I,'DEMAND'))*(106)/(H(I));**

```
* RESOLVING THE WHOLE MODEL FOR THE UPDATED HEADER          *
BALANCE
SOLVE STEAM USING LP MAXIMIZING EALL;

* DISPLAY STATMENTS TO DISPLAY DIFFERENT VARIABLES AND       *
VALUES
DISPLAY EALL.L;
DISPLAY ECOGEN.L;
DISPLAY EIND.L;
DISPLAY MASS.L;
DISPLAY EXCESSRC.L;
DISPLAY EXCESSNK.L;
DISPLAY MASSCD.L;
DISPLAY H;
```

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COGENERATION POTENTIAL
Equation Listing   SOLVE STEAM Using LP From line 404

---- EALLEQ =E= TOTAL COGENERATION POTENTIAL

EALLEQ.. EALL - ECOGEN =E= 0 ; (LHS = 0)

---- ECOG =E= POWER BY COGENERATION

ECOG.. - EIND(HP,MP) - EIND(HP,LP) - EIND(MP,LP) + ECOGEN =E= 0 ;
(LHS = 0)

---- EINDEQ =E= COGENERATION BETWEEN DIFFERENT HEADERS

EINDEQ(HP,MP).. - 0.000152128747748739*MASS(HP,MP) + EIND(HP,MP)
=E= 0 ;

   (LHS = 0)

EINDEQ(HP,LP).. - 0.000168216981209258*MASS(HP,LP) + EIND(HP,LP) =E= 0
;

   (LHS = 0)

EINDEQ(MP,LP).. - 1.60882334605192E-5*MASS(MP,LP) + EIND(MP,LP) =E= 0
;

   (LHS = 0)

---- SRCONSTRAINT =L= SOURCE CONSTRAINT EQUATION

SRCONSTRAINT(HP).. MASS(HP,MP) + MASS(HP,LP) + MASSCD(HP,COND)
=L=
   168696.05889162 ; (LHS = 0)

---- SKCONSTRAINT =E= SINK CONSTRAINT EQUATION

SKCONSTRAINT(MP).. MASS(HP,MP) =E= 12568.1343875023 ;

(LHS = 0, INFES = 12568.1343875023 ***)

SKCONSTRAINT(LP).. MASS(HP,LP) + MASS(MP,LP) =E= 146944.322582913 ;

(LHS = 0, INFES = 146944.322582913 ***)

---- FLOWCONSTRAINT =L= FLOW CONSTRAINT EQUATION

**FLOWCONSTRAINT(MP).. - MASS(HP,MP) + MASS(MP,LP) +
MASSCD(MP,COND) =L=
-12568.1343875023 ; (LHS = 0, INFES = 12568.1343875023 ***)**

**FLOWCONSTRAINT(LP).. - MASS(HP,LP) - MASS(MP,LP) +
MASSCD(LP,COND) =L=
-146944.322582913 ; (LHS = 0, INFES = 146944.322582913 ***)**

---- EXSTEAMSRC =E= EXCESS STEAM AT SURPLUS HEADER

**EXSTEAMSRC(HP).. MASS(HP,MP) + MASS(HP,LP) + EXCESSRC(HP) =E=
168696.05889162 ;**

(LHS = 0, INFES = 168696.05889162 ***)

---- EXSTEAMSNK =E= EXCESS STEAM AT DEFICIT HEADER

**EXSTEAMSNK(MP).. - MASS(HP,MP) + MASS(MP,LP) + EXCESSNK(MP)
=E=
-12568.1343875023 ; (LHS = 0, INFES = 12568.1343875023 ***)**

**EXSTEAMSNK(LP).. - MASS(HP,LP) - MASS(MP,LP) + EXCESSNK(LP) =E=
-146944.322582913 ; (LHS = 0, INFES = 146944.322582913 ***)**

---- MASS STEAM MASS FLOW BETWEEN DIFFERENT HEADERS (LB PER HOUR)

MASS(HP,MP)
       (.LO, .L, .UP = 0, 0, +INF)
   -0.0002 EINDEQ(HP,MP)
    1   SRCONSTRAINT(HP)
    1   SKCONSTRAINT(MP)
   -1   FLOWCONSTRAINT(MP)
    1   EXSTEAMSRC(HP)
   -1   EXSTEAMSNK(MP)

MASS(HP,LP)
       (.LO, .L, .UP = 0, 0, +INF)
   -0.0002 EINDEQ(HP,LP)
    1   SRCONSTRAINT(HP)
    1   SKCONSTRAINT(LP)
   -1   FLOWCONSTRAINT(LP)
    1   EXSTEAMSRC(HP)
   -1   EXSTEAMSNK(LP)

MASS(MP,LP)
       (.LO, .L, .UP = 0, 0, +INF)
 -1.608823E-5 EINDEQ(MP,LP)
    1   SKCONSTRAINT(LP)
    1   FLOWCONSTRAINT(MP)
   -1   FLOWCONSTRAINT(LP)
    1   EXSTEAMSNK(MP)
   -1   EXSTEAMSNK(LP)

---- EALL TOTAL COGENERATION POTENTIAL (MMBTU PER HOUR)

EALL
       (.LO, .L, .UP = -INF, 0, +INF)
    1   EALLEQ

---- EIND COGENERATION BETWEEN DIFFERENT HEADERS (MMBTU PER HOUR)

EIND(HP,MP)
       (.LO, .L, .UP = 0, 0, +INF)
  -1   **ECOG**
  1   **EINDEQ(HP,MP)**

EIND(HP,LP)
       (.LO, .L, .UP = 0, 0, +INF)
  -1   **ECOG**
  1   **EINDEQ(HP,LP)**

EIND(MP,LP)
       (.LO, .L, .UP = 0, 0, +INF)
  -1   **ECOG**
  1   **EINDEQ(MP,LP)**

---- EXCESSRC EXCESS STEAM AT DIFFERENT SURPLUS HEADERS (LB PER HOUR)

EXCESSRC(HP)
      (.LO, .L, .UP = -INF, 0, +INF)
  1   **EXSTEAMSRC(HP)**

---- EXCESSNK EXCESS STEAM AT DIFFERENT DEFICIT HEADERS (LB PER HOUR)

EXCESSNK(MP)
      (.LO, .L, .UP = -INF, 0, +INF)
  1   **EXSTEAMSNK(MP)**

EXCESSNK(LP)
      (.LO, .L, .UP = -INF, 0, +INF)
  1   **EXSTEAMSNK(LP)**

---- MASSCD FLOW TO CONDENSING TURBINE

MASSCD(HP,COND)
      (.LO, .L, .UP = 0, 0, +INF)
  1   **SRCONSTRAINT(HP)**

MASSCD(MP,COND)
       (.LO, .L, .UP = 0, 0, +INF)
   1   **FLOWCONSTRAINT(MP)**

MASSCD(LP,COND)
       (.LO, .L, .UP = 0, 0, +INF)
   1   **FLOWCONSTRAINT(LP)**

---- ECOGEN POWER BY COGENERATION

ECOGEN
       (.LO, .L, .UP = 0, 0, +INF)
  -1   **EALLEQ**
   1   **ECOG**

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COGENERATION POTENTIAL
Model Statistics   SOLVE STEAM Using LP From line 404

MODEL STATISTICS

BLOCKS OF EQUATIONS   8   SINGLE EQUATIONS   13
BLOCKS OF VARIABLES   7   SINGLE VARIABLES   14
NON ZERO ELEMENTS   33

GENERATION TIME   =   0.060 SECONDS   1.6 Mb   WIN210-134

EXECUTION TIME   =   0.060 SECONDS   1.6 Mb   WIN210-134

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COGENERATION POTENTIAL
Solution Report   SOLVE STEAM Using LP From line 404

          S O L V E   S U M M A R Y

   MODEL   STEAM   OBJECTIVE EALL
   TYPE   LP   DIRECTION MAXIMIZE
   SOLVER CPLEX   FROM LINE 404

**** SOLVER STATUS   1 NORMAL COMPLETION
**** MODEL STATUS   1 OPTIMAL
**** OBJECTIVE VALUE   26.6305

 RESOURCE USAGE, LIMIT   0.030   1000.000
 ITERATION COUNT, LIMIT   0   10000

GAMS/Cplex   May 15, 2003 WIN.CP.CP 21.0 023.025.041.VIS For Cplex 8.1
Cplex 8.1.0, GAMS Link 23

Optimal solution found.
Objective :   26.630505

          LOWER   LEVEL   UPPER   MARGINAL

---- EQU EALLEQ   .   .   .   1.000
---- EQU ECOG   .   .   .   1.000

 EALLEQ TOTAL COGENERATION POTENTIAL
 ECOG POWER BY COGENERATION

---- EQU EINDEQ COGENERATION BETWEEN DIFFERENT HEADERS

     LOWER   LEVEL   UPPER   MARGINAL

HP.MP   .   .   .   1.000
HP.LP   .   .   .   1.000
MP.LP   .   .   .   1.000

---- EQU SRCONSTRAINT SOURCE CONSTRAINT EQUATION

     LOWER   LEVEL   UPPER   MARGINAL

HP   -INF 1.5951E+5 1.6870E+5   .

---- EQU SKCONSTRAINT SINK CONSTRAINT EQUATION

     LOWER   LEVEL   UPPER   MARGINAL

MP 12568.134 12568.134 12568.134 1.5213E-4
LP 1.4694E+5 1.4694E+5 1.4694E+5 1.6822E-4

---- EQU FLOWCONSTRAINT FLOW CONSTRAINT EQUATION

     LOWER   LEVEL   UPPER   MARGINAL

MP   -INF -1.257E+4 -1.257E+4   .
LP   -INF -1.469E+5 -1.469E+5   .

---- EQU EXSTEAMSRC EXCESS STEAM AT SURPLUS HEADER

     LOWER   LEVEL   UPPER   MARGINAL

HP 1.6870E+5 1.6870E+5 1.6870E+5   EPS

---- EQU EXSTEAMSNK EXCESS STEAM AT DEFICIT HEADER

     LOWER   LEVEL   UPPER   MARGINAL

MP -1.257E+4 -1.257E+4 -1.257E+4   EPS
LP -1.469E+5 -1.469E+5 -1.469E+5   EPS

---- VAR MASS STEAM MASS FLOW BETWEEN DIFFERENT HEADERS (LB PER HOUR)

       LOWER   LEVEL   UPPER   MARGINAL

HP.MP   .   12568.134   +INF   .
HP.LP   .   1.4694E+5   +INF   .
MP.LP   .   .   +INF -1.521E-4

          LOWER   LEVEL   UPPER   MARGINAL

---- VAR EALL   -INF   26.631   +INF   .

  EALL TOTAL COGENERATION POTENTIAL (MMBTU PER HOUR)

---- VAR EIND COGENERATION BETWEEN DIFFERENT HEADERS (MMBTU PER HOUR)

LOWER   LEVEL   UPPER   MARGINAL

HP.MP   .   1.912   +INF   .
HP.LP   .   24.719   +INF   .
MP.LP   .   .   +INF   .

---- VAR EXCESSRC EXCESS STEAM AT DIFFERENT SURPLUS HEADERS (LB PER HOUR)

LOWER   LEVEL   UPPER   MARGINAL

HP   -INF 9183.602   +INF   .

---- VAR EXCESSNK EXCESS STEAM AT DIFFERENT DEFICIT HEADERS (LB PER HOUR)

LOWER   LEVEL   UPPER   MARGINAL

MP   -INF   .   +INF   .
LP   -INF   .   +INF   .

---- VAR MASSCD FLOW TO CONDENSING TURBINE

LOWER   LEVEL   UPPER   MARGINAL

HP.COND   .   .   +INF   EPS
MP.COND   .   .   +INF   EPS
LP.COND   .   .   +INF   EPS

LOWER   LEVEL   UPPER   MARGINAL

---- VAR ECOGEN   .   26.631   +INF   .

ECOGEN POWER BY COGENERATION

**** REPORT SUMMARY :   0   NONOPT
0 INFEASIBLE
0 UNBOUNDED

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COGENERATION POTENTIAL
E x e c u t i o n

----   408 VARIABLE EALL.L   =   26.631 TOTAL COGENERATION PO

                                      TENTIAL (MMBTU PER HO
                                      UR)

----   409 VARIABLE ECOGEN.L   =   26.631 POWER BY COGENERATION

----   410 VARIABLE EIND.L COGENERATION BETWEEN DIFFERENT HEADERS (MMBTU PER H
                  OUR)

         MP   LP

HP   1.912   24.719

----   411 VARIABLE MASS.L STEAM MASS FLOW BETWEEN DIFFERENT HEADERS (LB PER H
                  OUR)

         MP   LP

HP   12568.134 146944.323

----   412 VARIABLE EXCESSRC.L EXCESS STEAM AT DIFFERENT SURPLUS HEADERS (LB P
                  ER HOUR)

HP 9183.602

----   413 VARIABLE EXCESSNK.L EXCESS STEAM AT DIFFERENT DEFICIT HEADERS (LB P
                  ER HOUR)

413 VARIABLE EXCESSNK.L EXCESS STEAM AT DIFFERENT DEFICIT HEADERS (LB P
ER HOUR)

(ALL   0.000)

----   414 VARIABLE MASSCD.L FLOW TO CONDENSING TURBINE

(ALL   0.000)

---- EALLEQ =E= TOTAL COGENERATION POTENTIAL

EALLEQ.. EALL - ECOGEN =E= 0 ; (LHS = 0)

---- ECOG =E= POWER BY COGENERATION

**ECOG.. - EIND(HP,MP) - EIND(HP,LP) - EIND(MP,LP) + ECOGEN =E= 0 ;
(LHS = 0)**

---- EINDEQ =E= COGENERATION BETWEEN DIFFERENT HEADERS

**EINDEQ(HP,MP).. - 0.000152128747748739*MASS(HP,MP) + EIND(HP,MP)
=E= 0 ;**

   (LHS = 0)

**EINDEQ(HP,LP).. - 0.000168216981209258*MASS(HP,LP) + EIND(HP,LP) =E= 0
;**

   (LHS = 0)

**EINDEQ(MP,LP).. - 1.60882334605192E-5*MASS(MP,LP) + EIND(MP,LP) =E= 0
;**

   (LHS = 0)

---- SRCONSTRAINT =L= SOURCE CONSTRAINT EQUATION

**SRCONSTRAINT(HP).. MASS(HP,MP) + MASS(HP,LP) + MASSCD(HP,COND)
=L=
   168696.05889162 ; (LHS = 159512.456970416)**

---- SKCONSTRAINT =E= SINK CONSTRAINT EQUATION

SKCONSTRAINT(MP).. MASS(HP,MP) =E= 21751.2512466374 ;

(LHS = 12568.1343875023, INFES = 9183.11685913504 ***)

SKCONSTRAINT(LP).. MASS(HP,LP) + MASS(MP,LP) =E= 146944.322582913 ;

   (LHS = 146944.322582913)

---- FLOWCONSTRAINT =L= FLOW CONSTRAINT EQUATION

FLOWCONSTRAINT(MP).. - MASS(HP,MP) + MASS(MP,LP) +
MASSCD(MP,COND) =L=
   -21751.2512466374 ; (LHS = -12568.1343875023, INFES = 9183.11685913504
***)

FLOWCONSTRAINT(LP).. - MASS(HP,LP) - MASS(MP,LP) +
MASSCD(LP,COND) =L=
   -146944.322582913 ; (LHS = -146944.322582913)

---- EXSTEAMSRC =E= EXCESS STEAM AT SURPLUS HEADER

EXSTEAMSRC(HP).. MASS(HP,MP) + MASS(HP,LP) + EXCESSRC(HP) =E=
168696.05889162 ;

   (LHS = 168696.05889162)

---- EXSTEAMSNK =E= EXCESS STEAM AT DEFICIT HEADER

EXSTEAMSNK(MP).. - MASS(HP,MP) + MASS(MP,LP) + EXCESSNK(MP)
=E=
   -21751.2512466374 ; (LHS = -12568.1343875023, INFES = 9183.11685913504
***)

EXSTEAMSNK(LP).. - MASS(HP,LP) - MASS(MP,LP) + EXCESSNK(LP) =E=
   -146944.322582913 ; (LHS = -146944.322582913)

---- MASS STEAM MASS FLOW BETWEEN DIFFERENT HEADERS (LB PER HOUR)

MASS(HP,MP)
          (.LO, .L, .UP = 0, 12568.1343875023, +INF)
    -0.0002 EINDEQ(HP,MP)
     1   SRCONSTRAINT(HP)
     1   SKCONSTRAINT(MP)
    -1   FLOWCONSTRAINT(MP)
     1   EXSTEAMSRC(HP)
    -1   EXSTEAMSNK(MP)

MASS(HP,LP)
          (.LO, .L, .UP = 0, 146944.322582913, +INF)
    -0.0002 EINDEQ(HP,LP)
     1   SRCONSTRAINT(HP)
     1   SKCONSTRAINT(LP)
    -1   FLOWCONSTRAINT(LP)
     1   EXSTEAMSRC(HP)
    -1   EXSTEAMSNK(LP)

MASS(MP,LP)
          (.LO, .L, .UP = 0, 0, +INF)
  -1.608823E-5 EINDEQ(MP,LP)
     1   SKCONSTRAINT(LP)
     1   FLOWCONSTRAINT(MP)
    -1   FLOWCONSTRAINT(LP)
     1   EXSTEAMSNK(MP)
    -1   EXSTEAMSNK(LP)

---- EALL TOTAL COGENERATION POTENTIAL (MMBTU PER HOUR)

EALL
          (.LO, .L, .UP = -INF, 26.6305048966457, +INF)
     1   EALLEQ

---- EIND COGENERATION BETWEEN DIFFERENT HEADERS (MMBTU PER HOUR)

EIND(HP,MP)
        (.LO, .L, .UP = 0, 1.9119745459086, +INF)
   -1   ECOG
    1   EINDEQ(HP,MP)

EIND(HP,LP)
        (.LO, .L, .UP = 0, 24.7185303507371, +INF)
   -1   ECOG
    1   EINDEQ(HP,LP)

EIND(MP,LP)
        (.LO, .L, .UP = 0, 0, +INF)
   -1   ECOG
    1   EINDEQ(MP,LP)

---- EXCESSRC EXCESS STEAM AT DIFFERENT SURPLUS HEADERS (LB PER HOUR)

EXCESSRC(HP)
        (.LO, .L, .UP = -INF, 9183.60192120478, +INF)
    1   EXSTEAMSRC(HP)

---- EXCESSNK EXCESS STEAM AT DIFFERENT DEFICIT HEADERS (LB PER HOUR)

EXCESSNK(MP)
        (.LO, .L, .UP = -INF, 0, +INF)
    1   EXSTEAMSNK(MP)

EXCESSNK(LP)
        (.LO, .L, .UP = -INF, 0, +INF)
    1   EXSTEAMSNK(LP)

---- MASSCD FLOW TO CONDENSING TURBINE

MASSCD(HP,COND)
        (.LO, .L, .UP = 0, 0, +INF)
    1   SRCONSTRAINT(HP)

MASSCD(MP,COND)
      (.LO, .L, .UP = 0, 0, +INF)
   1   FLOWCONSTRAINT(MP)

MASSCD(LP,COND)
      (.LO, .L, .UP = 0, 0, +INF)
   1   FLOWCONSTRAINT(LP)

---- ECOGEN POWER BY COGENERATION

ECOGEN
      (.LO, .L, .UP = 0, 26.6305048966457, +INF)
  -1   EALLEQ
   1   ECOG

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COGENERATION POTENTIAL
Model Statistics   SOLVE STEAM Using LP From line 419

MODEL STATISTICS

BLOCKS OF EQUATIONS   8   SINGLE EQUATIONS   13
BLOCKS OF VARIABLES   7   SINGLE VARIABLES   14
NON ZERO ELEMENTS   33

GENERATION TIME   =   0.040 SECONDS   1.5 Mb   WIN210-134

EXECUTION TIME   =   0.040 SECONDS   1.5 Mb   WIN210-134

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COGENERATION POTENTIAL
Solution Report   SOLVE STEAM Using LP From line 419

        S O L V E   S U M M A R Y

   MODEL   STEAM   OBJECTIVE EALL
   TYPE   LP   DIRECTION MAXIMIZE
   SOLVER CPLEX   FROM LINE 419

**** SOLVER STATUS   1 NORMAL COMPLETION
**** MODEL STATUS   1 OPTIMAL
**** OBJECTIVE VALUE   28.0275

 RESOURCE USAGE, LIMIT   0.010   1000.000
 ITERATION COUNT, LIMIT   0   10000

GAMS/Cplex   May 15, 2003 WIN.CP.CP 21.0 023.025.041.VIS For Cplex 8.1
Cplex 8.1.0, GAMS Link 23

Optimal solution found.
Objective :   28.027521

        LOWER   LEVEL   UPPER   MARGINAL

---- EQU EALLEQ   .   .   .   1.000
---- EQU ECOG   .   .   .   1.000

 EALLEQ TOTAL COGENERATION POTENTIAL
 ECOG POWER BY COGENERATION

---- EQU EINDEQ COGENERATION BETWEEN DIFFERENT HEADERS

     LOWER   LEVEL   UPPER   MARGINAL

HP.MP   .   .   .   1.000
HP.LP   .   .   .   1.000
MP.LP   .   .   .   1.000

---- EQU SRCONSTRAINT SOURCE CONSTRAINT EQUATION

    LOWER   LEVEL   UPPER   MARGINAL

---- VAR EIND COGENERATION BETWEEN DIFFERENT HEADERS (MMBTU PER HOUR)

	LOWER	LEVEL	UPPER	MARGINAL
HP.MP	.	3.309	+INF	.
HP.LP	.	24.719	+INF	.
MP.LP	.	.	+INF	.

---- VAR EXCESSRC EXCESS STEAM AT DIFFERENT SURPLUS HEADERS (LB PER HOUR)

	LOWER	LEVEL	UPPER	MARGINAL
HP	-INF	0.485	+INF	.

---- VAR EXCESSNK EXCESS STEAM AT DIFFERENT DEFICIT HEADERS (LB PER HOUR)

	LOWER	LEVEL	UPPER	MARGINAL
MP	-INF	.	+INF	.
LP	-INF	.	+INF	.

---- VAR MASSCD FLOW TO CONDENSING TURBINE

	LOWER	LEVEL	UPPER	MARGINAL
HP.COND	.	.	+INF	EPS
MP.COND	.	.	+INF	EPS
LP.COND	.	.	+INF	EPS

	LOWER	LEVEL	UPPER	MARGINAL
---- VAR ECOGEN	.	28.028	+INF	.

 ECOGEN POWER BY COGENERATION

**** REPORT SUMMARY :   0   NONOPT
              0 INFEASIBLE
              0 UNBOUNDED

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COGENERATION POTENTIAL
E x e c u t i o n

----   422 VARIABLE EALL.L   =   28.028 TOTAL COGENERATION PO

                                    TENTIAL (MMBTU PER HO
                                    UR)

----   423 VARIABLE ECOGEN.L   =   28.028 POWER BY COGENERATION

----   424 VARIABLE EIND.L COGENERATION BETWEEN DIFFERENT HEADERS (MMBTU PER H
                   OUR)

         MP   LP

HP   3.309   24.719

----   425 VARIABLE MASS.L STEAM MASS FLOW BETWEEN DIFFERENT HEADERS (LB PER H
                   OUR)

         MP   LP

HP   21751.251 146944.323

----   426 VARIABLE EXCESSRC.L EXCESS STEAM AT DIFFERENT SURPLUS HEADERS (LB P
                   ER HOUR)

HP 0.485

----   427 VARIABLE EXCESSNK.L EXCESS STEAM AT DIFFERENT DEFICIT HEADERS (LB P
                   ER HOUR)

427 VARIABLE EXCESSNK.L EXCESS STEAM AT DIFFERENT DEFICIT HEADERS (LB P
                ER HOUR)

       (ALL   0.000)

----   428 VARIABLE MASSCD.L FLOW TO CONDENSING TURBINE

       (ALL   0.000)

----   429 PARAMETER H ENTHALPY AT HEADER CONDITION (BTU PER LB)

HP 1410.821,   MP 1193.495,   LP 1170.511

EXECUTION TIME   =   0.030 SECONDS   1.5 Mb   WIN210-134

USER: Advanced GAMS Class   G040108:1749AJ-WIN
  Jan., 2004 College Station, Texas   DC4545

**** FILE SUMMARY

INPUT   C:\DOCUME~1\RUBAYA~1\DESKTOP\THESIS~1\TH09F5~1.GMS
OUTPUT   C:\WINDOWS\GAMSDIR\TH09F5~1.LST

VITA

Rubayat Mahmud, son of Mohsin Ali Munshi and Mahmuda Yasmin, was born on December 28th, 1974, in Chuadanga, Bangladesh. He received his Bachelor of Science degree in chemical engineering in December 1999, from Bangladesh University of Engineering & Technology. After receiving his B.S. degree, he studied one semester in the MBA program, in 2000, at the Institute of Business Administration, under Dhaka University. Then he joined Texas Tech University in January 2001. He earned his Master of Science degree in chemical engineering in August 2002, from Texas Tech University. He joined the PhD program in chemical engineering at Texas A&M University in August 2002. He married Farzana Diba in June 9th, 2003. They were blessed with a beautiful daughter Raisa Tarannum Mahmud in June 22nd, 2004.

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