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Optimization and Supervisory Control of Cogeneration Systems

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Optimization and Supervisory Control of Cogeneration Systems

by

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Abstract

Optimization and supervisory control of cogeneration systems is studied in this thesis. Optimizing a cogeneration system is not only an optimization problem, but also a control one. The optimization addresses the problem of both equipment selection and equipment operation by handling it as a mixed integer non-linear programming (MINLP) problem. The MINLP problem is then solved using a novel three stage linear programming algorithm. The results show significant cost benefits across wide operation scenarios in the model cogeneration systems. The supervisory control addresses the problem of taking the plant to its optimal operating point while maintaining the critical plant parameters. A fuzzy supervisory control is developed and tested on a virtual cogeneration plant developed in SIMULINK. The results show that the fuzzy controller performs to expectation and tracks the optimal setpoints while maintaining critical plant parameters within limits.

Preface

It was a warm sunny day, as days usually are in Chennai, India, when I got the assignment. The gas turbine major overhaul had just been completed. Everyone was secretly pleased and relieved to get the thing working after taking it apart piece by piece and reassembling it. Close to the equivalent of a million dollars had been spent on the overhaul and the plant management needed some justification for the cost. The overworked team of engineers and technicians just wanted to go home. It was decided to conduct an efficiency analysis of the machine before and after overhaul to know about the performance improvement. The decision reached my manager and the work to me. As an electrical engineer who had close to zero knowledge of thermodynamics, I was in a difficult situation. However my immediate supervisor and thermodynamics master of the department decided to help me out. He gave me a dusty old book to read (yes, it was a bunch of documents released by Bureau of Energy Efficiency (BEE) India, but the dusty old book had a more dramatic effect). My research took me from simple efficiency calculation to complex thermodynamic equations; which I must confess I still do not understand completely, from LP problems to NP hard MINLP questions and finally to a subject I love – control theory. The thesis lying in front of you is the culmination of the work I started almost three years ago on that warm sunny day.

The work in general can be divided into two parts, optimization and supervisory control. Optimization of cogeneration systems is discussed in chapter 2 through 4, while supervisory control is discussed in chapters 4 through 7. Optimization of cogeneration systems is a well-known topic. However, most of the formulations do not tackle the wide operational scenarios faced in cogeneration systems. The novel contribution of the thesis is the development of a three stage algorithm that can be customised to solve diverse operational scenarios. Fuzzy control of power plant is well discussed in the literature, but most of the formulations tend to focus on individual control loops. Using fuzzy to achieve plant wide control in a power plant is seldom discussed and to the best of my knowledge has never been attempted on a cogeneration plant. The original contribution stems from the same. The thesis proposes a high level supervisory fuzzy control system that will work in tandem with a plant optimizer.

It is my personal belief that no work is complete if not applied in real life. Though the thesis reaches an end with a virtual simulation, I believe the research done here gives a solid platform to work towards a viable real life implementation of optimizing a cogeneration system.

Acknowledgements

None of this would have been possible if Prof. Om Malik had not taken me under his guidance. He was the person who believed in me and to that no amount of dedication is grand enough. He gave me freedom to do my own research and his sage advice in academic matters helped me avoid numerous pitfalls. I would also like to thank Prof. Laleh Behjat, Prof. Hamid Zareipour, Prof. Jeff Piper, Prof. Ed Nowicki and Prof. Chris Macnab whose courses I attended. They have played an invaluable part in moulding the thesis as it is seen today.

The inspiration for the thesis began from my industrial experience and I am indebted to all my colleagues but I would like to single out five gentlemen who have shaped me for who I am today. My first manager Mr. K. Rajasekaran whose “yes boss” phrase has become a part of my life as his cool demeanour. Mr. S. Srinivasan, who taught me more about practical and theoretical engineering than any one single person on this planet. Mr. E. Rajendran with whom I have shared many a lighter moments during morose work hours. Mr. R. Thyagarajan my partner in crime to umpteen experiments with million dollar plant and control equipment, and finally Mr. S. Ramasamy (affectionately called GT Rams) to whom I owe my practical knowledge about cogeneration systems.

The first honours in friendship belong to Pamela Manjarres and Syed Sabbir Ahmed. We took the same classes, worked on the same projects, lived and laughed the same during the last two years. To me they were the family I had in Canada. If the thesis has good grammar and language I owe that to Virali Shah who proof read my thesis.

All work and no play makes a dull day, and it would only be appropriate if I mention with whom I had fun. Robinder Khehra, Akshay Marahte, Suhas Illiah, Anand Narayanan, Rachel L’orsa, Gagan Mudher and Edward Dizon belong to the illustrious group who were steadfast with me during my highs and lows.

The acknowledgement would not be complete if I did not mention my parents. They supported me through years of education and the years in this master’s program was no exception. I would like to believe I have made them proud. My father would still nag me to get a PhD, as he has one too; though my mother would beg to differ as she believes that we already have one PhD too much in our family.

To the tireless and dedicated professionals in the power industry

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List of Abbreviations

Abbreviation	Definition
BFW	Boiler Feed Water
DM Water	De Mineralized Water
DCS	Distributed Control System
DMC	Dynamic Matrix Control
FLC	Fuzzy Logic Control
GPC	Generalized Predictive Control
GT	Gas Turbine
GUI	Graphical User Interface
HP	High Pressure
HRSG	Heat Recovery Steam Generator
IFC	Incremental Fuel Consumption
ISG	Incremental Steam Generation
LC	Loss Coefficient
LP	Low Pressure
MIDACO	Mixed Integer Ant Colony Optimization
MINLP	Mixed Integer Non Linear Programming
MIP	Mixed Integer Programming
MP	Medium Pressure
NLF	No Load Fuel consumption
NLP	Non-Linear Programming
NSP	Negative Set Point
PRDS	Pressure Reducer De-Superheater
P SP	Positive Set Point
SFR	Steam Fuel Ratio
sN SP	Small Negative Set Point
SP	Set Point
sP SP	Small Positive Set Point
ZSP	Zero Set Point

Chapter One: Introduction

1.1 Overview of Cogeneration Systems

Traditional power plants are typically based on Rankine (steam) or Bryton (gas) thermal cycles. Here the energy conversion takes place from chemical energy (stored in fuel) to thermal energy (present in steam/combustion gases), to mechanical energy produced by a turbine, and finally to electrical energy generated in the alternator. The thermal cycles used have an inherent disadvantage of rejecting heat energy to a thermal sink. In steam cycle, bulk of thermal energy is lost in the condenser, while in gas cycle the thermal energy is lost through the hot flue gases in the stack. As a consequence thermal efficiencies of power plants are low, about 30-40% for a steam cycle and 30-35% for a gas cycle power plant [1].

Cogeneration systems, unlike traditional power plants, produce both electric power and process heat. The process heat, usually in the form of steam is used by downstream industrial plants or for district heating. Due to the fact that the thermal energy is not rejected but converted into process heat, cogeneration systems have high efficiency, typically about 80-85% [1].
Cogeneration systems can operate in different thermal cycles, but are broadly classified into two types.

- Steam cycle cogeneration systems
- Gas cycle cogeneration systems

A steam cycle cogeneration system operates on the principle of modified Rankine cycle. It consists of a battery of boilers producing high pressure steam which is then used to run a steam turbine. Steam turbines in cogeneration systems are generally of the back-pressure, single extraction or double extraction type [1]. The extraction steam from the turbine is used to feed low pressure steam headers in the system.

A gas cycle cogeneration system operates on the principle of modified Bryton cycle. It consists of a gas turbine coupled to the generator which produces electrical power. The high temperature exhaust flue gases from the gas turbine are directed to a downstream Heat Recovery Steam Generation (HRSG) unit. Thus the gas turbine combined with the HRSG forms the heart of the cogeneration system.

In addition, all cogeneration systems have plant auxiliary systems that are necessary for normal plant operation [1]. They include a deaerator system that produces Boiler Feed Water

(BFW), heat exchangers for cooling and process heating, cooling towers for rejecting waste heat, Pressure Reducer De-Superheater (PRDS) for interconnection of steam headers etc. To handle additional steam load, boiler houses are a common part of the cogeneration system.

Most cogeneration plants operate on the n+1 concept; i.e. for a block load of “n”, there is always “n+1” power and steam generation capacity available. This increases system reliability as the downstream process units won’t be affected by a single generator or boiler trip. However, cogeneration plant optimization is seldom done and it is left at the discretion of the plant operator to maintain the generator and boiler load depending on the process requirements. The combination of the above factors provides immense opportunity to optimize the plant operation so as to reduce operating cost.

1.2 Research Objectives

Optimizing a cogeneration system is not only a complex optimization problem but also a control one. The thesis addresses both these problems and can be broadly divided into two parts

- Part 1 : Optimization of cogeneration systems (Chapters 2-4)
- Part 2 : Control of cogeneration systems (Chapters 5-7)

The research objectives are formulated independently for both parts. A brief overview of the background work and the research objectives is given below.

1.2.1 Optimization of Cogeneration Systems

Academic papers that discuss optimization of cogeneration systems primarily focus on reducing the operating cost. The operation of these plants is optimized by reducing the fuel consumption of the power and steam generation units, and also by bringing down auxiliary costs associated with boiler feed water, steam conversion, etc. On an average optimization brings about 7-8% reduction in total cost [2]. Optimization studies done on cogeneration plants diverge on the approach of modeling and solving the system. The constraints and cost functions vary from one plant to another, making the optimization problem unique for each model.

Different solution techniques have been applied to solve a variety of problems related to the cogeneration systems. They can broadly be classified into the non-linear and integer programming categories. In order to solve these nonlinear problems several methods are used. In [3] constraints are relaxed to the point where the electric and steam systems are separated and treated as separate linear programming problems. Another approach as seen in [2] utilized nonlinear solution techniques such as Generalized Reduce Gradient method. The study done in [4] exhibits the use of integer variables in a cogeneration plant optimization. They are introduced as on/off (binary) decision variables for equipment on the plant that results in a mixed integer linear program, worked out with the CPLXTM Optimization solver.

Two main problem statements addressed in this thesis are:

- Selecting which equipment to run under the specified constraints (an integer programming problem)
- How to run the selected equipment so as to achieve minimum operating cost (a linear programming problem)

Combining the two to achieve minimum cost of operation, the optimization problem at hand becomes a Mixed Integer Non-Linear Programming (MINLP) problem. Based on the above the research objectives for the optimization part of the thesis are defined as follows:

- Modelling cogeneration plant equipment and headers for formulating the optimization problem
- Developing an algorithm for solving the optimization problem at hand
- Extending the algorithm to solve optimization problems under various operating conditions
- Validating and comparing the results with a black box solver.

1.2.2 Control of Cogeneration Systems

Once the optimal plant operating conditions are known from an optimization run, the next step is to take the plant from its present operating condition to the optimal operating point. Model Predictive Control (MPC) techniques like Dynamic Matrix Control (DMC) have been extensively used in chemical and petrochemical industries for the past twenty years and are still the most commonly used advanced control systems in an industrial setting [5]-[9]. However, implementation of MPC in power industry is still in its nascent stages. The few reported implementations use a reduced order non-linear plant models and ad-hoc MPC techniques [10]. In academic circles, for power plant control, predictive control techniques like Generalised Predictive Control (GPC) [11]-[12]; intelligent control techniques like neural-network control [13]-[14], fuzzy control [15]-[18] and neuro-fuzzy control [19]-[20] have been discussed. However high level controls for cogeneration systems have seldom been discussed in the literature. As cogeneration systems are popular in petrochemical plants, a few industrial implementations of high level control for cogeneration have been done using DMC. An example of DMC used for cogeneration is shown in [21].

The primary research objective for the control part of the thesis is the development of a high level control system that has the following characteristics:

- Take the plant from one operating point to another safely. In other words, the deviations from critical plant parameters like desired power and steam header pressure levels must be kept to the minimum and within limits.
- Take minimum possible time to take plant from one point to another so that the next optimization run can begin.
- Be capable of handling system disturbances like changes in process power and steam load while keeping the critical system parameters within limits.
- Be hierarchically a high-level supervisory controller and must seamlessly integrate with the existing plant control system.
- Be easy to setup, require minimum system identification tests and should minimize plant downtime during implementation.

1.3 Major Contributions

The major research contributions of the optimization part of the thesis are as below:

- ➢ Development of optimization models for ten different types of cogeneration plant equipment and headers
- ➢ Development of a framework that supports automated problem formation for a given plant model
- ➢ Development of a novel three stage linear programming algorithm based on constraint relaxation that solves the optimization problem under various operating conditions

The major research contributions for the control part of the thesis are as below:

- ➢ Development of a supervisory fuzzy set point controller with the following characteristics:
  - Works in tandem with a plant optimizer and takes the plant to optimal operating point.
  - Achieves plant wide control by keeping the plant critical parameters under check during transition to optimal operating points and system disturbances.

In addition to the above a cogeneration simulation suite has been developed to aid the research work. The package has three main software modules:

- Plant Graphical user interface (GUI), that mimics a plant operator station
- Integrated optimizer, where the three stage optimization algorithm is implemented
- Plant real time module, where the dynamic plant system and the supervisory control system are simulated.

The software modules provide a good platform not only to add and test new plant models but also to test new advanced control algorithms.

1.4 Thesis Structure

- Chapter 2: The optimization models of the plant equipment and headers used in the thesis are described in this chapter. The chapter also gives an overview of modelling framework used and describes its advantages and limitations.
- Chapter 3: This chapter details how the optimization problem is formulated and solved. A detailed description of the novel three stage algorithms is given along with the extensions used to solve the problem under various operating conditions.
- Chapter 4: The algorithm developed in Chapter 3 is tested on two model cogeneration plants. This chapter describes the model plants used and proceeds to the results. The analysis and comparison of the results with a black box solver are also presented in this chapter.
- Chapter 5: This chapter details the dynamic models developed for simulating the cogeneration plant. Individual model simulations and model descriptions are discussed.
- Chapter 6: A detailed description of the fuzzy control scheme used to develop the supervisory controller is given in this chapter. Five different types of fuzzy controllers that are used in the main supervisory control are explained in detail. A brief overview of the control hierarchy in modern process and power plants is also given.
- Chapter 7: This chapter gives the results of applying the supervisory control scheme in a model cogeneration plant. The simulation of the plant under transition from normal to optimal conditions is shown along with plant disturbance tests. This chapter also describes the working of the software suite developed during the research.
- Chapter 8: The final chapter summarizes the results, the major research contributions of the thesis and its importance.

Chapter Two: Approach to Optimization Modelling

Plant models developed in this project are both statistical and thermodynamic in nature. Statistical models are chosen when the thermodynamic models of the equipment are complex or unknown and the thermodynamic models are chosen if they can be linearized.

Two different types of plant models are used in this thesis. One represents a cogeneration facility in a Refinery complex and the other represents a district heat and power utility system. The plant equipment details and configuration for the two model plants are given in Appendices A and B, respectively.

Even though the number and configuration of the plant equipment is different for each model cogeneration system, it is seen that both have similar equipment. They can be summarized as below:

- Gas Turbine: Industrial gas turbine unit used for power generation. Hot flue gas from the GT is routed to a steam generator for producing process steam.
- Heat Recovery Steam Generator: Unfired Boiler unit present in the downstream of a GT. It takes the hot flue gas from the GT and uses it for producing process steam.
- Boiler: Utility level fired boiler unit used for steam generation. Typically cogeneration facilities use oil/gas fired boiler units.
- Deaerator: Converts DM water into boiler feed water by physically stripping gases and dissolved oxygen.
- Pressure Reducer De-Superheater: Provides interconnection between a high pressure steam header to a low pressure steam header and facilitates steam flow balance.
- Plant headers: Headers are usually pipes that carry a particular fluid across the plant. Different types of headers are present in a cogeneration unit, but the most common ones are the DM water header, Boiler Feed Water (BFW) header, steam header and fuel header which carry DM water, BFW, process steam and fuel, respectively.

2.1 Modelling limitations

- Rigorous thermodynamic models for power plant equipment are fundamentally nonlinear. However, models derived in this project for plant and plant auxiliaries are linear.
- The models do not give accurate predictions for transient plant conditions (plant exigencies, start-ups and shutdowns).

2.2 Modelling Framework

The complex plant is divided into distinct equipment and headers. Each equipment and header will have at least one of the following attributes:

- Unique input variables
- Unique output variable
- Unique equality/inequality at the equipment/header boundary

The above framework has the following advantages:

- Expresses the distinct relationship between the inputs and outputs at the equipment/header battery limit.
- Modelling of the plant becomes easier as there are well defined variables and equations associated with each equipment/header.
- Suits well to an algorithmic format that can be used for automated constraint listing.

Based on the above framework the equipment and headers as listed in Table 2.1 were chosen. Appendix C gives an overview of how the steady state optimization model was developed and validated for a GT-HRSG unit.

Table 2.1 - Plant Equipment and Header List

Plant Equipment	Plant Headers
1. Gas Turbine (GT)	1. DM(Demineralized)-Water Header
2. Heat Recovery Steam Generator (HRSG)	2. Boiler feed water (BFW) Header
3. Boiler	3. Steam Header
4. De-aerator	4. Fuel Header
5. Pressure Reducer De-Superheater (PRDS)	5. Power Header

2.3 Plant and Equipment Modelling

The traditional approach to optimization is to reduce the number of variables in the system. This approach works well for small plants, but as the plant gets bigger and more complex, derivation of these intricate equations becomes tedious and prone to error. In addition, the validity of the derived equations is hard to ascertain and leads to ill conditioned optimization problems.

To simplify the process of setting up the optimization problem, the modelling framework has been extended, such that each equipment or header defined in the plant will have the following:

- Input variables (Continuous)
- Output variables (Continuous)
- Status variables (Binary)
- Equation set (Equality/Inequality)
- Variable boundary (Upper/Lower)

Typically both statistical and thermodynamic equations are used for modelling [7]. A similar approach is adopted here. Data from a cogeneration plant in a refinery is used to validate the derived equations. The variables and equations associated with each equipment/header are presented below.

2.3.1 Gas Turbine (GT) Model

The GT model is shown in Figure 2.1, while Table 2.2 gives the GT model variables list.

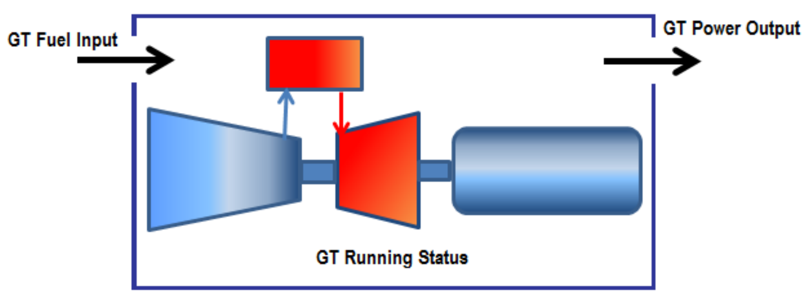


Figure 2.1 - Gas Turbine Model (Optimization)

Table 2.2 - Gas Turbine Variables List

Unique Variable List				
	Description	Variable type	Abbreviation	Unit
Input Variables	GT fuel input	Continuous	fuel_gt	t/h
Output variables	GT power output	Continuous	power_gt	MW
Status Variables	GT running status	Boolean	run_stat_gt	Nil

Equation set for Gas Turbine model

- Power-fuel equality

$$run\_stat\_gt * fuel\_gt = run\_stat\_gt * (IFC * power\_gt + NLF) \qquad (2.1)$$

where IFC - Incremental fuel consumption per MW (t/h/MW)

NLF - No load fuel consumption (T/h)

2.3.2 Heat Recovery Steam Generator (HRSG) Model

The HRSG model is shown in Figure 2.2, while Table 2.3 gives the HRSG model variables list.

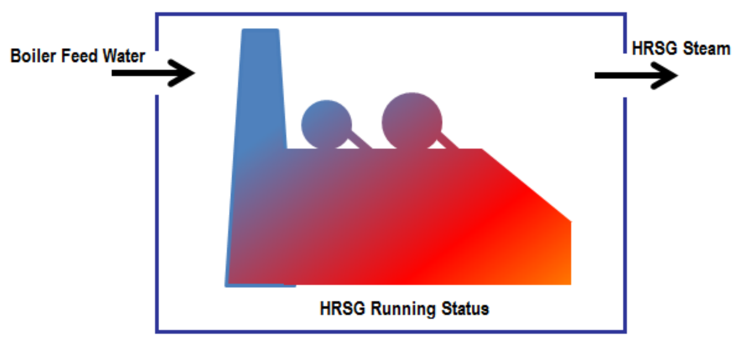


Figure 2.2 - HRSG Model (Optimization)

Table 2.3 - HRSG Variables List

Unique Variable List				
	Description	Variable type	Abbreviation	Unit
Input Variables	Boiler feed water	Continuous	bfw_hrsg	t/h
Output variables	Steam output	Continuous	steam_hrsg	t/h
Status Variables	HRSG running status	Boolean	run_stat_hrsg	Nil

Equation set for HRSG Model

- Power-Steam equality

$$run\_stat\_hrsg * steam\_hrsg = run\_stat\_hrsg * (ISG * power\_gt + NLS) \qquad (2.2)$$

- BFW-Steam equality -

$$run\_stat\_hrsg * bfw\_hrsg \quad = run\_stat\_hrsg * LC \; * steam\_hrsg \qquad (2.3)$$

where ISG - Incremental steam generation per MW (t/h/MW)

NLS - No load steam generation (t/h)

LC - HRSG loss coefficient

2.3.3 Deaerator Model

The deaerator model is shown in Figure 2.3, while Table 2.4 gives the variables list.

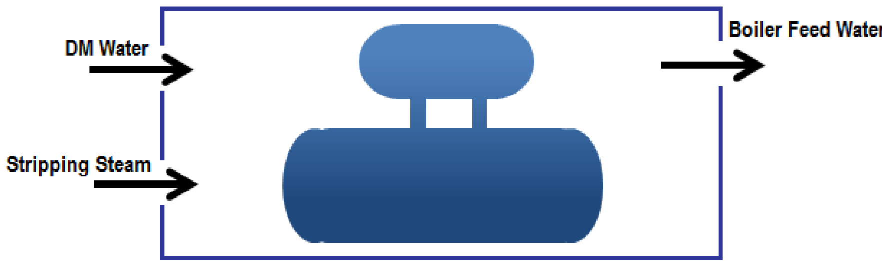


Figure 2.3 - Deaerator Model (Optimization)

Table 2.4 - Deaerator Variables List

Unique Variable List				
	Description	Variable type	Abbreviation	Unit
Input Variables	DM water	Continuous	water_dea	t/h
	Stripping steam	Continuous	steam_dea	t/h
Output variables	Boiler feed water	Continuous	bfw_out_dea	t/h
Status Variables	Nil	Nil	Nil	Nil

Equations set for Deaerator Model

- DM water – BFW equality

$$water\_dea = K * out\_bfw\_dea \qquad (2.4)$$

- Steam input – BFW equality

$$steam\_dea = (1 - K) * out\_bfw\_dea \qquad (2.5)$$

where K is a deaerator constant given by the thermodynamic equation

$$K = \frac{\eta steam - \eta bfw}{\eta steam - \eta dm\_water} \qquad (2.6)$$

where ηsteam - Enthalpy of incoming stripping steam

ηbfw - Enthalpy of boiler feed water

ηdm_water - Enthalpy of incoming DM water

2.3.4 Pressure Reducer De-Superheater (PRDS) Model

The PRDS Model is shown in Figure 2.4, while Table 2.5 gives the model variables list.

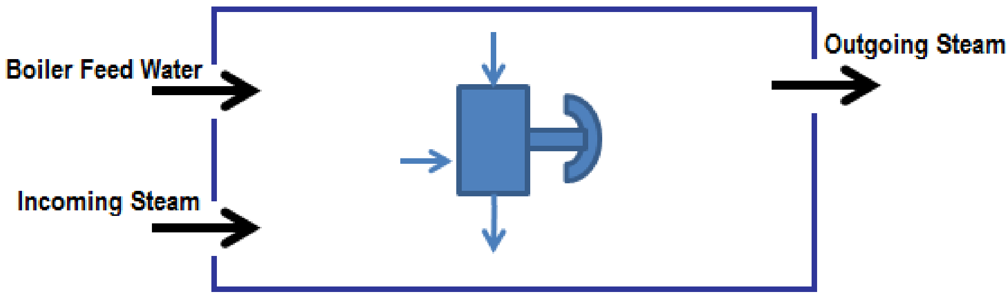


Figure 2.4 - PRDS Model (Optimization)

Table 2.5 - PRDS Variables List

Unique Variable List				
	Description	Variable type	Abbreviation	Unit
Input Variables	Boiler feed water	Continuous	bfw_prds	t/h
	Incoming steam	Continuous	steam_in_prds	t/h
Output variables	Outgoing steam	Continuous	steam_out_prds	t/h
Status Variables	Nil	Nil	Nil	Nil

Equations set for PRDS

- Steam input/ Steam output equality

$$steam\_in\_prds = P * steam\_out\_prds \tag{2.7}$$

- BFW input/ Steam output equality

$$bfw\_prds = (1 - P) * steam\_out\_prds \tag{2.8}$$

where K is a deaerator constant given by the thermodynamic equation

$$P = \frac{\eta steam\_out - \eta bfw}{\eta steam\_in - \eta bfw} \tag{2.9}$$

where ηsteam_in - Enthalpy of incoming steam

ηbfw - Enthalpy of boiler feed water

ηsteam_out - Enthalpy of outgoing steam

2.3.5 Steam Boiler Model

The boiler model is shown in Figure 2.5, while Table 2.6 gives the model variables list.

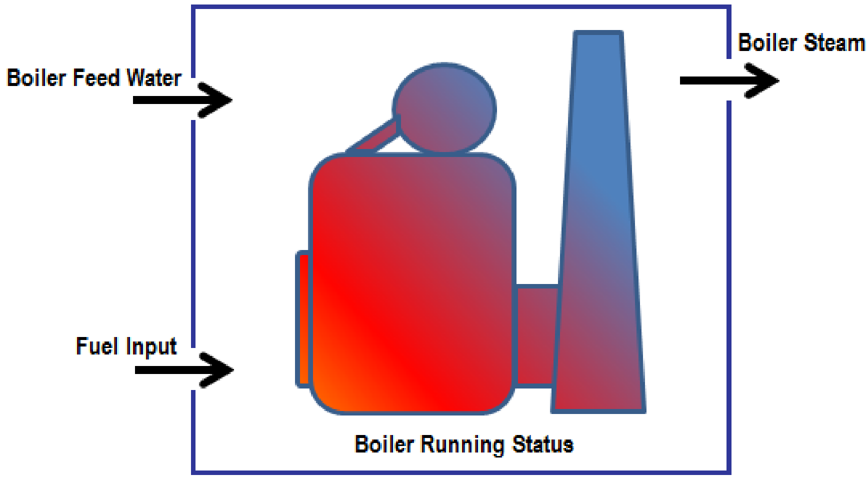


Figure 2.5 - Boiler Model (Optimization)

Table 2.6 - Boiler Variables List

Unique Variable List				
	Description	Variable type	Abbreviation	Unit
Input Variables	Boiler feed water	Continuous	bfw_boiler	t/h
	Fuel input	Continuous	fuel_boiler	t/h
Output variables	Steam output	Continuous	steam_boiler	t/h
Status Variables	Steam Running status	Boolean	run_stat_boiler	Nil

Equations set for Steam Steam

- BFW / Steam equality

$$run\_stat\_blr * bfw\_boiler = run\_stat * LC * steam\_boiler \tag{2.10}$$

- Fuel / Steam equality

$$run\_stat\_blr * fuel\_boiler = run\_stat * \frac{1}{SFR} * steam\_boiler \tag{2.11}$$

where LC - BFW loss coefficient of the steam

SFR - Steam fuel ration of the steam

2.3.6 De-Mineralised (DM) Water Header Model

The DM water header model is shown in Figure 2.6, while Table 2.7 gives the model variables list.

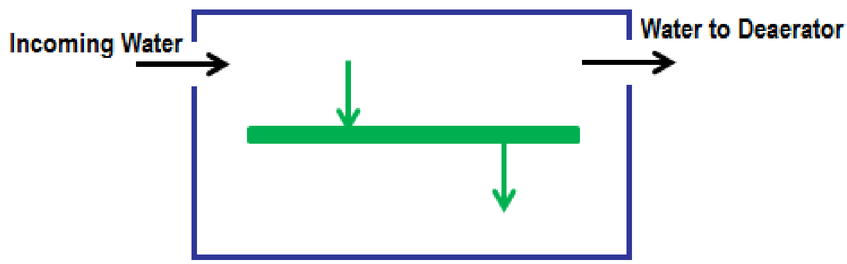


Figure 2.6 - DM Water Header Model (Optimization)

Table 2.7 - DM Water Header Variables List

Unique Variable List				
	Description	Variable type	Abbreviation	Unit
Input Variables	DM water incoming	Continuous	input_water_hdr	t/h
Output variables	Nil	Nil	Nil	Nil
Status Variables	Nil	Nil	Nil	Nil

Equation Set for DM Water Header

- DM water header balance equality

$$input\_water\_hdr = \sum_{i=1}^{nd} water\_dea_i \tag{2.12}$$

where nd - Number of deaerators connected to the water header

2.3.7 Boiler Feed Water (BFW) Header Model

The BFW header model is shown in Figure 2.7, while Table 2.8 gives the model variables list.

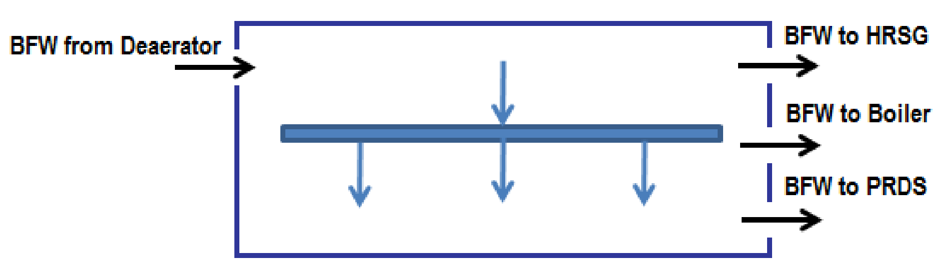


Figure 2.7 - BFW Header Model (Optimization)

Table 2.8 - BFW Header Variables List

Unique Variable List				
	Description	Variable type	Abbreviation	Unit
Input Variables	BFW from deaerator	Continuous	in_bfw_hdr	t/h
Output variables	Nil	Nil	Nil	Nil
Status Variables	Nil	Nil	Nil	Nil

Equation set for BFW Header

- BFW header balance equality -

$$in\_bfw\_hdr = \sum_{i=1}^{nh} bfw\_hrsg_i + \sum_{i=1}^{nb} bfw\_boiler_i + \sum_{i=1}^{np} water\_prds_i \tag{2.13}$$

where nh - Number of HRSG connected to the header

nb - Number of Steams connected to the header

np - Number of PRDS connected to the header

2.3.8 Fuel Header Model

The fuel header model is shown in Figure 2.8, while Table 2.9 gives the model variables list.

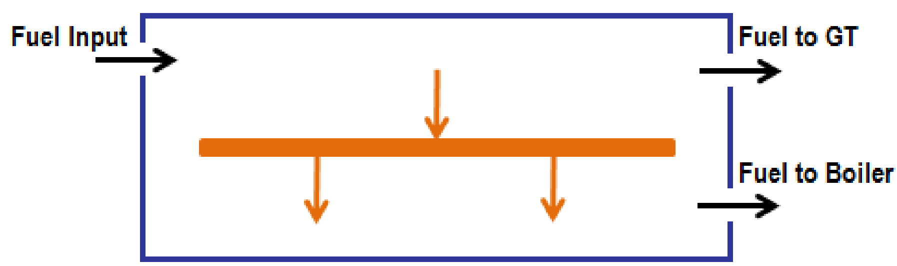


Figure 2.8 - Fuel Header Model (Optimization)

Table 2.9 - Fuel Header Variables List

Unique Variable List				
	Description	Variable type	Abbreviation	Unit
Input Variables	Fuel incoming	Continuous	in_fuel_hdr	t/h
Output variables	Nil	Nil	Nil	Nil
Status Variables	Nil	Nil	Nil	Nil

Equation set for Fuel Header

- Fuel header balancing equality

$$input\_fuel\_hdr = \sum_{i=1}^{ng} fuel\_gt_i + \sum_{i=1}^{nb} fuel\_boiler_i \tag{2.14}$$

where *ng* - Number of GTs connected to the header

nb - Number of steams connected to the header

2.3.9 Steam Header Model

The steam header model is shown in Figure 2.9. The steam header model has no unique variables of its own and all the incoming and outgoing variables are pre-existing.

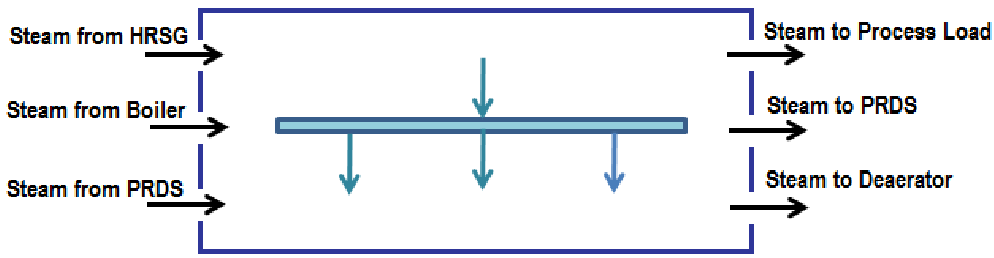


Figure 2.9 - Steam Header Model (Optimization)

Equation set for Steam Header

- Steam header Balance inequality

$$\sum_{i=1}^{nh} run\_stat\_hrsg * steam\_hrsg_i + \sum_{i=1}^{nb} run\_stat\_blr * steam\_boiler_i + \sum_{i=1}^{npo} steam\_prds_i \geq$$

$$steam\_process\_load + \sum_{i=1}^{npi} steam\_in\_prds_i + \sum_{i=1}^{nd} steam\_dea_i \tag{2.15}$$

where nh - number of HRSG connected to the header

nb - number of steams connected to the header

npo - number of PRDS transporting steam out of the header

npi - number of PRDS transporting steam into the header

nd - number of Deaerator connected to the header.

2.3.10 Power Header Model

The power header model is shown in Figure 2.10, while Table 2.10 gives the model variables list.

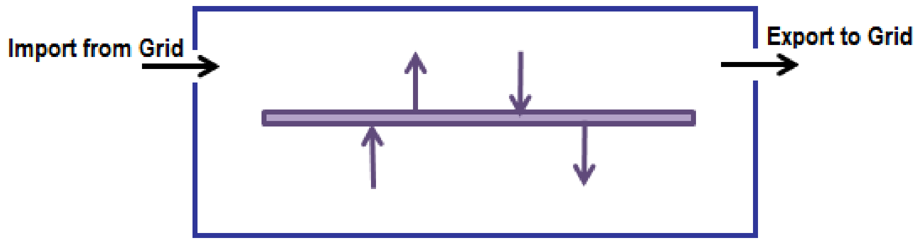


Figure 2.10 - Power Header Model (Optimization)

Table 2.10 - Power Header Variables List

Unique Variable List				
	Description	Variable type	Abbreviation	Unit
Input Variables	Import from grid	Continuous	import_power	MW
Output variables	Export to grid	Continuous	export_power	MW
Status Variables	Nil	Nil	Nil	Nil

Equation set for Power Header

- Power header balancing equality

$$\sum_{i=1}^{ng} run\_gt * \ power\_gt_i + import\_power = export\_power \ + process\_load \quad (2.16)$$

where *ng* - Number of generators connected to the header

2.4 Variables and Constrains in Optimization

The preceding section gives an overall outlook of the optimization models. In the optimization problem, the variables and equations are used as decision variables and constrains, respectively. For example, in the GT model, there are two unique variables and one equation; that forms the two decision variables and one constraint per GT in the optimization routine. Additionally, the models split the operational cost into three headers, namely the power header, fuel header and water header; that represent the power cost, fuel cost and water cost, respectively. This separation helps in automated problem formulation and simplifies the optimization routine. Table 2.4 gives an overview of all the variables and constraints in each of the equipment and header model.

Table 2.11 - Overview of Equipment and Headers

Overview of Variables and Constraints per Equipment/header					
Sl No	Equipment/Header	No of Unique Variables	No of Equalities	No of Inequalities	Cost function bias
1	Gas turbine	2	1	0	Nil
2	HRSG	2 to 4	2 to 4	0	Nil
3	Steam	3	2	0	Nil
4	Deaerator	3	3	0	Nil
5	PRDS	3	2	0	Nil
6	Water header	1	1	0	Water cost
7	BFW header	1	1	0	Nil
8	Steam header	0	0	1	Nil
9	Fuel header	1	1	0	Fuel cost
10	Power header	2	1	0	Power cost

Chapter Three: Optimization Problem

3.1 Problem Formulation

From the modelling equations given in chapter two, the cogeneration optimization problem can be summarized as

Minimize the Cost Function

$$Total\ cost = \sum_{i=1}^{nfh} fuel.costi * input\_fuel\_hdr$$
$$+ \sum_{i=1}^{nwh} DMwater.cost * input\_water\_hdr$$
$$+ \sum_{i=1}^{nph} import.cost_i * import\ power_i - export.cost_i$$
$$* expor\square power_i$$

where nfh - Number of fuel headers

nwh - Number of water headers

nph - Number of power headers

subject to

- GT constraints (2.1)

$$run\_stat\_gt * fuel\_gt = run\_stat\_gt * (IFC * power_{gt} + NLF)$$

- HRSG constraints (2.2), (2.3)

$$run\_stat\_hrsg * steam\_hrsg = run\_stat\_hrsg * (ISG * power_{gt} + NLS)$$

$$run_{stat_{hrsg}} * bfw_{hrsg} = run\_stat\_hrsg * LC * steam\_hrsg$$

- Boiler constraints (2.4), (2.5)

$$run\_stat\_blr * bfw\_boiler = run\_stat\_blr * LC * steam\_boiler$$

$$run\_stat\_blr * fuel\_boiler = run\_stat\_blr * \frac{1}{SFR} * steam\_boiler$$

- Deaerator constraint (2.7), (2.8)

$$water\_dea = K * out\_bfw\_dea$$

$$steam\_dea = (1 - K) * out\_bfw\_dea$$

- PRDS constraint (2.10), (2.11)

$$steam\_in\_prds = P * steam\_out\_prds$$

$$bfw\_prds = (1 - P) * steam\_out\_prds$$

- Water header constraint (2.12)

$$input\_water\_hdr = \sum_{i=1}^{nd} water\_dea_i$$

- BFW header constraint (2.13)

$$in\_bfw\_hdr = \sum_{i=1}^{nh} bfw\_hrsg_i + \sum_{i=1}^{nb} bfw\_boiler_i + \sum_{i=1}^{np} water\_prds_i$$

- Steam header constraint (2.14)

$$\sum_{i=1}^{nh} run\_stat\_hrsg * steam\_hrsg_i + \sum_{i=1}^{nb} run\_stat\_blr * steam\_boiler_i$$

$$+ \sum_{i=1}^{npo} steam\_prds_i \geq$$

$$steam\_process\_load + \sum_{i=1}^{npi} steam\_in\_prds_i + \sum_{i=1}^{nd} steam\_dea_i$$

- Fuel header constraint (2.15)

$$input\_fuel\_hdr = \sum_{i=1}^{ng} fuel\_gt_i + \sum_{i=1}^{nb} fuel\_boiler_i$$

- Power header constraint (2.16)

$$\sum_{i=1}^{ng} run\_gt * power\_gt_i + import\_power = export\_power + process\_load$$

3.2 Characteristics of the Optimization Problem

From the problem formulation in section 3.1 it is observed that the objective function is linear in nature. However, the constraints are both integer (running status) and real (GT power, boiler steam etc.). In addition, product of integer and real variables exists that turns the problem into a mixed integer nonlinear programming problem (MINLP). The general form of MINLP is as follows:

$$\begin{cases} \text{minimize } f(x,y) \\ \text{subject to } C(x,y) \\ x \in X, y \in Y \text{ integer} \end{cases} \tag{3.1}$$

In equation 3.1, x represents the real continuous variables like boiler and generator load set points, while y represents the integer binary values used for running status indication.

MINLP problems are difficult to solve, because they combine the difficulties of both of their subclasses; the combinatorial nature of mixed integer programs (MIP) and the difficulty in solving non-convex (and even convex) nonlinear programs (NLP). Subclasses MIP and NLP belong to the class of theoretically difficult problems (NP-complete) [5]. NP complete means that the solution can be verified in polynomial time but there are no algorithms to compute the solution in the same time frame.

3.2.1 Algorithms for MINLP problems

Outer Approximation (OA) methods [22]-[23], Branch-and-Bound (B&B) [24]-[25], Extended Cutting Plane methods [26], and Generalized Bender's Decomposition (GBD) [27] for solving MINLPs have been discussed in the literature since the early 1980's. These approaches generally rely on the successive solutions of closely related NLP problems. In addition, OA and GBD require the successive solution of a related MIP problem. Thus solvers for MINLP are mostly built by combining LP, MIP and NLP solvers.

Due to the high complexity of MINLP and the wide range of applications that can be modelled as MINLPs, it is sometimes desirable to customize the MINLP solver for a specific application in order to achieve good computational performance [28]-[29]. In this thesis the above paradigm is adopted. A custom three stage linear programming approach is used where the LP relaxed problem is successively solved in three stages to get a final solution

3.3 Solving the Optimization Problem

The optimization problem listed in section 3.1 provides the formulation to solve the problem for lowest possible operating cost. However two more scenarios that are quite common in plant operation are:

- Achieving reliable power and steam
- Handling of islanded power and steam headers during plant maintenance activities.

The solution to the above is also addressed in the algorithms developed in this thesis.

3.3.1 Minimizing Operation Cost

For determining the minimum operating cost, a novel three stage linear programming algorithm is used. It is based on the following three important facts:

- If it is possible to set equipment lower boundaries to zero and run the optimization; then the optimum value of redundant equipment output will always go to zero
- If the optimum value of an equipment output variable is zero then it is no longer necessary to run the equipment and it can switched off
- The turning on and turning off of any equipment can be mapped in linear programming by pre-multiplying all the equations (equalities/inequalities) and variables associated with the equipment with the running status variable.

The algorithm can be summarized as below:

- Stage-0

1. Read data from plant data file and construct the system matrices.

2. Run the optimization with the above system matrices. This scenario is to see if the problem is feasible with all equipment turned ON.

- If feasible then proceed to step 3, else system is unfeasible; terminate.

- Stage-1

3. Modify the system matrices such that:

- All the lower boundaries of the equipment are set to zero.
- HRSG steam production is approximated to eliminate the no load steam production steam.

4. Run the optimization with the modified system matrices.

- If feasible proceed to stage-2, else solution in stage 0 is the optimum; terminate.

- Stage-2

5. If any of the turbines is producing 0 MW, turn OFF the generator by setting the status to zero.

6. Modify the system matrices such that:

   - The generation sources which are turned OFF are taken out of the system matrices.

   - Set the lower boundaries of the steam generation sources to zero.

7. Run the optimization with modified system matrices.

➢ Stage 3

8. If any of the steam sources is producing 0 t/h, turn OFF the source by setting running status to 0.

9. Modify the system matrices such that:

   - The steam sources that are turned OFF are taken out of the system matrices.

   - Set the lower boundaries to plant data specifications.

10. Run the optimization to get to the best possible solution.

Flowchart for the three stage optimization algorithm is shown in Figure 3.1.

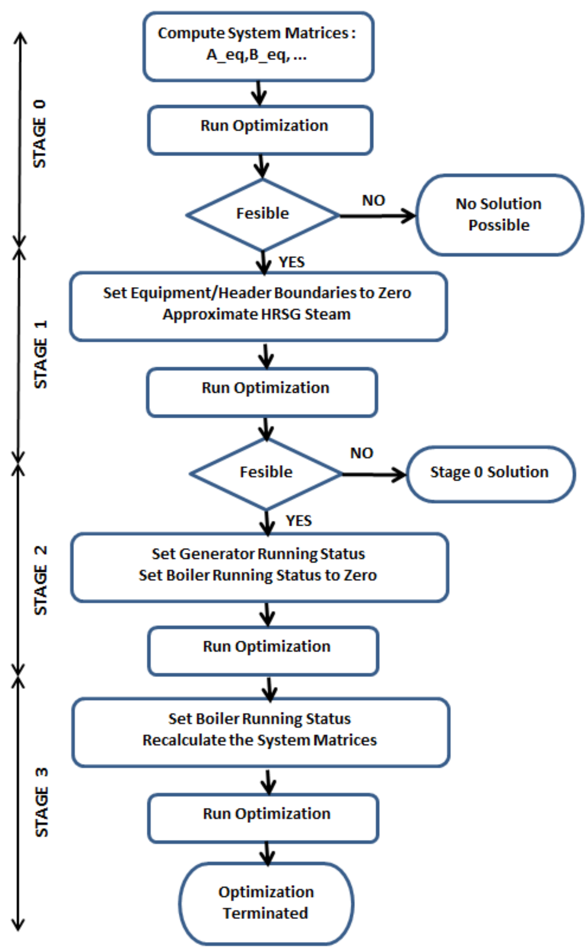


Figure 3.1 - Three Stage Optimization Algorithm

From the flowchart in Figure 3.1 it can be observed that the algorithm avoids the integer variables in the optimization problem by separately choosing them outside the optimization routine. In stage 2 the generator running status variables are decided in accordance with the optimization results from stage 1 and in stage 3 the steam sources status variables are decided in accordance with the optimization results from stage 2. Thus before the optimization run in stage 3, the integer variables that give the best possible solution are already known, thus addressing the problem of selection of equipment. Now that it is known which equipment has to be run, the problem of how to run the selected equipment is found by solving the optimization step in stage 3. The output of stage 3 gives the setpoint of the generators and the steam sources that will give the minimum possible cost.

3.3.2 Reliable Power and Steam

The primary focus of reliable operation is to ensure that the downstream units are not affected due to an unexpected trip of a generator or boiler. Most of the cogeneration units follow the n+1 concept to achieve this objective; i.e. for a block load of N there is always n+1 generation capacity available [30]. This concept is usually applied to both steam and power networks.

Both the power and steam reliability calculations are incorporated into the three stage linear programming algorithm. The power reliability is an extension of stage 1 while steam reliability is implemented as an extended calculation of stage 2.

3.3.2.1 Reliable Power

For achieving reliable power, it is necessary to sustain the system load in case of a generator failure. In other words there should be sufficient capacity in the system to supply an additional load equal to the largest generating source apart from the usual process load. This forms the basis for the power reliability algorithm, which can be summarized as:

1. Check for number of generating sources in the power header:
    - If greater than one, then proceed to step 2, else reliable power cannot be achieved.
2. Find the largest generating source in the header and add the value to the process power load.
3. Run the optimization with the increased power load:
    - If the optimization is feasible, then proceed to step 4, else reliable power cannot be achieved.
4. Run stage 1 with increased power load so that one more additional generator gets selected.

5. Complete the three stage optimization process to get reliable power optimized setpoint values.

Flowchart for the power reliability algorithm discussed above is shown in Figure 3.2.

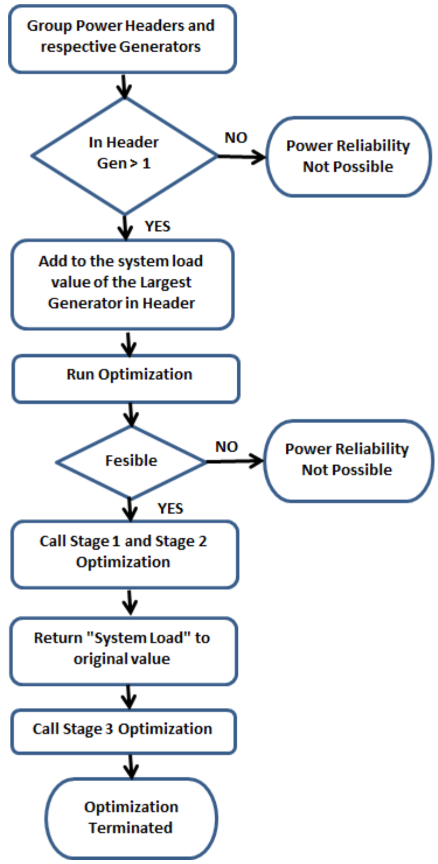


Figure 3.2 - Power Reliability Algorithm

3.3.2.2 Reliable Steam

Achieving steam reliability while optimizing cost at the same time is a challenging task due to the presence of multiple steam headers and associated header interconnections. In addition, the following facts add complexity to the steam reliability problem:

- It is not known how much steam production is necessary prior to the optimization run. This is due to the presence of auxiliary steam consumption equipment like deaerators.
- Steam production in steam sources like HRSG cannot be independently controlled and is dependent on the GT power production.
- PRDS that facilitates the interconnection of headers can supply only from a higher pressure header to a lower pressure header, and not the other way round.

The steam reliability algorithm tackles the above constraints is summarized as follows:

1. Arrange the steam headers in the plant from the highest order to the lowest.

2. Calculation for the header with the highest pressure:

2.1 Check the number of sources:

- If greater than one proceed to step 2.2;
- Else reliable power cannot be achieved in the header, proceed to step 3.

2.2 Find the largest generating source in the header and add the value to the process steam load.

2.3 Run the optimization with the increased process steam load:

- If the optimization is feasible; reliable power is achieved in header, keep the increased process load.
- Else reliable power cannot be achieved in header; change the steam load to normal process steam load.

3. Calculation for header with lower steam pressures:

3.1 Check the number of sources

- If greater than one proceed to step 3.2;
- Else check for interconnection to the header:
  - If interconnection is reliable then reliable steam is achieved.
  - Else reliable steam is not achieved in the header.

3.2 Find the largest source in the header and the steam from reliable steam headers.

3.4 Calculate the reliable steam needed in the header as the sum of the process steam load, the largest source and the negative of steam available from reliable interconnections.

3.5 Run the optimization with the increased steam load:

- If the optimization is feasible, then reliable power is achieved in header, keep the increased steam load.
- Else reliable power cannot be achieved in the header, change steam load back to normal steam load.

3.6 Check if the header is the last steam header:

- If yes, proceed to step 4; else go back to step 3.

4. Run the stage 2 optimization with the increased steam load so as to select additional steam generation source.

5. Continue with the three stage optimization to get reliable steam optimized values.

Flowchart for the steam reliability algorithm discussed above is shown in Figure 3.3.

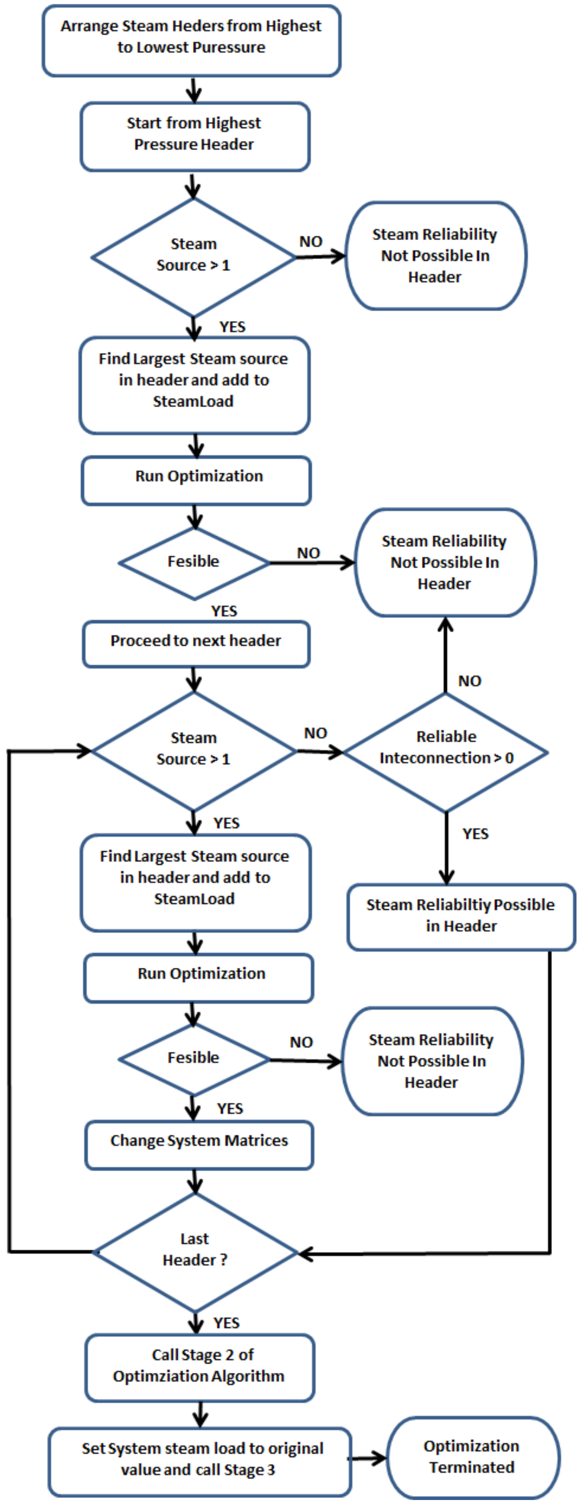


Figure 3.3 - Steam Reliability Algorithm

3.3.3 Islanded Power and Steam Headers

The presence of islanded power and steam headers is quite common in cogeneration systems and they arise due to the following factors:

- Plant emergency conditions like feeder outage.
- Performance of regular maintenance activities.
- Presence of an operation philosophy that calls for separate power and steam headers.

The three stage optimization and the modelling allow significant flexibility for modelling and detecting islanded power and steam headers. The important step is to list the islanded headers as separate and individual headers. The algorithm can be summarized as follows:

1. List the islanded plant headers as separate headers.
2. Construct the system matrices for the islanded headers.
3. Augment the system matrices to the main plant system matrices.
4. Run the three stage optimization with the augmented plant matrices:
    - If the optimization is feasible then it is the global minimum.
    - If not then the islanded header is unstable and will face a power or steam collapse.

The islanded header algorithm discussed above is given in Figure 3.4.

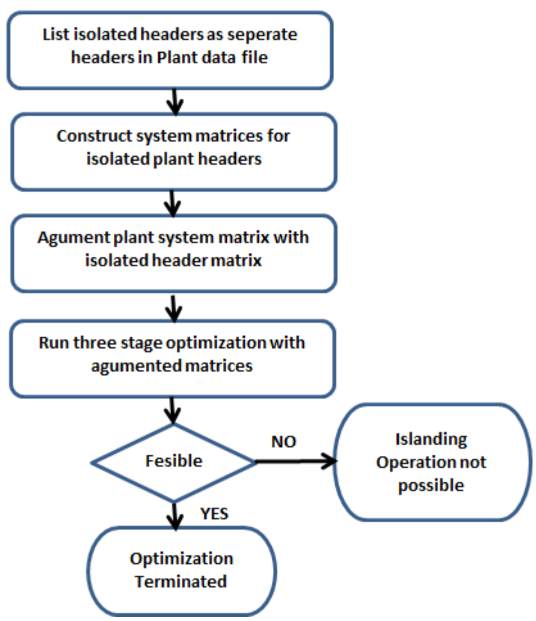


Figure 3.4 - Islanded Headers Algorithm

Chapter Four: Optimization of Cogeneration Systems

4.1 The Model Plants

To test the effectiveness of the three stage algorithm, a black box solver MIDACO [31]-[33] has been used to compare the results. MIDACO uses ant colony optimization and oracle penalty technique to solve MINLP problems [32]-[33]. The models chosen represent cogeneration systems in:

- A refinery complex
- District heat and power utility complex

The three stage algorithm was tested on both models, while MIDACO was tested only on the refinery complex. MIDACO was not tested on the district heat and power complex due to high variable count. The complete plant details along with the plant and equipment specifications are provided in Appendices A and B. Tables 4.1 – 4.2 give a brief overview of the equipment/headers present in the model plants. Table 4.3 shows the nature and size of the optimization problem presented by each plant.

Table 4.1 - Equipment and Headers in Plant Models

Equipment/Header	Refinery cogeneration	District heat and Power utility
Gas Turbine	3	30
HRSG	3	30
Boiler	3	35
Deaerator	2	10
PRDS	3	15
Water header	1	5
BFW header	3	15
Steam header	3	7
Fuel header	2	10
Power header	1	1

Table 4.2 - Power and Steam Loads in Model Plants

Generation & Average load	Refinery cogeneration	District heat and Power utility
Installed power capacity	66 MW	1050 MW
Installed Steam capacity	519 t/h	7650 t/h
Average Power load	40 MW	750 MW
Average Steam load	400 t/h	5600 t/h

Table 4.3 - Optimization Characteristics of Model Plants

Optimization Characteristics	Refinery cogeneration	District heat and Power utility
Total No of variables	56	397
Total no of continuous variables	50	332
Total no of integer variables	6	65
Total no of equations	43	255
Total no of equalities	40	248
Total no of Inequalities	3	7

4.2 Optimization Scenarios

The following operational scenarios have been simulated in the model plants and the optimization in the operational scenarios was done using both the three stage optimization algorithm and MIDACO. A brief overview of the operational scenarios is given below:

- Normal Operation

  In this case a feasible operation point is chosen that satisfies all the operational and equipment constraints. This normal operation cost is the base reference with which the cost benefits are calculated.

- Optimization with all known running equipment

  In this case all the equipment that is running is known beforehand. This operational scenario is essentially a linear programming problem as there is no equipment selection involved and serves as the basis for showing the effectiveness of equipment selection.

- Optimization for lowest cost

  In this scenario, the optimization also involves equipment selection. This task is performed in the two test plants as below:

- Refinery Cogeneration System: The system selects the generators and the boilers to be run such that the operational cost is the minimum while satisfying power and steam demand simultaneously.

- District Power and Utility: The system selects the generators and the boilers to be run in the 5 power plants in the utility such that the operation cost is the lowest, while satisfying power and steam demand simultaneously.

- Optimization under power and steam reliability

  In this scenario, the optimization takes into account the reliable power and steam criterion. This task is performed in the two test plants as below:

  - Refinery Cogeneration System: Tripping of any one boiler or generator should not affect the critical power and steam loads in the process plants.

  - District Power and Utility: Tripping of any one generator or boiler should not cause any disruptions in power and steam supply to the customers.

- Optimization under islanded power and steam headers

  In this scenario, the optimization takes into account the islanded power and steam headers in the system that supply power and steam to a local load. This task is performed in the two test plants as below:

  - Refinery Cogeneration System: It is assumed that for maintenance reasons the electrical system is islanded in such a way that GT-1 and 2 form one power island while GT-3 alone forms another island. Similarly, the MP steam header is divided into two, with HRSG 1, 2 and 3 supplying on an islanded header while, Boilers 1, 2 and 3 feed into a separate steam header.

  - District Power and Utility: It is assumed that for maintenance reasons one Power block, consisting of Plant 5 is islanded from the rest of the system and is catering to its own power and steam load.

4.3 Optimization Results for Three Stage Algorithm

Table 4.4 shows the equipment selection results while Table 4.5 shows the cost reduction achieved in the model plants with the three stage optimization algorithm.

Table 4.4 - Equipment Selection with Three Stage Optimization Algorithm

Scenario	Equipment	Refinery Cogeneration System		District Heat and Power Utility	
		Turned ON	Turned OFF	Turned ON	Turned OFF
Normal Operation	GT/HRSG	3	0	30	0
	Boiler	3	0	35	0
Optimised with all equipment running	GT/HRSG	3	0	30	0
	Boiler	3	0	35	0
Optimized for minimum cost	GT/HRSG	2	1	20	10
	Boiler	3	0	19	16
Optimized for power and steam reliability	GT/HRSG	3	0	21	9
	Boiler	3	0	21	14
Optimization under islanded condition	GT/HRSG	3	0	20	10
	Boiler	3	0	22	13

Table 4.5 - Cost Reduction with Three Stage Optimization Algorithm

Scenario	Cost factors	Refinery Cogeneration System	District Heat and Power Utility
Normal Operation	Cost of operation	$ 55746.42	$ 619792.63
Optimized with all equipment running	Cost of operation	$ 54176.89	$ 591405.52
	Cost saving/year	$ 13.75 million	$ 248.66 million
	% Reduction	2.89%	4.79%
Optimized for minimum cost	Cost of operation	$ 53161.29	$ 577713.01
	Cost saving /year	$ 25.04 million	$ 368.61 million
	% Reduction	4.86%	7.28%
Optimized for power and steam reliability	Cost of operation	$ 54176.89	$ 578966.42
	Cost saving /year	$ 13.75 million	$ 357.63 million
	% Reduction	2.89%	7.05%
Optimized under islanded condition	Cost of operation	$ 54367.64	$ 582501.96
	Cost saving /year	$ 12.01 million	$ 326.26 million
	% Reduction	2.52%	6.42%

4.4 Optimization Results for MIDACO

Table 4.6 shows the equipment selection results while Table 4.7 shows the cost reduction achieved in the model plants with MIDACO.

Table 4.6 - Equipment Selection with MIDACO

Scenario	Equipment	Refinery Cogeneration System	
		Turned ON	Turned OFF
Normal Operation	GT/HRSG	3	0
	Boiler	3	0
Optimised with all equipment running	GT/HRSG	3	0
	Boiler	3	0
Optimized for minimum cost	GT/HRSG	2	1
	Boiler	3	0
Optimized for power and steam reliability	GT/HRSG	3	0
	Boiler	3	0
Optimization under islanded condition	GT/HRSG	3	0
	Boiler	3	0

Table 4.7 - Cost Savings with MIDACO

Scenario	Cost factors	Refinery Cogeneration System
Normal Operation	Cost of operation	$ 55746.42
Optimized with all equipment running	Cost of operation	$ 54176.89
	Cost saving/year	$ 13.75 million
	% reduction	2.89%
Optimized for minimum cost	Cost of operation	$ 53387.61
	Cost saving/year	$ 20.66 million
	% reduction	4.41%
Optimized for power and steam reliability	Cost of operation	$ 54182.37453
	Cost saving/year	$ 13.7 million
	% reduction	2.83%
Optimized under islanded condition	Cost of operation	$ 56867.5723
	Cost saving/year	$ -9.82 million
	% reduction	-1.9%

4.5 Analysis and Observations

It can be observed that even optimization without equipment selection brings significant cost benefits; a 2.89% reduction in operation cost is seen in the refinery cogeneration unit while a 4.79% reduction is seen in the district heat and utility system. As compared to the 7-8% savings in [2], the savings here average about 4% as the reference point taken for comparison is a near optimal condition. In this comparison the three stage linear programming and MIDACO perform at par, with both methods reaching similar solutions.

Equipment selection brings further cost benefits. In the refinery cogeneration unit, with one GT shut down the reduction in cost is about 4.86% that equates to $ 25 million in cost reduction per year. The benefit is even more significant in the district heat and utility system. With shutting down of the 10 GTs and 16 boilers, a cost reduction of 7.8 % is achieved that equates to $ 368 million in savings annually. In this scenario of minimum cost, the three stage algorithm performed better than MIDACO. MIDACO's optimum solution yields a reduction of only 4.41% in the refinery cogeneration system as compared to 4.86% achieved by the three stage algorithm.

The additional constraint of reliability as expected raises the cost of operation. In the refinery cogeneration system, one generator that was shut down was turned ON again to achieve power reliability, but steam reliability could not be achieved due to lack of surplus capacity. In the district heat and utility system, one generator and two boilers were additionally turned ON to meet the power and steam reliability criteria. The cost benefits for the refinery cogeneration system came down to 2.89% that equates to $ 13.75 million in annual savings. The cost benefit for the district power and utility system also came down to 7.28% that equates to $ 357 Million in annual savings.

In the islanded power and steam header case, operation costs are highest in the simulated scenarios. This is due to the inability of the system to utilize islanded generators and boilers that had to supply a local load and could not deliver power or steam to other parts to the system. In the refinery cogeneration system all generators and boilers had to be turned on to support the islanded operation and this along with additional constraints reduced the cost benefit to 2.52%. This equates to $ 12 million in annual savings. In the district heat and utility system, one GT and three boilers were additionally turned ON to facilitate the islanded operation. This eroded the savings to 6.42% that equates to $ 326 million in annual savings.

It is also noted that the three stage algorithm arrives at solutions quickly; averaging less than 2 s for refinery cogeneration system and less than 4 s for district heat and utility system, MIDACO on the other hand took 1000-5000 s for the refinery cogeneration system. This is mainly due to the fact that the three stage linear programming uses both interior point method and sparse matrices to speed up the calculation. As the integer variables are independently calculated, this removes the additional overhead and speeds up the solution time. MIDACO being a black box solver calculates both the integer and continuous variables simultaneously. In addition, MIDACO has to stochastically generate all values that will not only give a feasible solution but will also reduce the function cost. Combination of the above factors gives the three stage algorithm a distinct edge over blackbox solvers such as MIDACO.

Chapter Five: Approach to Control Modelling

Developing a mathematical model of the process is often the first step in control design. Though the design of the fuzzy controller does not depend on internal process models, the models are important tools for the simulation and testing of the fuzzy controller in an offline environment. It is a well known fact that first or second order plus dead time model may in general represent process dynamics [34]. System identification tests used to develop these models can be broadly classified into parametric and non-parametric approaches. Transfer function model is one of the commonly used parametric models; the same is used here. To obtain the transfer function model, both open loop and closed loop tests can be performed in the plant equipment. If the process exhibits oscillatory tendencies to a step input, then a closed loop test is done; otherwise an open loop test is preferred for its simplicity.

5.1 Modelling Limitations

Rather than a full exhaustive dynamic model, relationship between select inputs and outputs in the given system are chosen. This can be observed in the boiler model, where a full exhaustive model will require relationships to be established between four control loops. The air control loop consists of inlet air flow, air pressure, air temperature, oxygen level, flue gas temperature and flue gas pressure as variables. The feed water control loop consists of feed water pressure, feed water flow, drum level and drum pressure as variables. Attemperator control loop consists of superheater stage-1 temperature, superheater stage-2 temperature and attemperator feed water flow as variables. Fuel control loop consists of fuel flow, steam flow as variables. In this study only the fuel control loop is considered as the control only affects this loop. Similar choices are made on all the control models.

5.2 Modelling Framework

A modelling framework has been developed to aid the development of the control models for this thesis. The modelling framework has two salient features

- Well defined inputs and outputs: The total plant is made up of a number of equipment and subsystems and each of them are modelled separately. By having well defined inputs and outputs it becomes easier to interconnect them and build the complex plant in the simulation environment. This can be observed in the gas turbine model and HRSG Model.

In the gas turbine model the GT power is an output and in the HRSG model the GT power is an input, thus the two models can be easily interconnected to from the combined GT-HRSG system. In addition having well defined plant models for equipment facilitates reuse of models with changes made only to the specific equipment characteristics.

- Synchronized characteristics: Though the plant models are dynamic in nature the steady state nature of the plant models is similar to the optimization models. This can be observed in the boiler models used for optimization and boiler models used for simulation. If in the optimization model the boiler fuel consumption is 10 t for 140 t/h of steam then the boiler dynamic model will consume the same during steady state conditions. This synchronized characteristic aids accurate real time simulation of the system in conjunction with the optimization of the plant.

To represent the measurement noise in transducers, white noise is added to the above mentioned models. This helps in simulating the control schemes close to the real plant conditions and helps to ascertain the robustness of the developed control system. Appendix D gives an overview of how the plant dynamic data was used to formulate the GT-HRSG dynamic model for simulation.

5.3 Plant Equipment Modelling

Based on the above framework six different kinds of equipment and headers have been modelled. The simulation environment chosen for the project is SIMULINK. The equipment and header models described here form the building block of the larger plant model that can be seen as a collection of interconnected equipment models.

5.3.1 Gas Turbine Model

Figure 5.1 shows the block diagram of the gas turbine model. Inputs to the GT dynamic model are the GT power setpoint and GT initial conditions while the outputs are GT power and GT fuel consumption. Internally the GT power setpoint is routed to a rate limiter that controls the rate at which the setpoint is incremented or decremented. This rate limiter thus helps in the smooth ramping up or ramping down of the power output in the GT. The output from the rate limiter then acts as the input to a saturation block that limits the maximum and minimum power output from the GT. This is then routed to a first order block with dead time that simulates the governor and the turbine characteristics. The output of this block is the GT power in real time. The block has a “GT characteristic" module where all the process parameters are stored. It is also used to achieve synchronization with the optimization model.

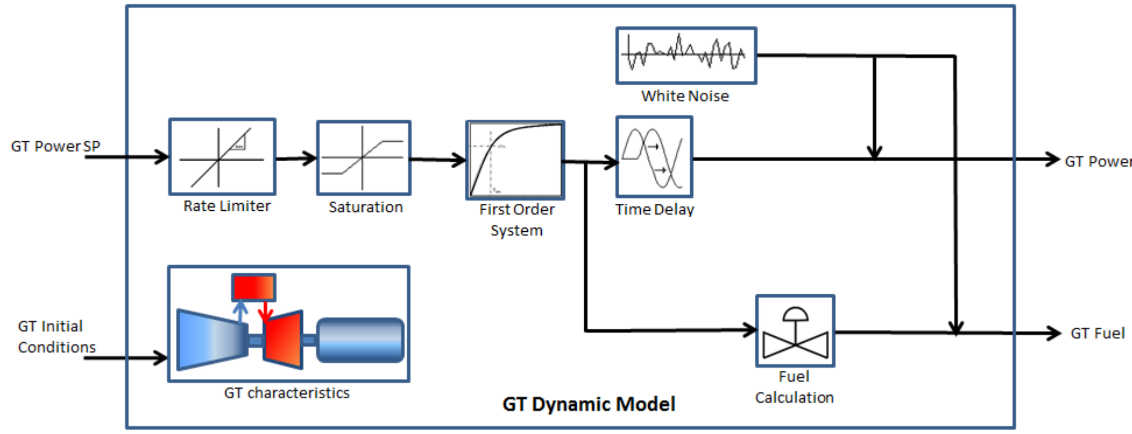


Figure 5.1 - Gas Turbine Model (Dynamic)

Figure 5.2 shows the GT power setpoint and the GT power over a 200 s time interval. The GT setpoint increases from 10 MW to 25 MW at time 15 s and remains constant during the entire simulation. It can be observed that the GT power output begins to rise immediately and ramps up smoothly to 25 MW in about 100 s; a ramp up rate of about 15 MW/m.

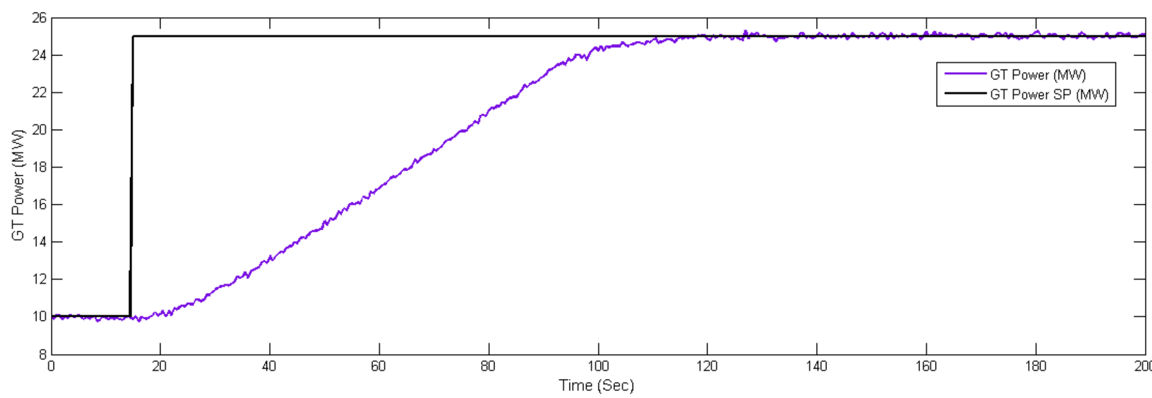


Figure 5.2 - GT Output Characteristics

5.3.2 Heat Recovery Steam Generator (HRSG) Model

Figure 5.3 shows the block diagram of the HRSG model. Inputs to the HRSG dynamic model are GT power and HRSG initial conditions while the outputs are HRSG steam and HRSG BFW consumption. Internally the GT power is routed to a damper control block that converts the GT power to a HRSG steam output value. By default, if no damper angle is specified then damper is assumed to be fully open to achieve maximum efficiency. The output of this block is then given to a saturation block that limits the maximum and minimum possible output from the HRSG. It is then routed to a second order system with dead time that simulates the HRSG dynamics. The block has a “HRSG characteristic” module where all the process parameters are stored and helps in achieving synchronization with the optimization models.

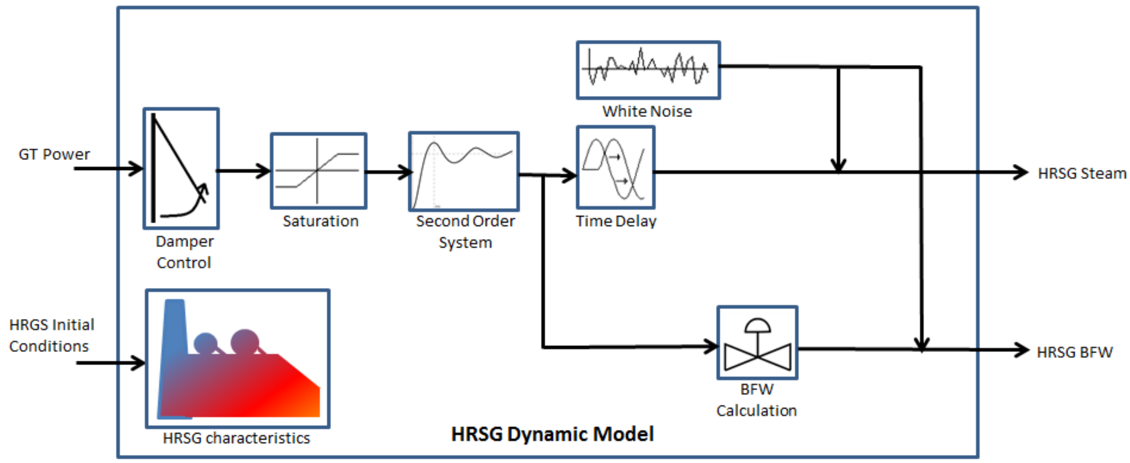


Figure 5.3 - HRSG Model (Dynamic)

Figure 5.4 shows the HRSG steam output when the GT power changes from 10MW to 15 MW. The output of the damper control block converts this GT power change into a steam setpoint change; from 25 t/h to 33 t/h. This input when fed into the second order system with dead time graph gives the actual steam rise dynamics. The HRSG HP Steam graph shows how the steam output rises from 25 t/h to 33 t/h in the course of 130 s. It can be observed that the output begins to rise only after about 5 - 6 s after the input is given. This delay can be attributed to the thermal capacitance of the HRSG unit.

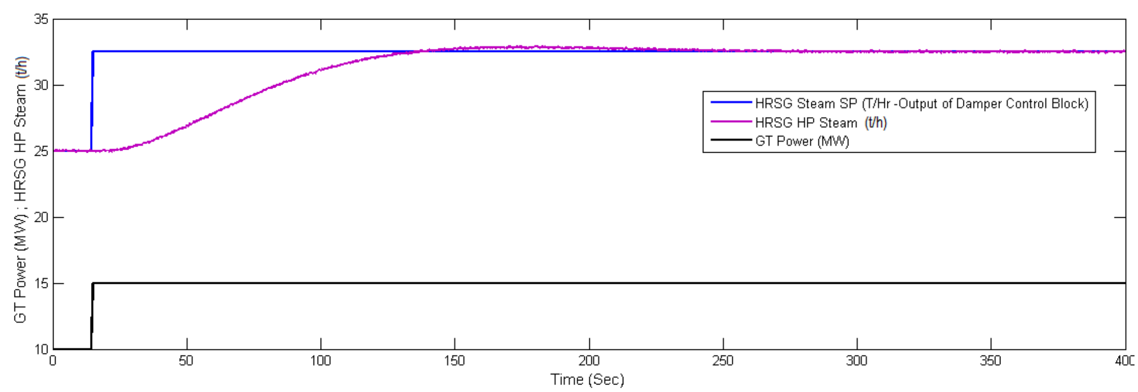


Figure 5.4 - HRSG Output Characteristics

5.3.3 Boiler Model

Figure 5.5 shows the block diagram of the boiler model. Inputs to the boiler model are boiler steam setpoint and boiler initial conditions while the outputs are boiler steam, boiler fuel consumption and boiler BFW consumption. Internally the boiler steam setpoint is routed to a rate limiter that controls the rate at which the boiler output can be changed safely. Boilers have high thermal capacitance. Thus fuel should be incremented at a controlled rate so as to achieve a smooth upswing and downswing of the boiler output. The rate limiter helps in achieving the same. The output of the rate limiter is fed to the saturation block that limits the maximum and minimum output from the boiler. Next the signal is given to a second order system with dead time block that simulates the boiler dynamics. The output of the block is the boiler steam. The module also has a "Boiler characteristic" block that is used for synchronizing with the optimization model.

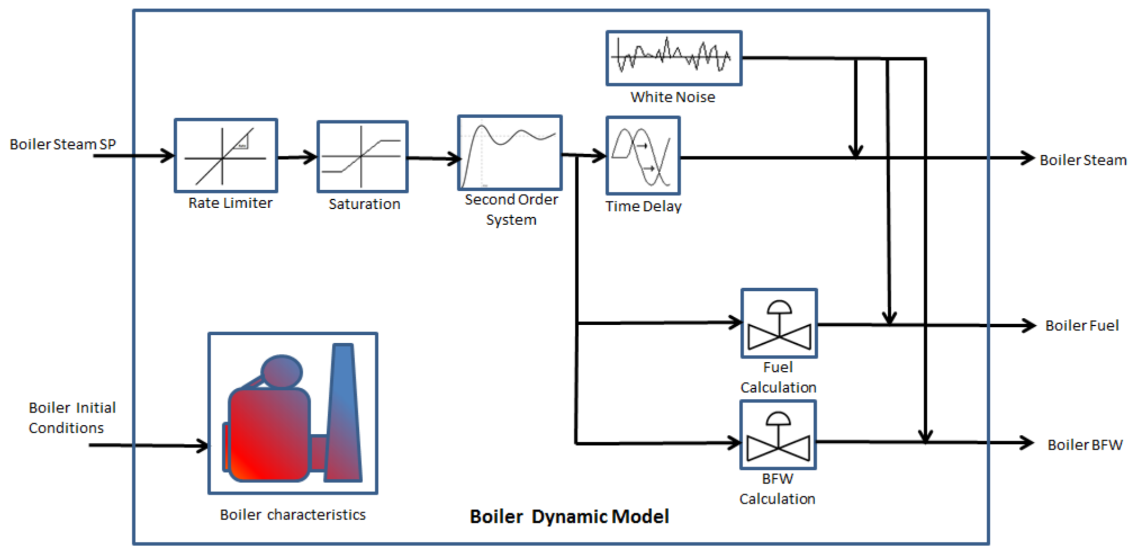


Figure 5.5 - Boiler Model (Dynamic)

Figure 5.6 shows the boiler steam output responding to a boiler steam setpoint. In the simulation the boiler setpoint is changed from 60 t/h to 90 t/h and the steam gradually rises to the setpoint in about 200 s; an average increase of 10 t/h every minute. It can be seen that the swing up has a very flat profile. This can be attributed to the rate limiter in the model. The output begins to rise a good 20 s after the input is given. This is because in an oil fired boiler the air is incremented first and then the fuel. In addition the huge thermal capacitance of the boiler also contributes to this delay.

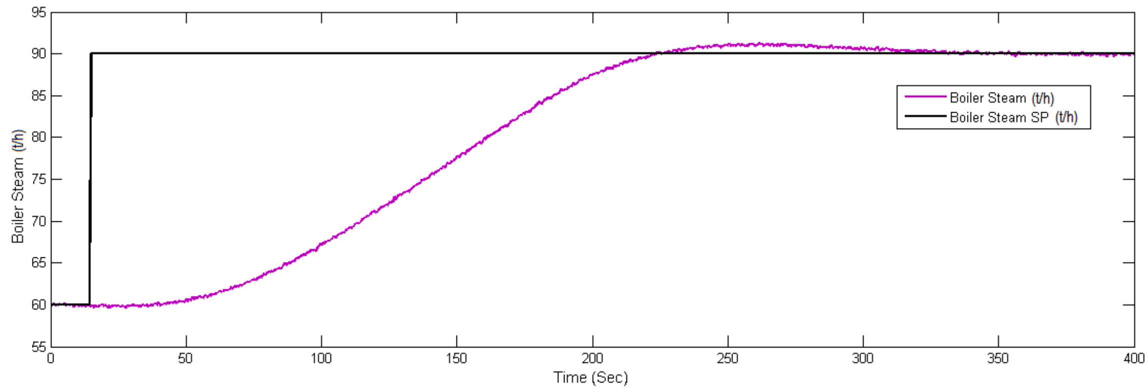


Figure 5.6 - Boiler Output Characteristics

5.3.4 Pressure Reducer De-Superheater (PRDS) Model

Figure 5.7 shows the block diagram of the PRDS model. The inputs to the model are PRDS steam setpoint and PRDS initial condition while the outputs are PRDS downstream steam, PRDS upstream steam consumption and PRDS BFW consumption. Internally the PRDS steam setpoint is routed to a saturation block that limits the maximum and minimum steam output for the PRDS. This signal is then fed to a second order system with dead time block that simulates the process dynamics of the PRDS. The output of this block is the PRDS downstream steam. The PRDS upstream steam and PRDS BFW are calculated using known thermodynamic equations from the "PRDS characteristic" module.

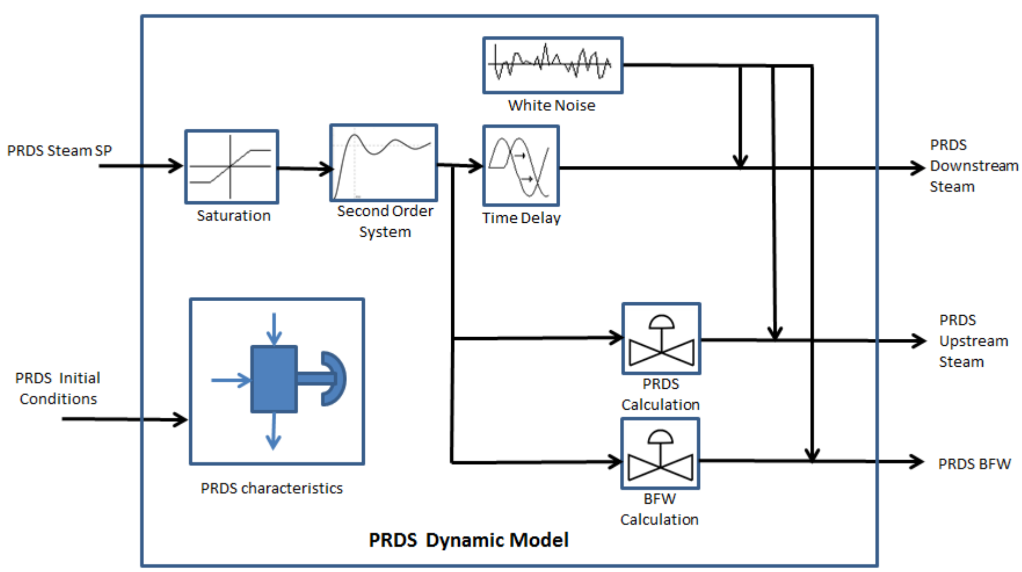


Figure 5.7 - PRDS Model (Dynamic)

Figure 5.8 shows the PRDS downstream steam v/s the PRDS steam setpoint. In the figure the PRDS steam set point is increased from 10 t/h to 40 t/h and the downstream steam rises to the setpoint in about 40 s and settles in about 120 s. The characteristic is a typical second order with dead time as there is no rate limiter in the model.

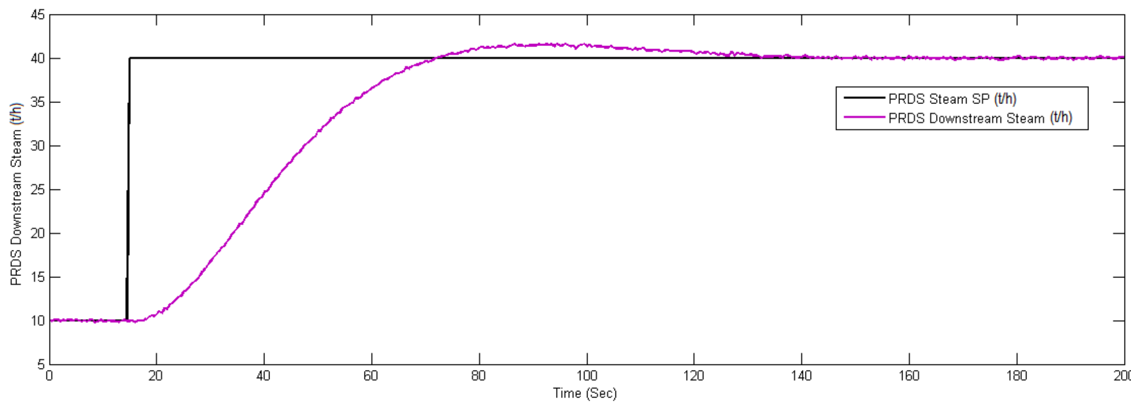


Figure 5.8 - PRDS Output Characteristics

5.3.5 Deaerator Model

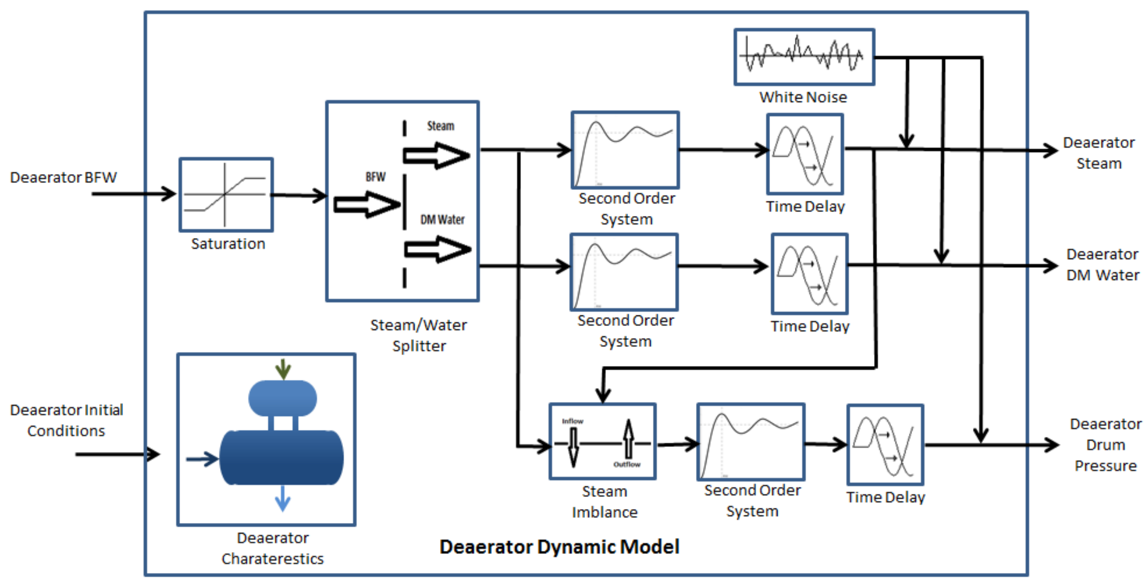
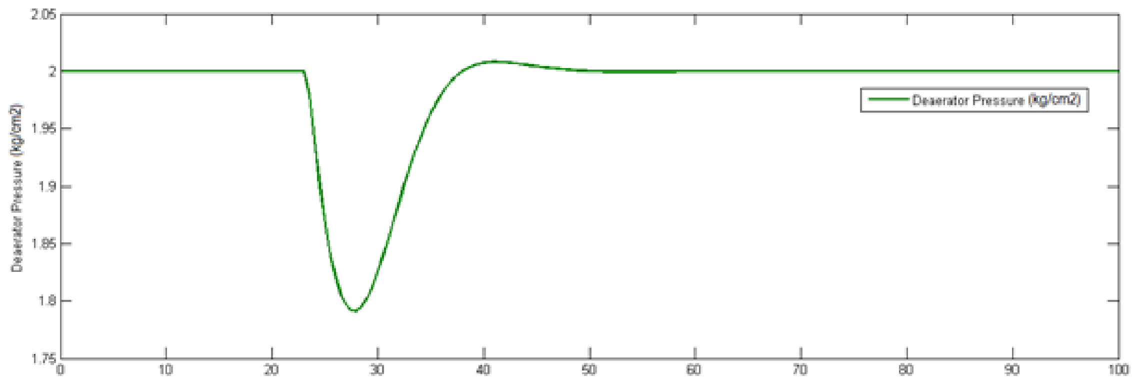
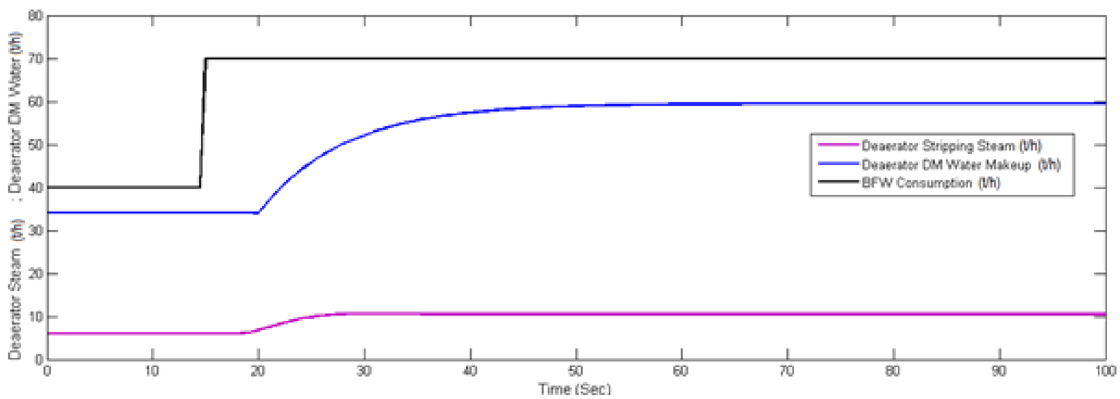


Figure 5.9 - Deaerator Model (Dynamic)

Figure 5.9 shows the block diagram of the deaerator model. The inputs to the deaerator model are BFW consumption and the deaerator initial conditions while the outputs are steam consumption, DM water consumption and deaerator drum pressure. The deaerator BFW is internally routed to a saturation block that limits the maximum BFW supply possible from the deaerator. The signal is then fed to the steam/water splitter that splits the BFW consumption to steam and DM water consumption from known thermodynamic equations, initial conditions and parameters available from "Deaerator characteristic" module. The required steam and DM water are then routed to their respective second order with dead time systems to simulate the process dynamics. In addition to this, a steam imbalance calculation is done so as to simulate the process dynamics of Deaerator drum pressure.



(a) Deaerator Pressure



(b) Deaerator Steam/DM water

Figure 5.10 - Deaerator Characteristics

Figure 5.10 shows the deaerator dynamics when the BFW consumption changes from 40 t/h to 70 t/h. The DM water begins to rise from its initial value of 34 t/h to 60 t/h as shown in the Deaerator DM water makeup graph. It is seen that the DM water starts to rise after 6 s. This is due to the fact that the DM water control valve is connected to the deaerator level controller and only when the level starts falling down, that the control action takes place. As the drum level begins to fall the deaerator drum pressure also begins to fall as shown in the deaerator pressure graph. This fall in pressure is sensed by the pressure controller that increases the steam input until the deaerator pressure reaches the initial operating point. The rise in the steam input to the deaerator can be observed in the deaerator stripping steam graph, Figure 10(b).

5.3.6 Steam Header Model

Figure 5.11 shows the block diagram of the deaerator model. The inputs to the steam header model are steam inflow, steam outflow and the steam header initial conditions while the outputs are the steam header pressure and the steam header blow-off. The steam inflow and steam outflow are routed to a steam imbalance calculator that calculates the change in pressure that will happen due to steam imbalance in the header. This signal is then routed to a second order system with dead time that simulates the process dynamics of a steam header pressure. The model also has a steam blow off point that takes input from the steam header pressure. If the pressure shoots up beyond a predetermined blow-off setpoint the steam blow-off valve releases steam into the atmosphere to bring down the header pressure. The steam header model also has "Steam Header characteristics" module for synchronization with the optimization model and to hold critical header parameters.

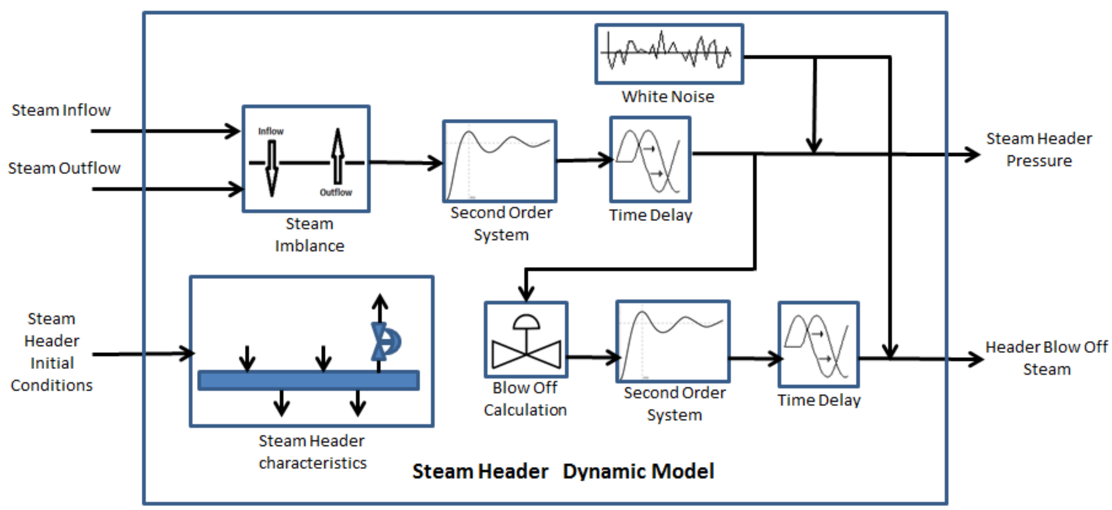
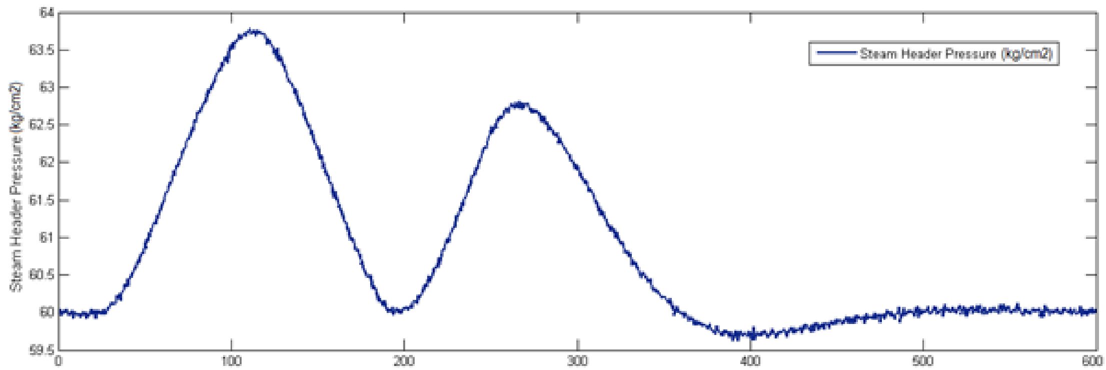
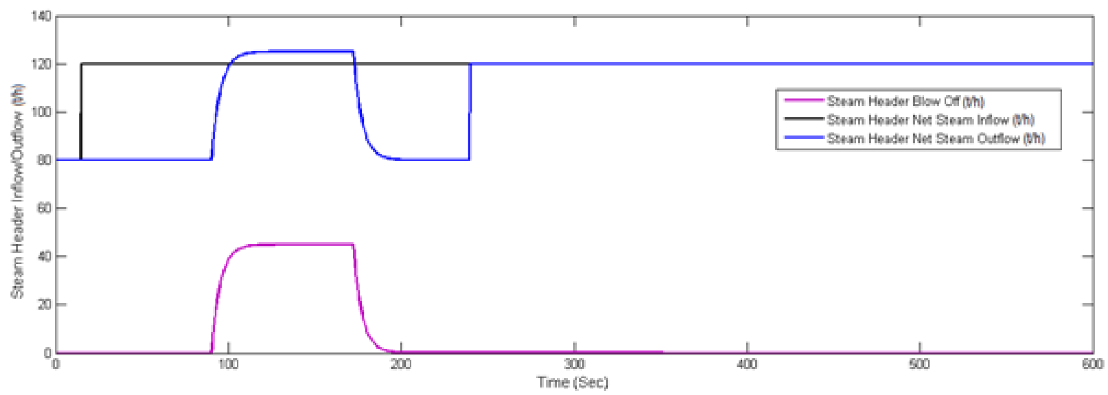


Figure 5.11 - Steam Header Model (Dynamic)

Dynamics of the pressure header due to changes in steam inflow and outflow are shown in Figure 5.12. For the first fifteen seconds of the simulation the inflow and the outflow are the same at 80 t/h and the header pressure remains constant at 60 kg/cm^2. Then the steam inflow suddenly increases to 120 t/h but the steam outflow remains the same. Now the pressure starts to increase and when it reaches 63.5 kg/cm^2 the steam blowoff is activated. The blow off header releases 50 t/h of steam into the atmosphere and the outflow increases to 130 t/h; this causes a reduction in the steam header pressure. As the steam header pressure reduces to around 60.5 hg/cm^2 the blow off valve closes and the outflow reduces to its pre blow-off value of 80 t/h, but as the inflow still remains at 120 t/h the steam pressure starts to increase again. However at 250 s, the outflow itself increases to 120 t/h and the steam pressure starts coming down and finally settles down at 60 kg/cm^2 at 500 s.



(a) Steam Header Pressure



(b) Steam Header Inflow/Outflow

Figure 5.12 - Steam Header Characteristics

Chapter Six: Control Schema

6.1 Control Hierarchy

Modern process and power plants have different levels of control with each level targeting a specific objective. The objectives range from low level plant control to high level performance analysis and report generation. Figure 6.1 shows a typical control hierarchy used in modern process plants. The same hierarchy can be successfully applied to a cogeneration system as well.

- Level 1 – Plant Equipment

Level 1 represents the plant equipment and the field transducers used for controlling and monitoring the equipment. In case of a cogeneration system the plant equipment can be a gas turbine or a boiler; and the field transducer can be the servo valve used to control the fuel flow to the turbine or a control valve used to control feed water flow into the boiler.

- Level 2 – Targeted Controllers

Level 2 represents the individual controllers that control a specific equipment or subsystem in the plant. A typical example is a turbine governor system that controls a gas turbine or a steam turbine. Plant Distributed Control System (DCS) also belongs to this category. Though typically a DCS controls a vast majority of the plant subsystems, it can be viewed as an aggregation of several controllers unified in a centralised control platform.

- Level 3 – Optimization and Multivariable Control

Level 3 represents the first supervisory control layer in the system. Usually both optimization and multivariable control are implemented in this level. Functionalities in level 3 can be broadly divided into three major modules

- Data Validation and Control: This module processes all the plant data both from the field and the individual controllers; validates the data and passes it on to the optimizer. The data validation is important as the quality of the optimization results depends on the quality of the input data. In addition, incorrect data may lead to system instability and wide operational swings. To avoid this, predetermined data validation rules are built to check field measurement. A simple example of data validation can be found in an extraction type steam turbine. In an extraction type steam turbine, total steam input to the turbine should be equal to the sum of extraction flow and condenser flow. If they do not match then an error has occurred in one or more of the flow meters connected to the turbine input, extraction and condenser. This error can be resolved in many ways. The most common methods used are to take the power balance in the turbine or to heuristically choose a transducer to have a correct reading and calculate the rest to resolve the error. If high fidelity mathematical models are available, then using model data as a substitute for real time operational data is also an option.

- Integrated Optimizer: This module collects the plant data from the data validator, the input data and optimization objectives from the operator, and generates optimal set points for the real time plant condition. The objective is usually to minimize the cost as seen in chapters 2-4, with additional goals depending on how the plant needs to be operated. The integrated optimizer need not always be online; usually the optimizer module can also be used for offline analysis. This is usually done for maintenance and operational planning of the system.
- Multivariable Controller - This module gets the optimal set points from the optimizer and takes the plant from the present operating condition to the optimal operating condition. The multivariable controller is usually a set point controller that interfaces directly with the plant low level controllers and DCS systems. In this thesis, the supervisory fuzzy controller is implemented in this module.

➢ Level 4 – Performance Monitoring

Layer 4 represents the second supervisory control layer. This layer, also called the enterprise resource layer, is used to present only the most critical plant data to the plant higher management. This data is mostly used by managers to know about overall plant performance and operation. The long term plant historian is usually implemented in this level. This layer can be broadly divided into two modules:

- Equipment Monitoring: This module monitors the healthiness of the plant equipment and is mostly used for maintenance planning. Most condition based maintenance systems are implemented in this module.
- Process Monitoring: This module monitors the plant operation and is used to make operational and process improvement decisions. Most energy and lifecycle management systems are implemented in this module.

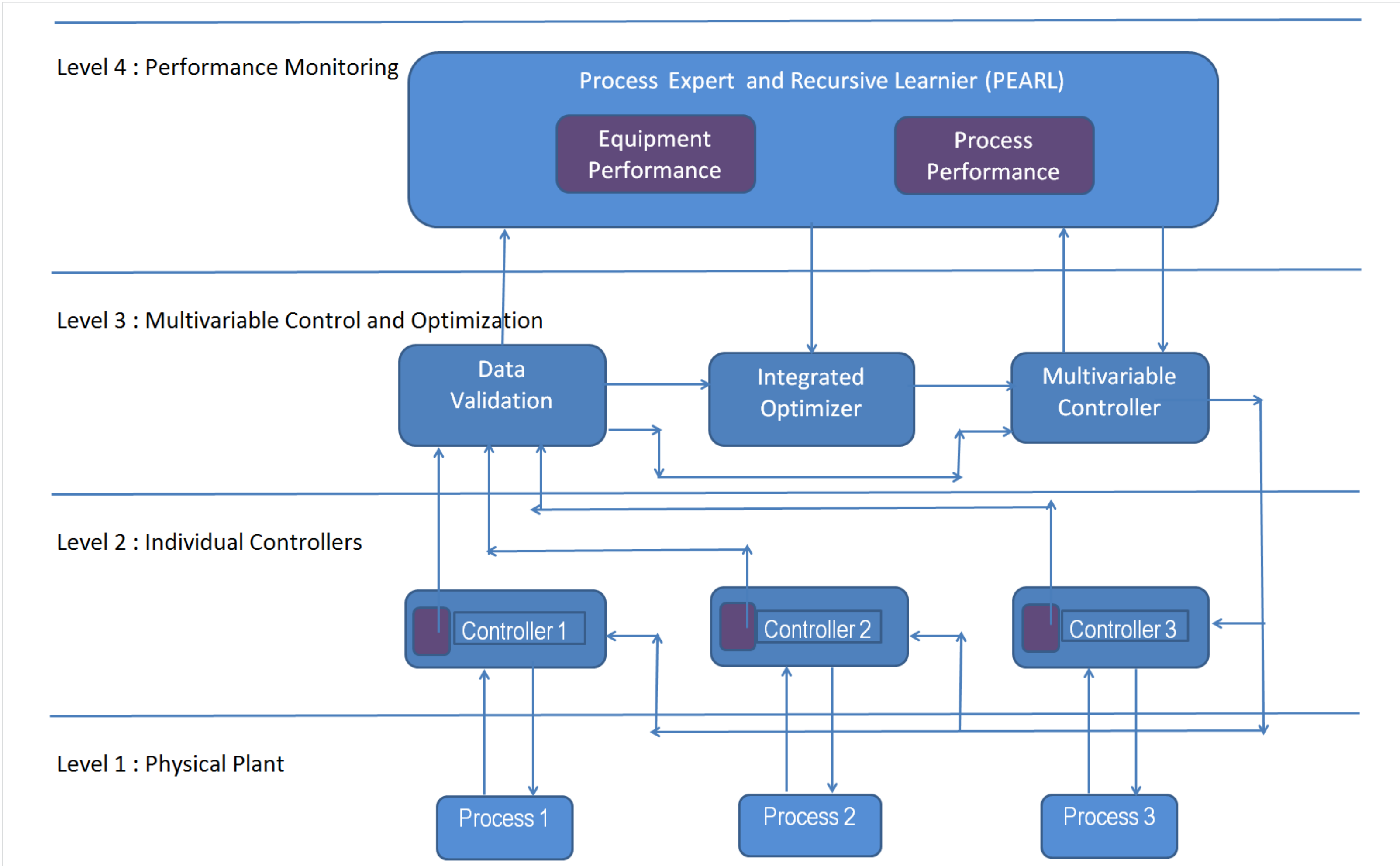


Figure 6.1 - Control Hierarchy

6.2 Fuzzy Control Scheme

In view of the control objectives specified in section 1.2.2 a fuzzy supervisory control is selected for plant control. Fuzzy supervisory control has the following advantages:

- Fuzzy control depends only on predetermined rule base and not on plant models, thus requires little or no plant system identification tests.
- It naturally handles multiple inputs and outputs. In addition the control can be split into smaller units and distributed in a specific way, each one assigned with a limited but specific purpose thus achieving plant wide control.
- It is inherently robust and does not need precise and noise free inputs. Thus control is not affected significantly by loss or malfunction of sensors.
- Fuzzy controller acting as a set point controller can be easily integrated into the existing plant control network with minimal downtime.

For cogeneration systems the control objectives can be summarised as below:

- Take generators and boiler units from present operating condition to optimal setpoint generated by the optimizer in minimum possible time.
- Keep power import/export from tie control within specified limits
- Keep steam header pressures in multiple header units within specified limits.

To judge the effectiveness of the control system, critical performance parameters have to be defined. In view of the control objectives specified for a cogeneration system, the following parameters are considered to be the performance indicators in this thesis:

- Deviation from optimal set point for generators and boilers
- Process settling time
- Maximum deviation in tie line power level
- Maximum deviation in steam header pressure level

In an ideal scenario, deviation from optimal set point must be zero under steady state conditions, the process settling time should be as fast as possible, with industry standard time less than thirty minutes, and the maximum deviation of critical parameters levels should be kept less than 10% of their required operating points.

To achieve the control and performance objectives five different fuzzy control schemes have been developed.

- Tie Power control scheme

  Any generator in the cogeneration system can operate in tie control mode. In this scheme the generator looks only for the tie power and tries to maintain it at the required operating point. The generator in this mode acts like a slack generator and takes care of the system disturbances and keeps the critical tie line power within limits.

- Generator set point control scheme

  Most of the generators in the cogeneration system will operate in the generator set point control mode. In this mode the fuzzy supervisory control takes the system from the present operating point to the optimal operating point.

- Steam header pressure control scheme

  Any boiler unit in the cogeneration system can operate in header pressure control mode. In this scheme the boiler looks only to maintain the steam header pressure it controls. The boiler in this mode acts like a slack boiler and takes care of the system disturbances, and keeps the steam header pressure within limits.

- Boiler set point control scheme

  Most of the boiler units will operate in the boiler set point control mode. In this mode the controller takes the boiler from the present operating point to the optimal operating point.

- PRDS set point control scheme

  All the PRDS in the system are put in the PRDS set point control scheme. In this mode the PRDS not only tries to reach its optimal operating point but also keeps in check the downstream/upstream header pressure.

All five control schemes are used in tandem in the cogeneration plant to achieve the control objectives. The fuzzy rules are designed so that the controller outputs do not conflict with one another. Detailed descriptions of the control schemes are given in the following sections.

6.3 Tie Power Control

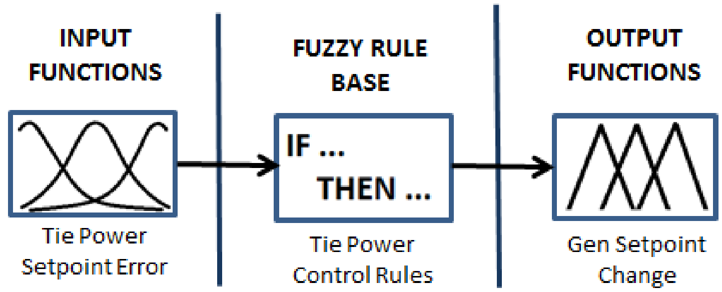


Figure 6.2 - Fuzzy Tie Power Control

Figure 6.2 shows the fuzzy tie power control scheme. Input to the tie power control scheme is the tie power set point error and the output is the power set point change of the generator that is operated in the tie power control scheme. In this mode the generator just tracks the tie power error and tries to bring it to zero. Usually one of the generators in the fleet is operated in tie power control scheme. This generator not only regulates the tie power to its set point while the other generators swing to meet their optimal set point conditions but also takes care of changes in the power loads that may happen in between an optimization run.

6.3.1 Input and Output Basis Functions

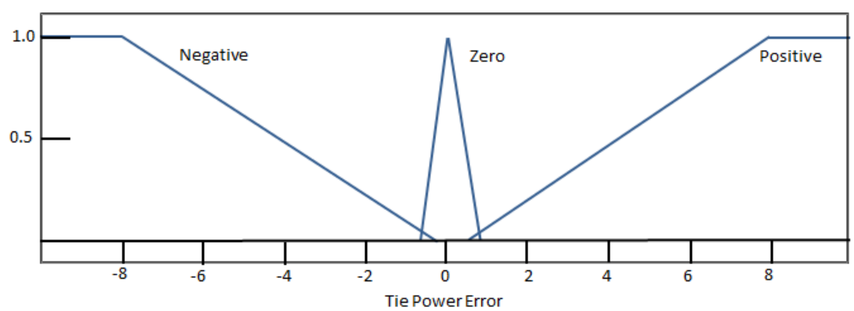


Figure 6.3 - Input Basis Functions (Tie Power Control)

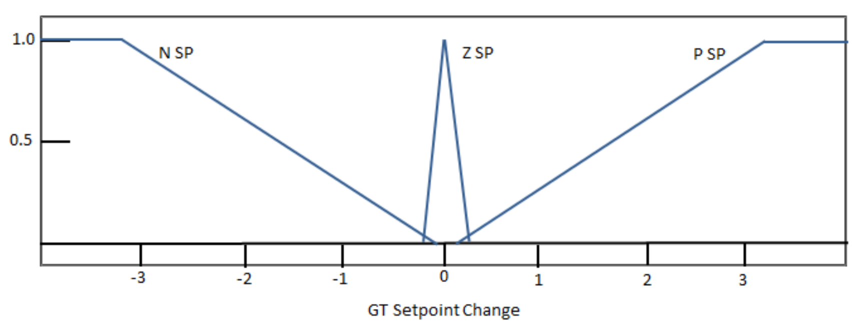


Figure 6.4 - Output Basis Functions (Tie Power Control)

Figures 6.3 and 6.4, respectively, show the input and output basis functions. The range of the input basis functions is chosen in consideration with the maximum allowable tie header error, while the output basis functions are chosen according to the GT set point rate limiter.

6.3.2 Fuzzy Rule Base

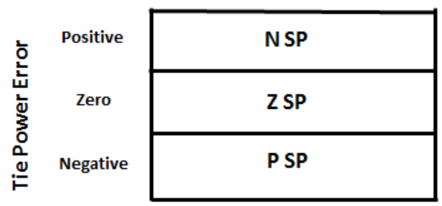


Figure 6.5 - Fuzzy Rule Base (Tie Power Control)

Figure 6.5 shows the fuzzy rule base for tie power control scheme. When the tie power error is negative it is necessary to increase the generator output to bring the tie error to zero. The fuzzy rule base does that by giving a ‘P SP’ or a positive set point change that increases the generator output. When the error is zero, it does nothing and gives a ‘Z SP’ or a zero set point change and during positive tie error conditions it gives a ‘N SP’ or a negative set point change.

6.3.3 Control Surface

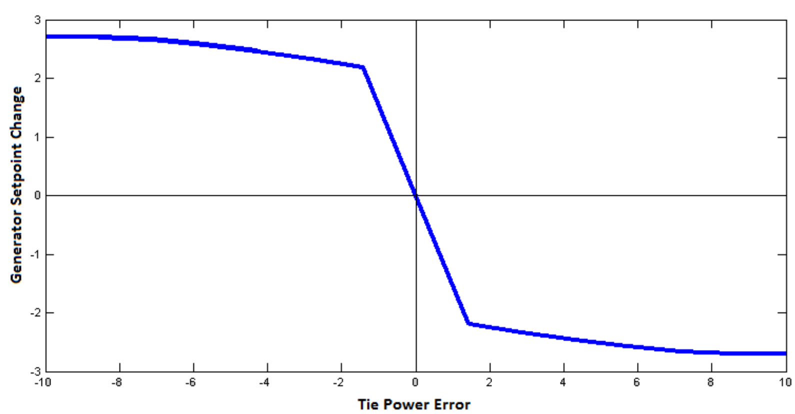


Figure 6.6 - Control Surface (Tie Power Control)

Figure 6.6 shows the control surface for the tie power controller. It can be observed that when the tie power error is positive the generator set point change is negative and when the tie power error is negative the generator set point error is positive.

6.4 Generator Setpoint Control

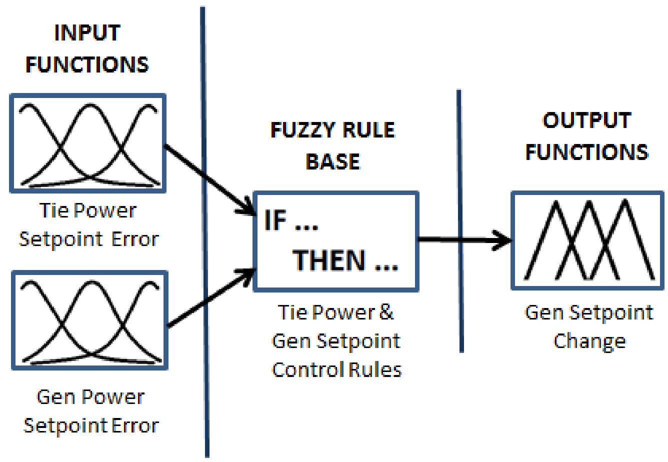


Figure 6.7 - Fuzzy Generator Set point Control

Figure 6.7 shows the fuzzy generator set point control scheme. Inputs to the control are the tie power setpoint error and the generator power set point error. The generator power set point error signifies the difference between the optimal operating condition and the present operating condition. This controller is primarily responsible for taking the generator to its optimal operating condition. Generators operating in this mode move towards their optimal set points and do not care if the load changes happen in the system.

6.4.1 Input and Output Basis Functions

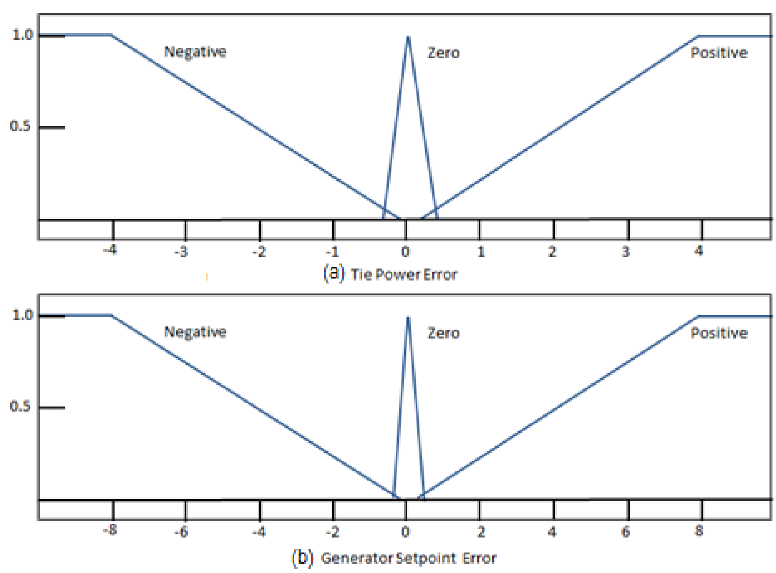


Figure 6.8 - Input Basis Functions (Generator Set point Control)

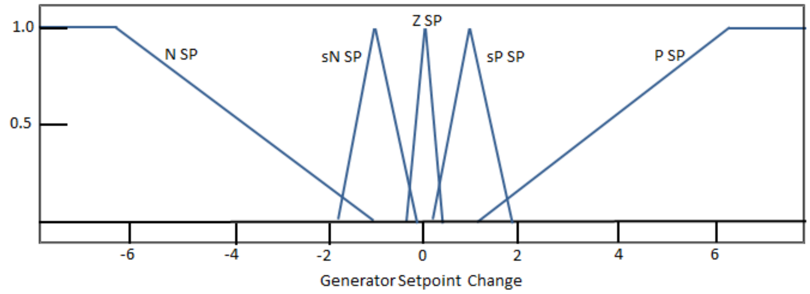


Figure 6.9 - Output Basis Functions (Generator Set point Control)

Figures 6.8 and 6.9, respectively, show the input and output basis functions. The range of the input basis functions is chosen in consideration with the maximum allowable tie header error and maximum generator set point error, while the output basis functions are chosen according to the GT set point rate limiter.

6.4.2 Fuzzy Rule Base

Tie Power Error \ GT Setpoint Error	Positive	Zero	Negative
Positive	N SP	Z SP	Z SP
Zero	sN SP	Z SP	sP SP
Negative	Z SP	Z SP	P SP

Figure 6.10 - Fuzzy Rule Base (Generator Set point Control)

Figure 6.10 shows the fuzzy rule base used in generator set point control. The fuzzy rule base is designed to mimic an intelligent operator. When both the tie power error and generator set point error are negative, both errors can be reduced by increasing the set point of the generator. The fuzzy rule base implements this by giving a “P SP” or positive set point. The reverse happens when both the tie power error and generator set point error are positive. In conditions when the tie power error is positive and generator set point error is negative or vice versa, the controller takes no action by giving a “Z SP” or zero set point change as any change will aggravate the situation of one of the control parameters. This is explained in the following example. Assume that the tie power error is positive and generator set point error is negative. If a positive set point change is given then generator set point error will reduce but the critical parameter tie power error will increase. Instead if a negative set point change is given then the tie power error will reduce but the plant will move away from the optimal generator operating point. Thus the best course of action is to wait until the conditions become favourable. As the main job of the controller is to take the generator to its optimal operating point, during conditions when the tie power error is zero and there is a positive or negative error generator set point, it issues a ‘sN SP’ (small negative set point) and ‘sP SP’ (small positive set point). Any change in the tie power error caused due to this action will be handled by the generator being put in “tie power control mode”. Finally when both the tie power error and generator power error are zero, the controller gives a zero set point change as both control objectives are achieved.

6.4.3 Control Surface

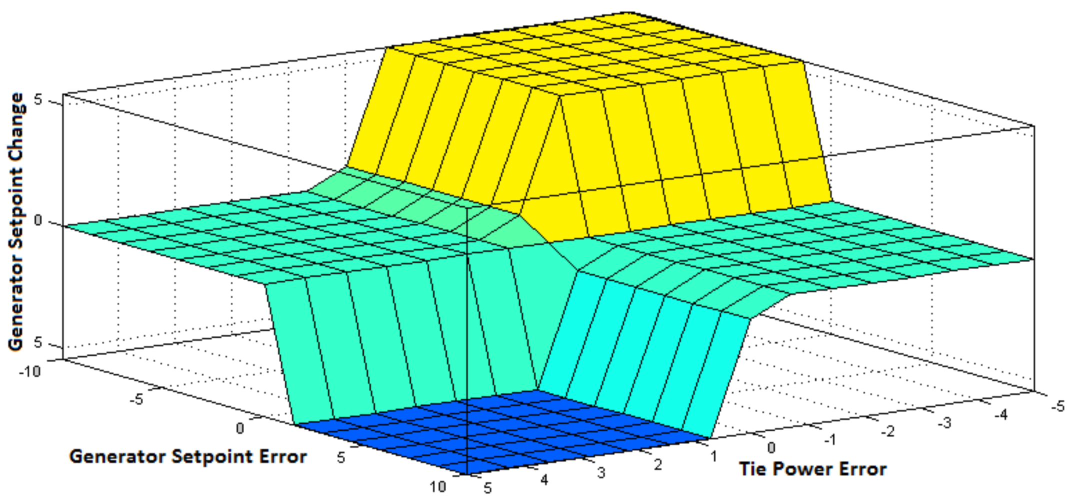


Figure 6.11 - Control Surface (Generator Set point Control)

Figure 6.11 shows the control surface for generator set point controller. It can be observed that when both the generator set point error and tie power error are positive the generator set point change is negative. When both the generator set point error and tie power error are negative the generator set point change is positive. It can be seen that the control surface output is non-zero at zero tie power error conditions. This is due to the small set point changes specified in the fuzzy rules. In conditions when tie power error and generator set point error are of opposite sign a zero set point change is the output.

6.5 Steam Header Pressure Control

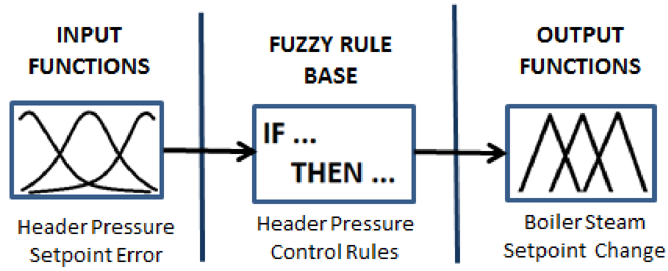


Figure 6.12 - Fuzzy Header Pressure Control

Figure 6.12 shows the fuzzy header pressure control scheme. In this mode the boiler just tracks the header pressure set point error and tries to bring it to zero. Usually one of the boilers in the boiler battery is operated in header pressure control scheme. This boiler not only regulates the header pressure to its set point while the other boilers swing to meet their optimal set point conditions but also handles changes in steam loads that may happen in between an optimization run.

6.5.1 Input and Output Basis Functions

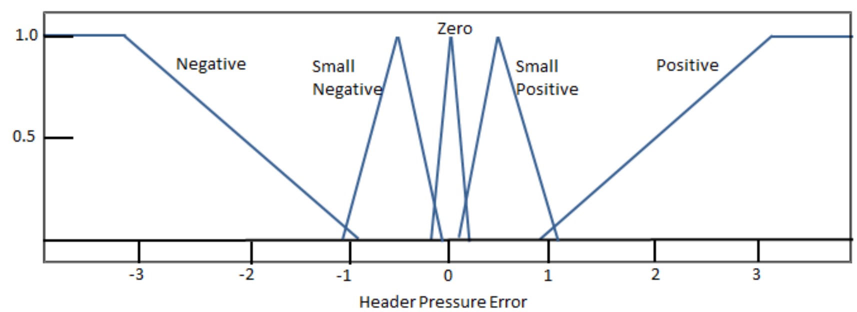


Figure 6.13 - Input Basis Function (Header Pressure Control)

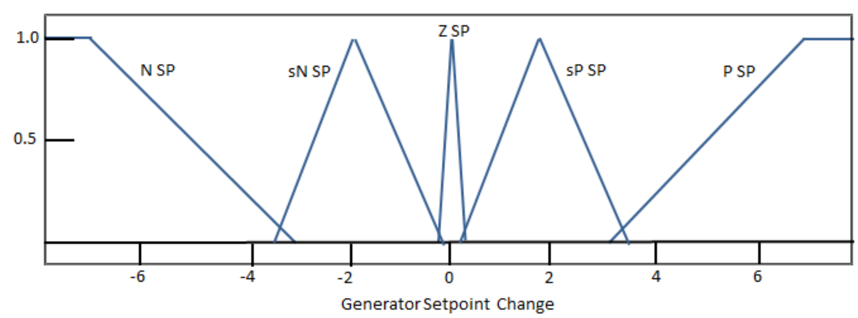


Figure 6.14 - Output Basis Functions (Header Pressure Control)

Figures 6.13 and 6.14, respectively, show the input and output basis functions. Range of the input basis functions is chosen in accordance with the maximum allowable header pressure error, while the output basis functions are chosen in line with the boiler set point rate limiter.

6.5.2 Fuzzy Rule Base

Header Pressure SP Error	
Positive	N SP
Small Positive	sN SP
Zero	Z SP
Small Negative	sP SP
Negative	P SP

Figure 6.15 - Fuzzy Rule Base (Header Pressure Control)

Figure 6.15 shows the fuzzy rule base for header pressure control. When the header pressure error is negative it is necessary to increase the boiler steam output to bring header pressure to its normal operating point. The fuzzy rule base does that by giving a 'P SP' or a positive set point change that increases the Boiler output. When the error is zero, it does nothing and gives a 'Z SP' or a zero set point change and when header pressure is high it gives a 'N SP' or a negative set point. In addition to this, to increase the speed of response, and bring the header pressure smoothly to its required operating point two additional basis functions have been used, thus the fuzzy controller gives a 'sN SP' (small negative set point) and 'sP SP'(small positive set point) when the header pressure set point errors are quite small.

6.5.3 Control Surface

Figure 6.16 - Control Surface (Header Pressure Control)

Figure 6.16 shows the control surface for the header pressure controller. It can be observed that when the header pressure error is positive the boiler set point change is negative and when the header pressure error is negative the boiler set point error is positive.

6.6 Boiler Setpoint Control

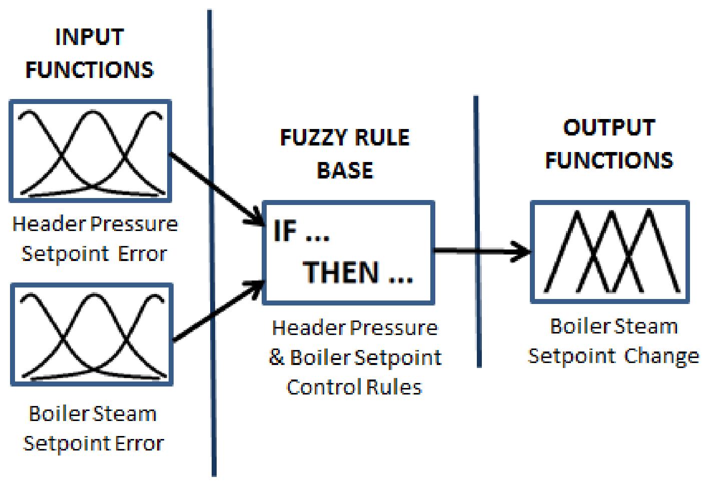


Figure 6.17 - Fuzzy Boiler Set point Control

Figure 6.17 shows the fuzzy boiler set point control scheme. The boiler steam set point error signifies the difference between the optimal operating condition and the present operating condition. Thus this controller is primarily responsible for taking the boiler to its optimal operating condition. Boilers put in this mode move towards their optimal set points and do not care for the load changes that can take place in the system.

6.6.1 Input and Output Basis Functions

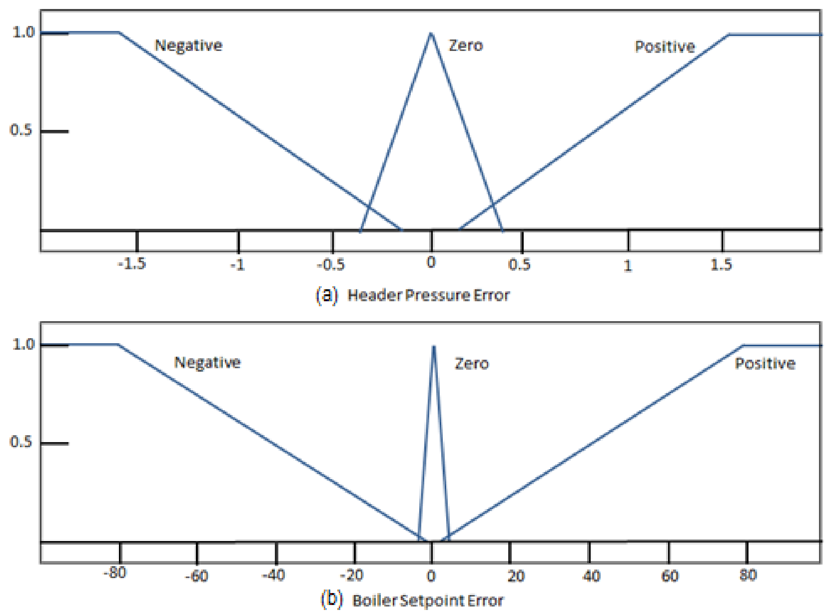


Figure 6.18 - Input Basis Functions (Boiler Set point Control)

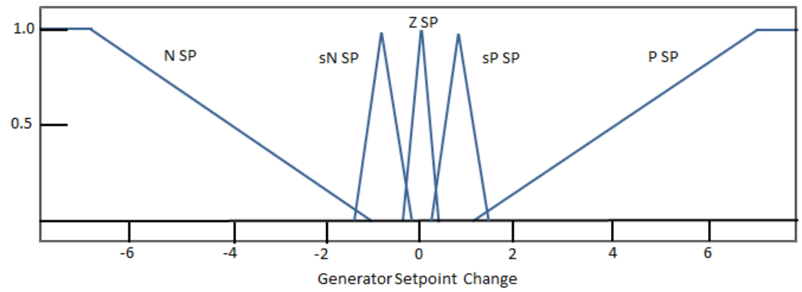


Figure 6.19 - Output Basis Function (Boiler Set point Control)

Figures 6.18 and 6.19, respectively, show the input and output basis functions. The range of the input basis functions is chosen in accordance with the maximum allowable header pressure error and maximum possible boiler set point error, while the output basis functions are chosen in line with the boiler set point rate limiter.

6.6.2 Fuzzy Rule Base

Header Pressure SP Error	Boiler SP Error: Positive	Zero	Negative
Positive	N SP	Z SP	Z SP
Zero	sN SP	Z SP	sP SP
Negative	Z SP	Z SP	P SP

Figure 6.20 - Fuzzy Rule Base (Boiler Set point Control)

Figure 6.20 shows the fuzzy rule base of the boiler set point control. The fuzzy rule base is very similar to the generator set point control and is designed to mimic human operator control. When both the header pressure error and boiler set point error are negative, both errors can be reduced by increasing the steam output of the boiler. The fuzzy rule base implements this by giving a 'P SP' or positive set point change to the boiler. The reverse occurs when both the header pressure error and boiler set point error are positive. In conditions when header pressure error is positive and boiler set point error is negative or vice versa, the controller takes no action by giving a 'Z SP' or zero set point change as any change will aggravate the situation of one of the control parameters. Assume that the header pressure is high and boiler set point error is negative. If a positive set point change is given then the boiler set point error will reduce but the critical parameter header pressure will increase. If a negative set point change is given then the header pressure will come down but the plant will move away from the optimal boiler operating point. Thus the best course of action is to wait until the conditions become favourable. As the main job of the controller is to take the boiler to its optimal operating point, during conditions when the header pressure set point error is zero and there is a positive or negative error in boiler set point, it issues a 'sN SP' (small negative set point) and 'sP SP' (small positive set point). Any change in the header pressure caused due to this action will be handled by the boiler put in "header pressure control mode". Finally when both the header pressure set point error and boiler set point error are zero, the controller gives a zero set point change as both control objectives are achieved.

6.6.3 Control Surface

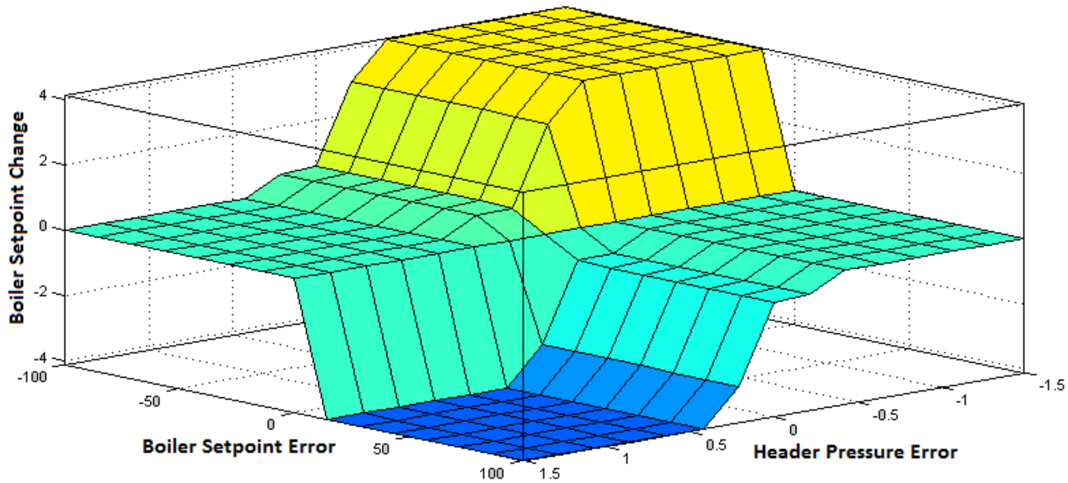


Figure 6.21 - Control Surface (Boiler Set point Control)

Figure 6.21 shows the control surface for boiler set point controller. It can be observed that when both the boiler set point error and header pressure error are positive the boiler set point change is negative. When both the boiler set point error and header pressure error are negative the boiler set point change is positive. It can be observed that the control surface output is non-zero at zero header pressure error conditions. This is due to the small set point changes specified in the fuzzy rules. In conditions when header pressure error and boiler set point error are of opposite signs, a zero set point change is the output.

6.7 PRDS Setpoint Control

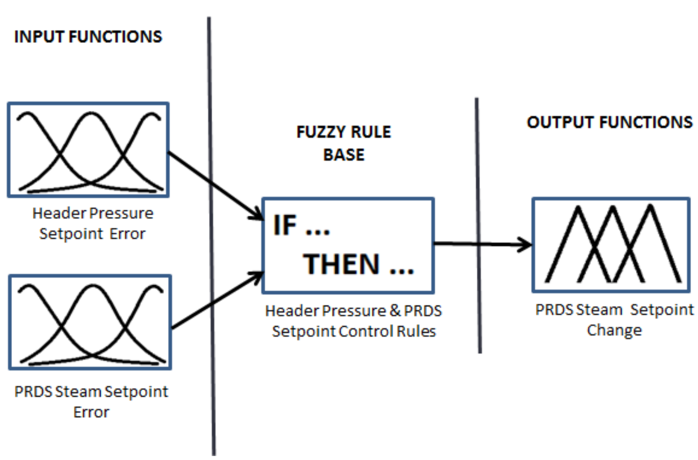


Figure 6.22 - Fuzzy PRDS Set point Control

Figure 6.22 shows the fuzzy PRDS set point control scheme. The PRDS steam set point error signifies the difference between the optimal operating point and the present operating condition. Thus this controller is primarily responsible for taking the PRDS to its optimal operating condition while maintaining the downstream steam header pressure.

6.7.1 Input and Output Basis Functions

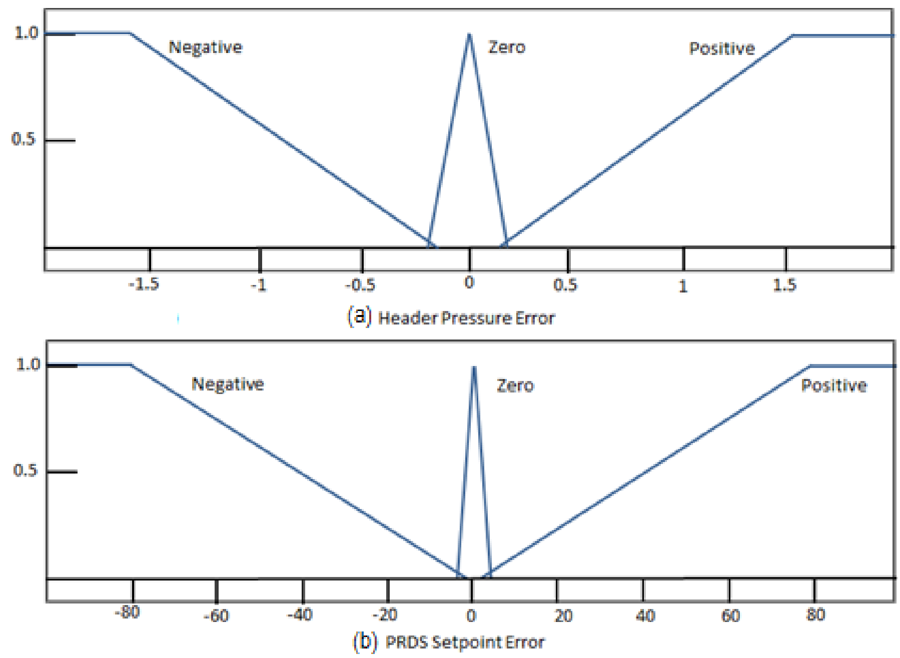


Figure 6.23 - Input Basis Functions (PRDS Setpoint Control)

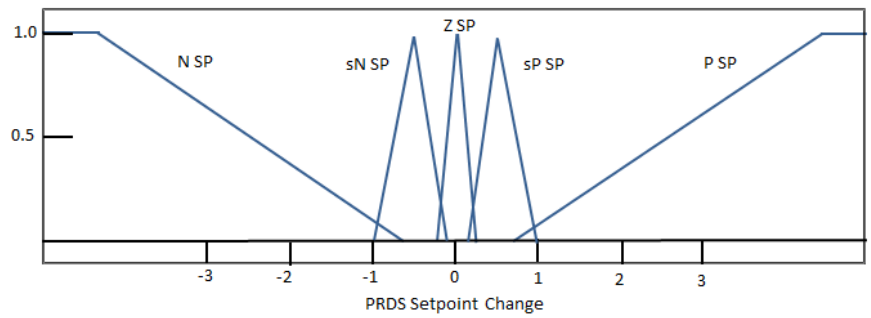


Figure 6.24 - Output Basis Functions (PRDS Set point Control)

Figures 6.23 and 6.24, respectively, show the input and output basis functions. The range of the input basis functions is chosen in accordance with the maximum allowable header pressure error and maximum possible PRDS set point error, while the output basis functions are chosen in line with the allowable PRDS rate change.

6.7.2 Fuzzy Rule Base

Header Pressure SP Error	PRDS SP Error: Positive	Zero	Negative
Positive	N SP	Z SP	Z SP
Zero	sN SP	Z SP	sP SP
Negative	Z SP	Z SP	P SP

Figure 6.25 - Fuzzy Rule Base (PRDS Set point Control)

Figure 6.25 shows the fuzzy rule base for PRDS control. The fuzzy rule base is very similar to the boiler set point control and is designed to mimic human operator control. When both the downstream header pressure error and PRDS set point error are negative, both errors can be reduced by increasing the downstream steam output from the PRDS. The fuzzy rule base implements this by giving a ‘P SP’ or positive set point change to the PRDS. The reverse occurs when both the header pressure error and boiler set point error are positive. In conditions when downstream header pressure error is positive and PRDS set point error are negative or vice versa, the controller takes no action by giving a ‘Z SP’ or zero set point change as any change will aggravate the situation of one of the control parameters. Assume that the downstream header pressure is high and PRDS set point error is negative. If a positive set point change is given then the PRDS set point error will reduce but the downstream header pressure will increase. If a negative set point change is given then the downstream header pressure will come down but the plant move away from the optimal PRDS operating point. Thus the best course of action is to wait till the conditions become favourable. As the main job of the controller is to take the PRDS to its optimal operating point, during conditions when the downstream header pressure set point error is zero and there is a positive or negative error in PRDS set point, it issues a ‘sN SP’ (small negative set point) and ‘sP SP’ (small positive set point). Any change in the header pressure caused due to this action will be handled by the boiler controlling the header and put in “header pressure control mode”. Finally when both the downstream header pressure set point error and PRDS set point error are zero, the controller gives a zero set point change as both control objectives are achieved.

6.7.3 Control Surface

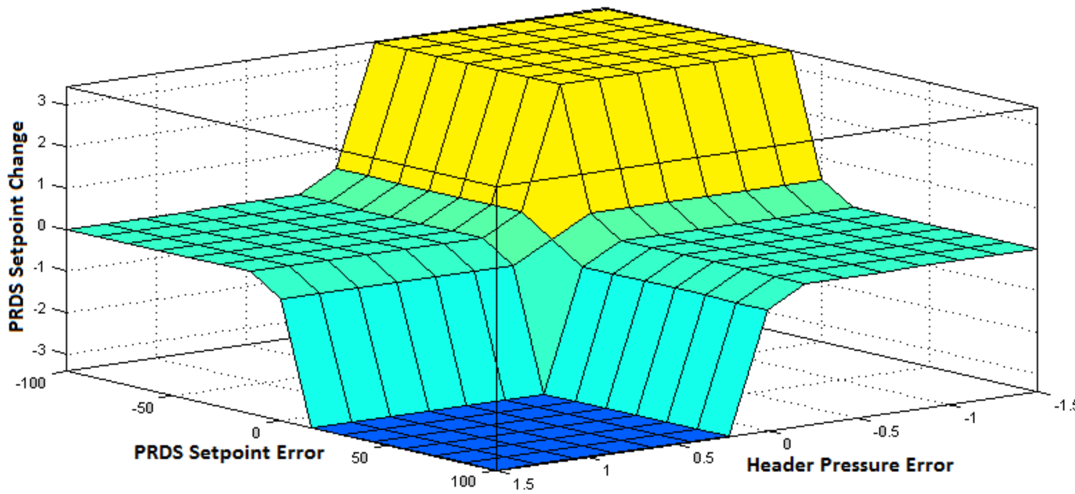


Figure 6.26 - Control Surface (PRDS Set point Control)

Figure 6.26 shows the control surface for PRDS set point controller. It can be observed that, when both the PRDS set point error and header pressure error are positive, the PRDS set point change is negative. When both the PRDS set point error and header pressure error are negative the PRDS set point change is positive. It can be observed that the control surface output is non-zero at zero header pressure error conditions. This is due to the small set point changes specified in the fuzzy rules. In conditions when header pressure error and boiler set point error are of opposite signs, a zero set point change is the output.

Chapter Seven: Supervisory Control of Cogeneration Systems

The model plant used to test the fuzzy control is the same refinery cogeneration plant used for testing the optimization algorithm (Appendix A). Before the fuzzy supervisory control was implemented, alternate control strategies were tested. Section 7.1 gives a brief description of the alternate control strategies and the associated plant response. Three separate simulation studies involving the fuzzy supervisory control under different conditions have been performed to judge the effectiveness of the control scheme. Section 7.2 showcases a simulation when the plant is taken from an initial operating point to optimal operating point. Sections 7.3 and 7.4 describe cases when the process power and steam load changes occur while the plant is still moving towards its optimal operating point.

7.1 Alternate Control Strategies

Prior to implementing the fuzzy supervisory control, two different control strategies were tested on the plant to test the performance of alternate supervisory control. The control was tested on the same refinery cogeneration plant. Brief description of the control techniques and the plant response to the control is described below.

7.1.1 Constant Setpoint Control

To know if a supervisory control was needed in addition to the targeted controllers available for each plant process, a constant setpoint control test was conducted. Here the set point of the plant equipment was kept constant at the optimal setpoint for the entire duration of the simulation. Thus in effect the supervisory scheme just takes the optimal setpoint from the optimizer and gives out a constant setpoint value equal to its optimal setpoint. This is a simple method of control that will work for some cases, due to the presence of targeted low level controllers that take care of individual process dynamics.

Tie power and the associated GT power changes when a constant set point control is applied are shown in Fig. 7.1. Here the GT's are fed their optimal setpoints of 8 MW, 18 MW and 22 MW for GT-1, GT-2 and GT-3, respectively. The GT's reach their optimal setpoint values in less than 100 s but the tie power has an enormous swing of about 2 MW when the GT's were proceeding to their optimal setpoint. This huge tie power control error is unacceptable for normal operation.

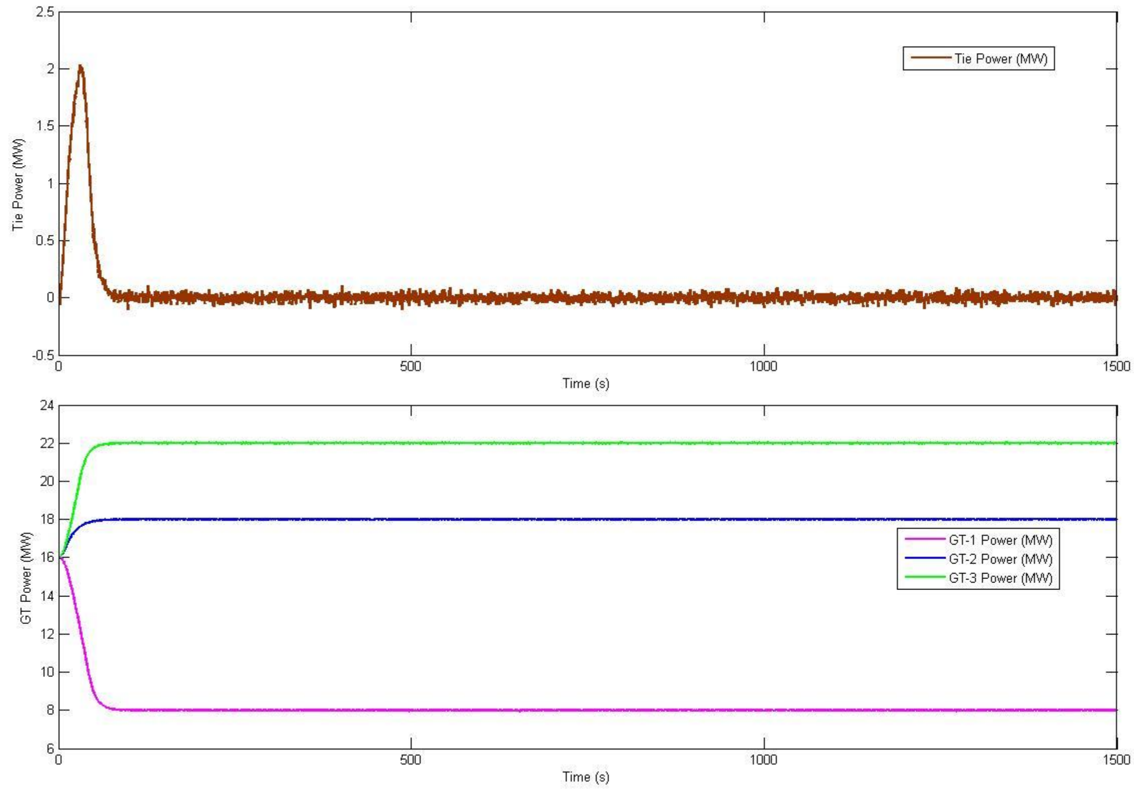


Figure 7.1 - Tie Power Control (Constant Setpoint Control)

Figure 7.2 shows the MP header pressure swing when the constant setpoint control is applied to boilers. Here, Boiler-1, Boiler-2 and Boiler-3 are given a constant optimal setpoint value of 120 t/h, 104 t/h and 30 t/h, respectively. As the boilers proceed to their optimal setpoints, the MP header pressure, that is to be maintained at 14 kg/cm^2, reaches a peak value of 15.5 kg/cm^2. This huge pressure variation is quite close to the blow off set point of the steam header and cannot be allowed for normal operation. In addition, PRDS-1 and PRDS-3 are also given constant set points of 63 t/h and 140 t/h, respectively. They also swing to their set points in less than 100 s. It is seen that all the boilers reach their optimal set point values in less than 300 sec and the system attains a steady state operation in about 500 s.

From the above it can be seen that the constant set point control has the advantage that it is simple to implement and the process settling time is quite fast, but it fails to control huge variations in the critical parameters such as tie power and header pressures, and thus cannot be used as a reliable plant control.

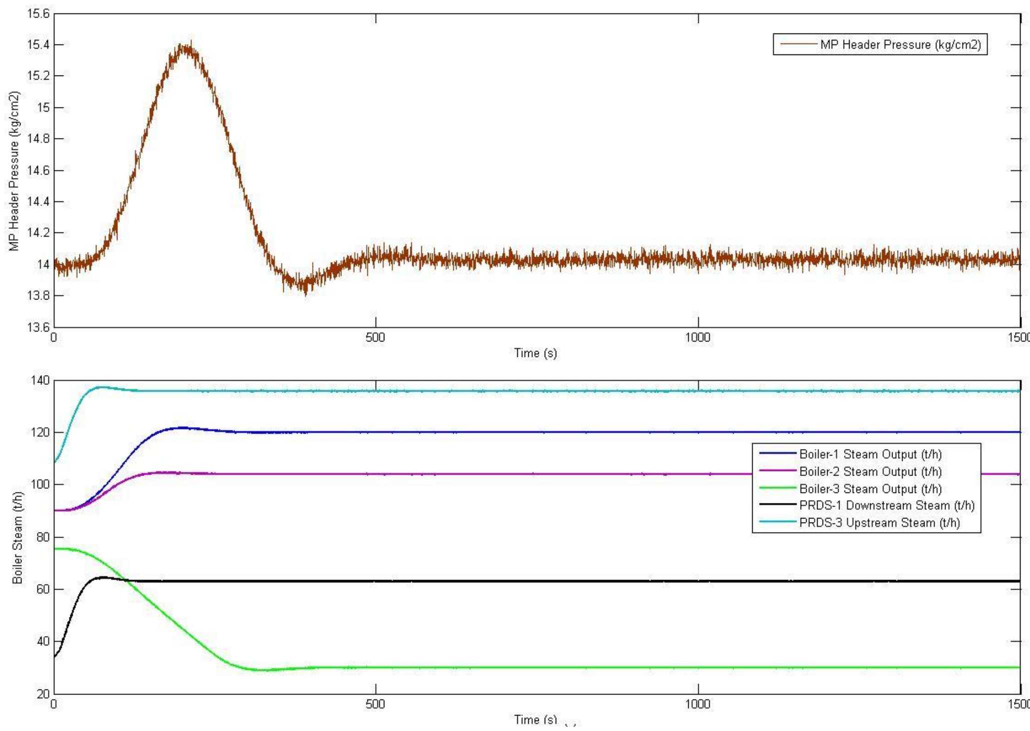


Figure 7.2 - MP Header Control (Constant Setpoint Control)

7.1.2 Heuristic Setpoint Control

Heuristic setpoint control uses ad-hoc fuzzy implementation based on expert process knowledge to control the plant parameters. There are three modes of operation, the generator control mode, the boiler control mode and PRDS control mode. Rules governing the generator control mode, boiler control mode and PRDS control mode are similar to the fuzzy rules used by "Generator setpoint control", "Boiler setpoint control" and " PRDS setpoint control" explained in section 6.4, 6.6 and 6.7, respectively. The heuristic controller acts as an intelligent operator and takes control action only when the conditions are favourable. But unlike the fuzzy controller there are no membership functions. The output is a simple proportional control where the output depends on a gain and the error.

In the generator control mode, the output is a product of a predetermined gain, tie power error and generator setpoint error. In the boiler control mode, the output is a product of a predetermined gain, header pressure error and boiler setpoint error. In the PRDS control mode the control action is a product of a predetermined gain, PRDS setpoint error and downstream or upstream header pressure error.

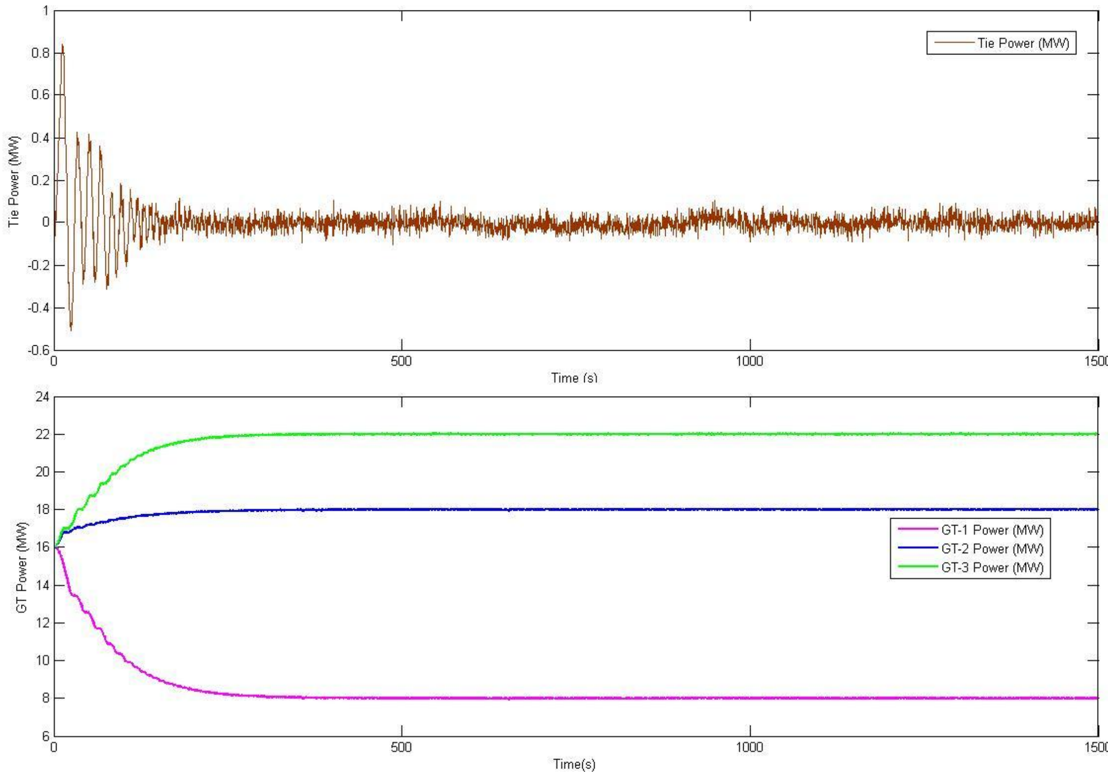


Figure 7.3 - Tie Power Control (Heuristic Setpoint Control)

Figure 7.3 shows the tie power error and the GT power swings when the heuristic controller is implemented in the plant. The GT's reach their respective setpoints in less than 150s and the tie power settles down at around the same time. However, the tie power exhibits a maximum swing of 0.8 MW and the tie power error oscillates quite a bit before settling down.

Figure 7.4 shows the MP header pressure and the boiler swings for a heuristic control action. Here too the header pressure shows wide pressure variation, a maximum negative deviation of 0.8 kg/cm^2 and a maximum positive deviation of 1.2 kg/cm^2. Though the negative swing is acceptable the positive swing is again a bit high for normal operation. The boilers reach their optimal operating point in less than 500 s and the total plant reaches steady state operation in about 700 s.

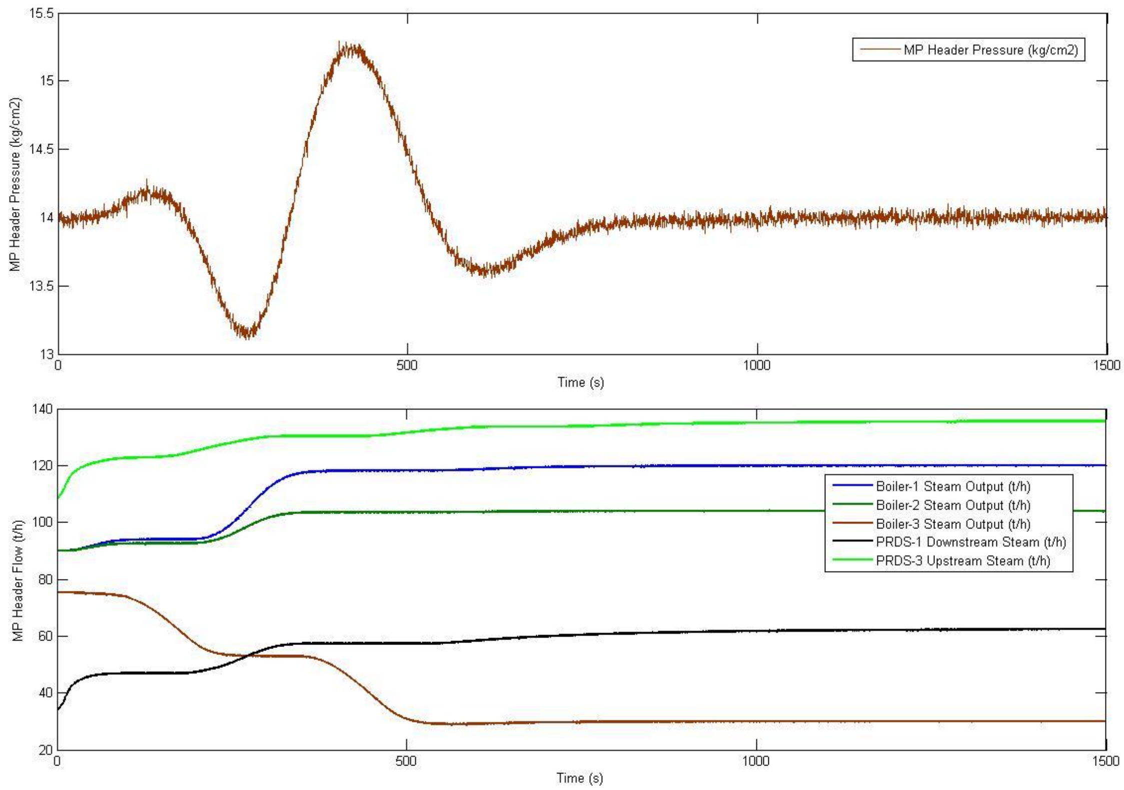


Figure 7.4 - MP Header Control (Heuristic Setpoint Control)

The heuristic controller has the advantage that it is fairly simple to implement and has good settling time. However the main controlling factor is the gain that determines the speed of response. The gain has a direct effect on the speed of response until it hits the rate limiter limits imposed by the targeted controllers on turbine and boilers. It was found during testing that by reducing the gain, the peak deviations of critical parameters could be reduced but the plant took a longer time to settle. If the gain was increased, the peak deviations increased but the plant settled down in a much smaller time. It was also observed that setting a particular gain worked well for one simulation but not for another. Thus the main disadvantage of this method was coming up with a gain that would work for all operational scenarios.

In addition, the controller in its present form could not handle disturbances and there was no dedicated generator or boiler control mode for controlling tie power or header pressure on its own.

7.2 Simulation with Optimization Parameters

Table 7.1 - Plant Parameters for Simulation

Plant Parameters	Initial Plant Conditions	Optimized Plant Conditions
Plant Equipment Parameters		
GT - 1 Power	16 MW	8 MW
GT - 2 Power	16 MW	18 MW
GT - 3 Power	16 MW	22 MW
HRSG - 1 HP Steam	33 t/h	21 t/h
HRSG - 1 MP Steam	4.5 t/h	3.7 t/h
HRSG - 2 HP Steam	33.5 t/h	36.5 t/h
HRSG - 2 MP Steam	4.6 t/h	4.8 t/h
HRSG - 3 HP Steam	36.4 t/h	46.3 t/h
HRSG - 3 MP Steam	5.9 t/h	6.65 t/h
Boiler - 1 MP Steam	90 t/h	120 t/h
Boiler - 2 MP Steam	90 t/h	104 t/h
Boiler - 3 MP Steam	75.5 t/h	30 t/h
PRDS - 1 Downstream Steam	34.5 t/h	63 t/h
PRDS - 2 Downstream Steam	28 t/h	0 t/h
PRDS - 3 Downstream Steam	112 t/h	140 t/h
Critical Plant Parameters		
Tie Power	0 MW	0 MW
HP Steam Header Pressure	50 kg/cm^2	50 kg/cm^2
MP Steam Header Pressure	14 kg/cm^2	14 kg/cm^2
MP Steam Header Pressure	5 kg/cm^2	5 kg/cm^2
Process Power and Steam Loads		
Process Power Load	48 MW	
Process HP Steam Load	50 t/h	
Process MP Steam Load	138 t/h	
Process LP Steam Load	140 t/h	

Table 7.1 shows the plant initial conditions and the plant optimal operating point. The fuzzy controller is designed to take the plant from its initial condition to its optimal operating point while maintaining the critical plant parameters shown in Table 7.1. The supervisory control used to control the plant consists of nine individual fuzzy controllers; three for gas turbines, three for boilers and three for PRDSs. GT-1 and GT-3 were operated in generator set point control mode while GT-2 was put in the tie power control mode. Boiler-1 and Boiler-3 were operated in boiler set point control mode while Boiler-3 was put in the header pressure control mode. All the PRDS's were put

in PRDS set point control mode. The simulation ran for a time span 1500 s in the real time plant model developed in SIMULINK. The results are shown below.

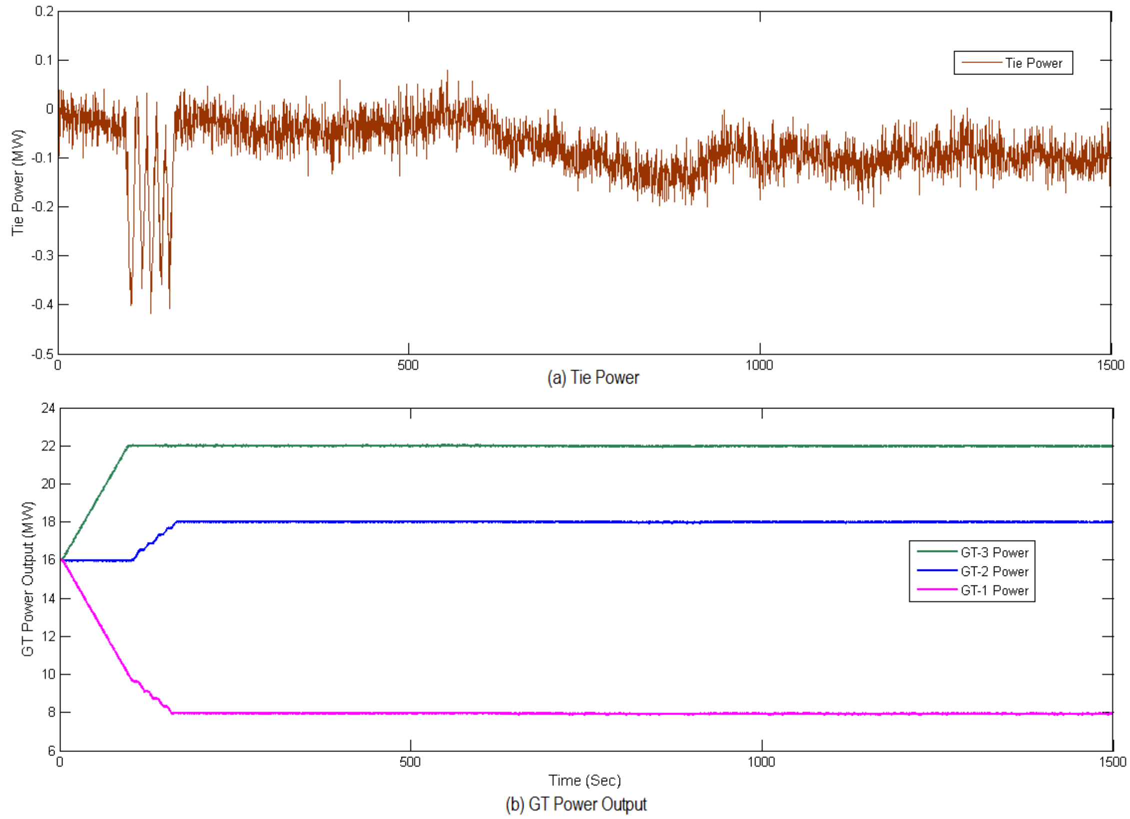


Figure 7.5 - Tie and GT Power

Figure 7.5 shows the tie and GT power changes during the simulation. It is observed that from the initial conditions of 16 MW GT-3 reaches its optimal operating point in about 100 s. During this time GT-1 load is going down, but with no generator able to match the power reduction the tie power also begins to go negative as seen in Fig. 7.5(a). Now GT-2 that is in tie power control comes into play and increases its power output. Thus as GT-1 lowers its output GT-2 increases its own to keep the tie power at a constant level. All the GTs reach their optimal operating point in about 180 s. It is observed that the maximum deviation in the tie power is only 0.4 MW that is well within the required control range.

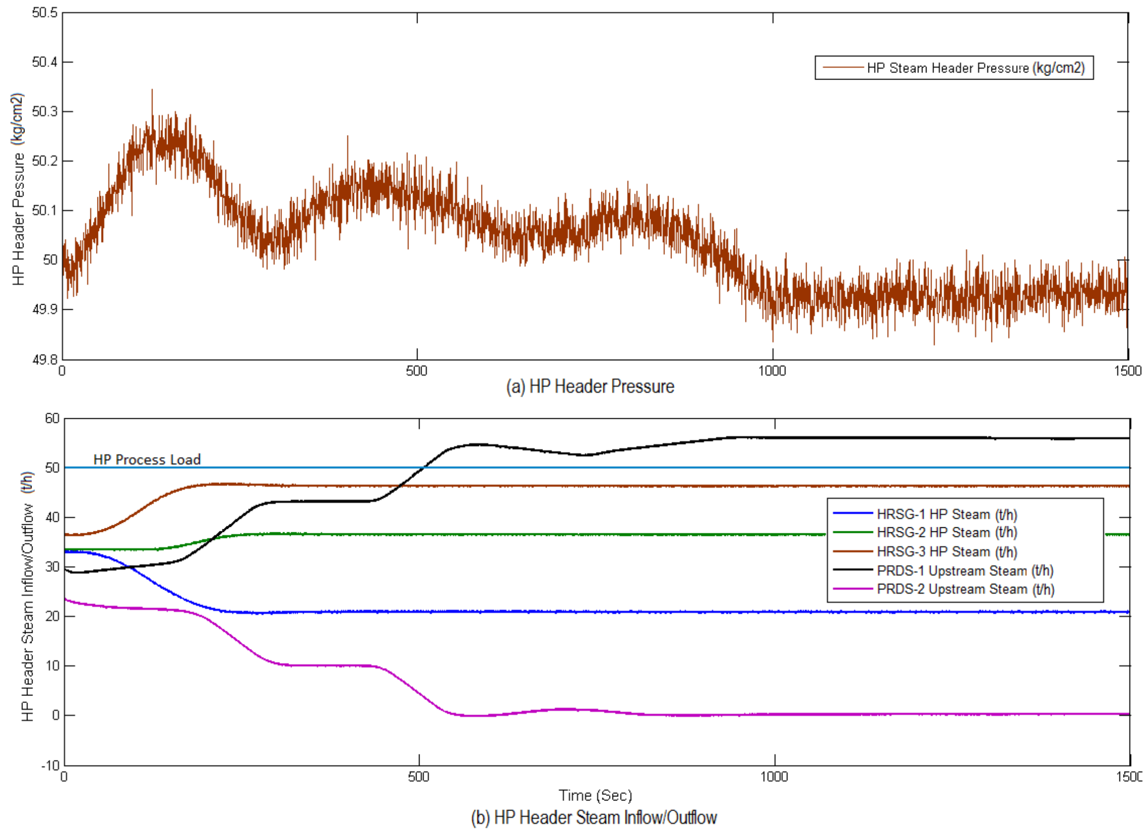


Figure 7.6 - MP Header Flow

Figure 7.6 shows the HP header pressure and the major steam inflow and outflow from the HP header. HRSG-1, HRSG-2 and HRSG-3 HP steam represent the major steam inflow into the header, while PRDS-1 and PRDS-2 upstream flow and HP process load represent the major steam outflow. It is observed that the HRSG steam follows the GT power variations and settles down in less than 240 s. The PRDS flow is usually dependent on the downstream pressure regulation and settles down at around 1000 s. About the same time the header pressure also stabilises at 50 kg/cm^2, which is the required operating point of the header. It is observed that though there are significant changes in inflow and outflow, the maximum header pressure variation is positive 0.25 kg/cm^2, that is well within the required operational limits.

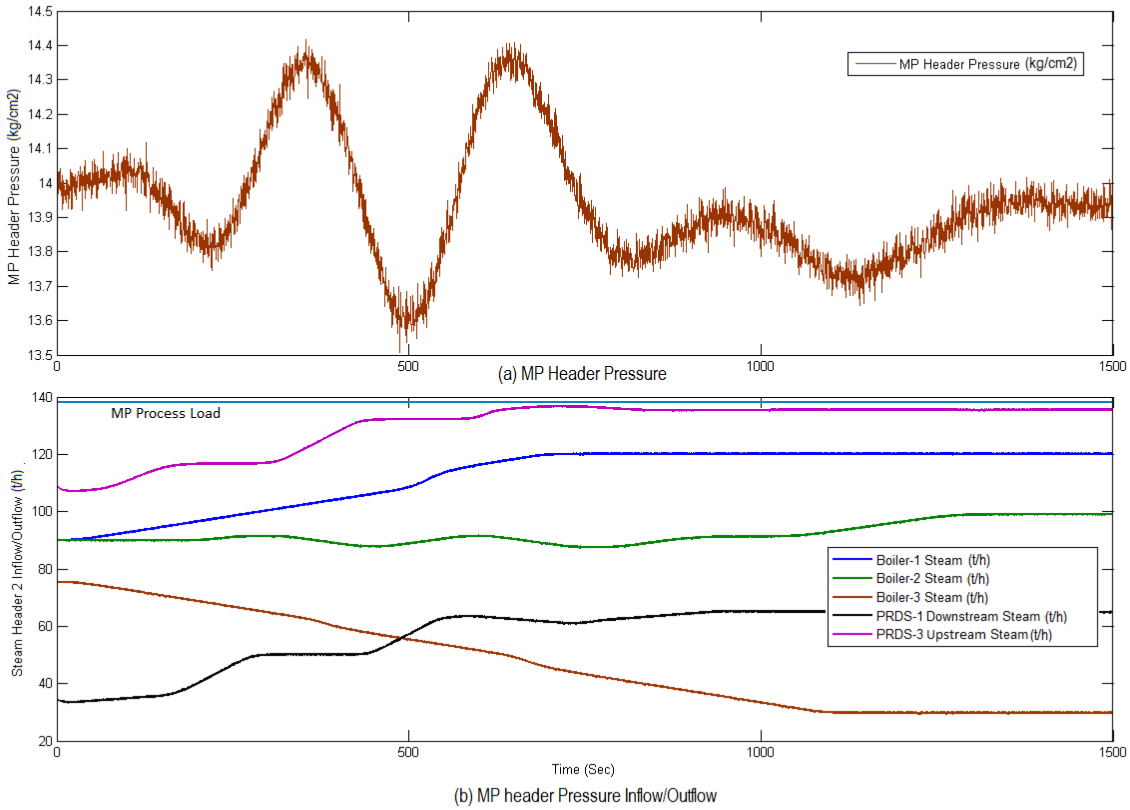


Figure 7.7 - MP Header Pressure and Flow

Figure 7.7 shows the MP steam header pressure and the major inflow and outflow from the MP header. Boiler-1, Boiler-2, Boiler-3 and PRDS-1 Downstream steam represent the major inflow while MP process steam load and the PRDS-3 Upstream Steam represent the major outflow from the MP header. It can be seen that Boiler-1 and Boiler-3 swing to their optimal set point and reach their optimal set points in 800 s and 1200 s, respectively. It can be observed that Boiler-2 that is in header pressure control mode increases its output when the header pressure goes low and reduces its output when the header pressure goes high. As the MP process load remains the same throughout, Boiler-2 finally settles down at its optimal operating point at around 1400 s. About the same time the MP header pressure settles down at 14 kg/cm^2 which is the required operating point. It is seen that even with large steam swings in the boilers and the PRDS the controllers work well and the critical steam pressure is controlled well. During the whole simulation period the MP header pressure had a maximum positive deviation of 0.4 kg/cm^2 and maximum negative deviation of 0.4 kg/cm^2; that is well within the required operational limits.

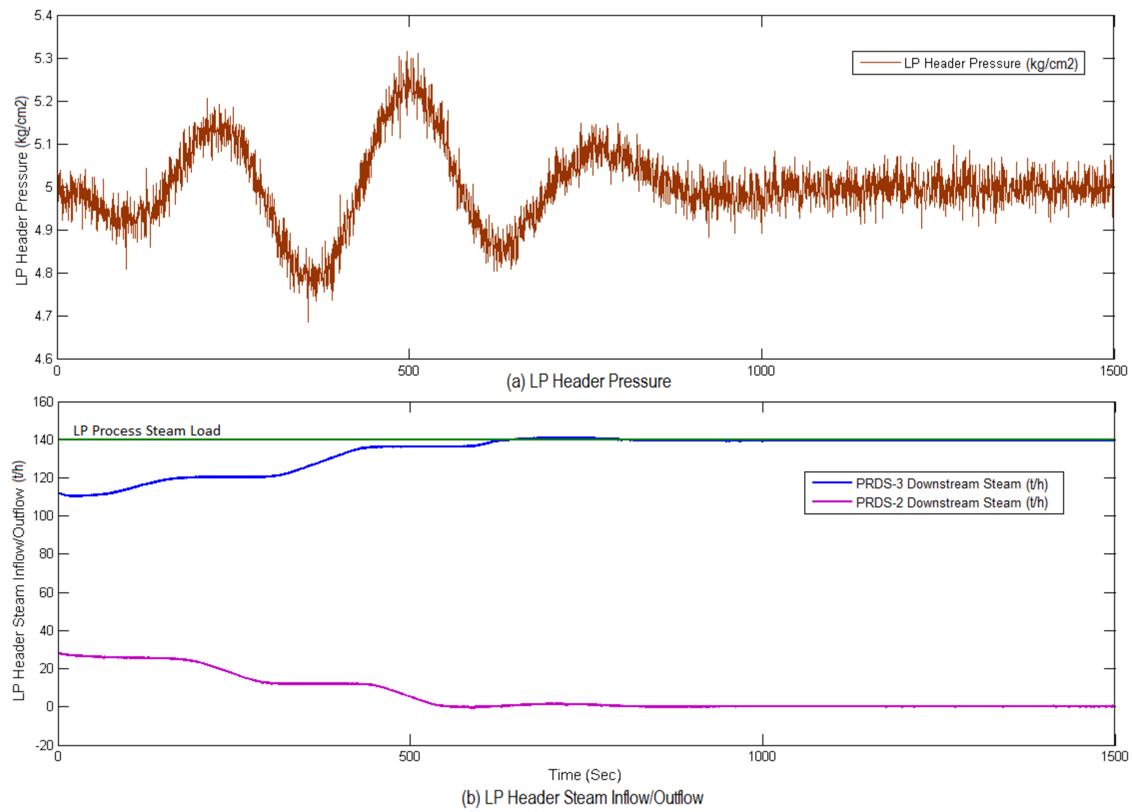


Figure 7.8 - LP Header Pressure and Flow

Figure 7.8 shows the LP header pressure and the major inflow and outflow from the LP steam header. PRDS-2 and PRDS-3 downstream steam represents the major inflow into the header while LP process steam load represents its major outflow. It is observed that the PRDSs reach their steady state conditions in about 800 s and at about the same time the header pressure also stabilises to 5 kg/cm^2, the required operating point. It is seen that the PRDS controllers work well and the maximum positive deviation in header pressure is only about 0.25 kg/cm^2 and maximum negative deviation is about 0.2 kg/cm^2; that is well within the operational limits.

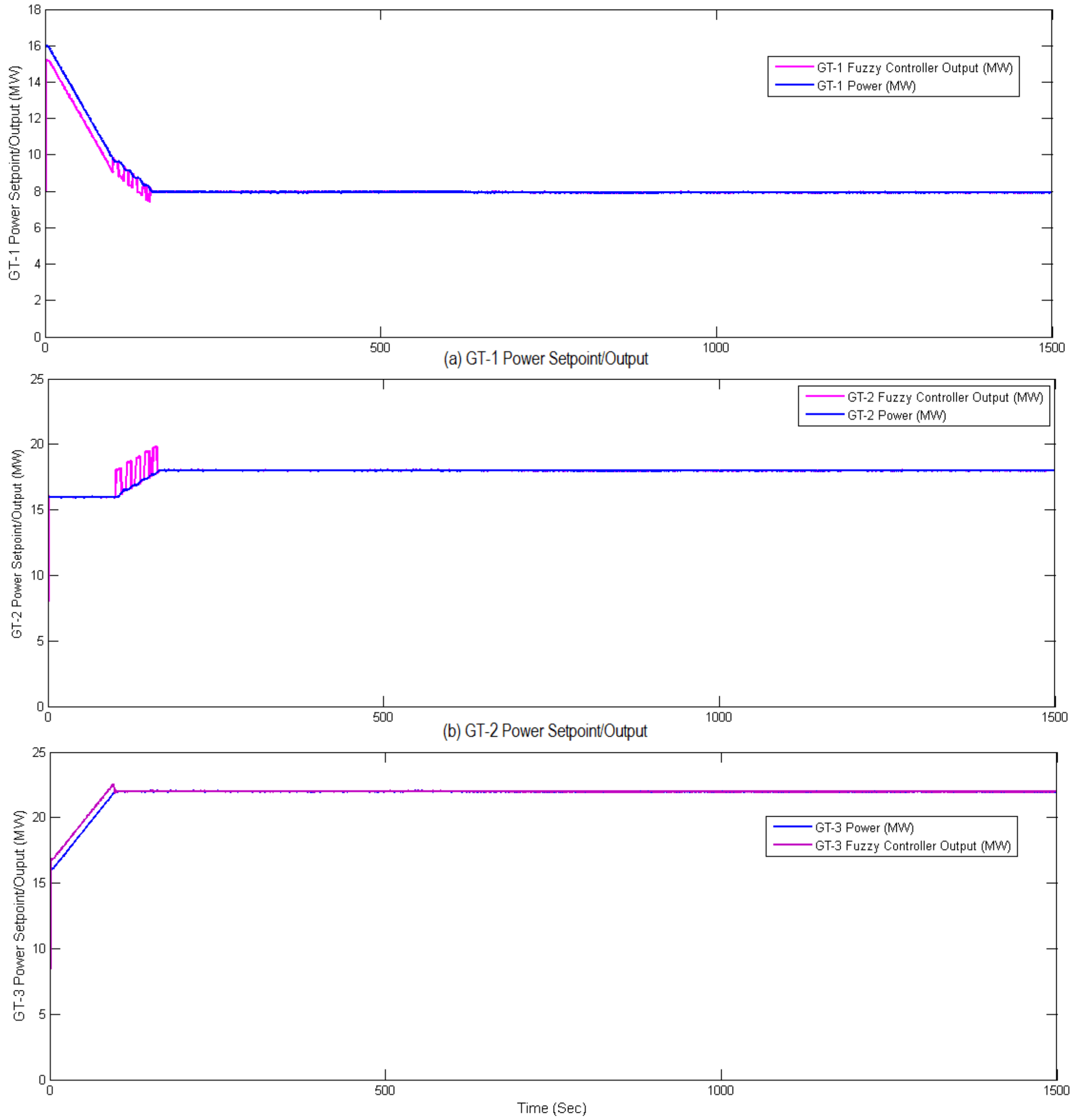


Figure 7.9 - Fuzzy Controllers v/s Real Time Output (Generators)

Figure 7.9 shows the GT fuzzy controller output and GT real time output during the simulation studies. It is observed that the controller performs to expectations, and closely matches the real time output of the generators. In GT-2 which is put in tie power control mode, the fuzzy controller output has peaks as it was designed to tackle large tie power errors.

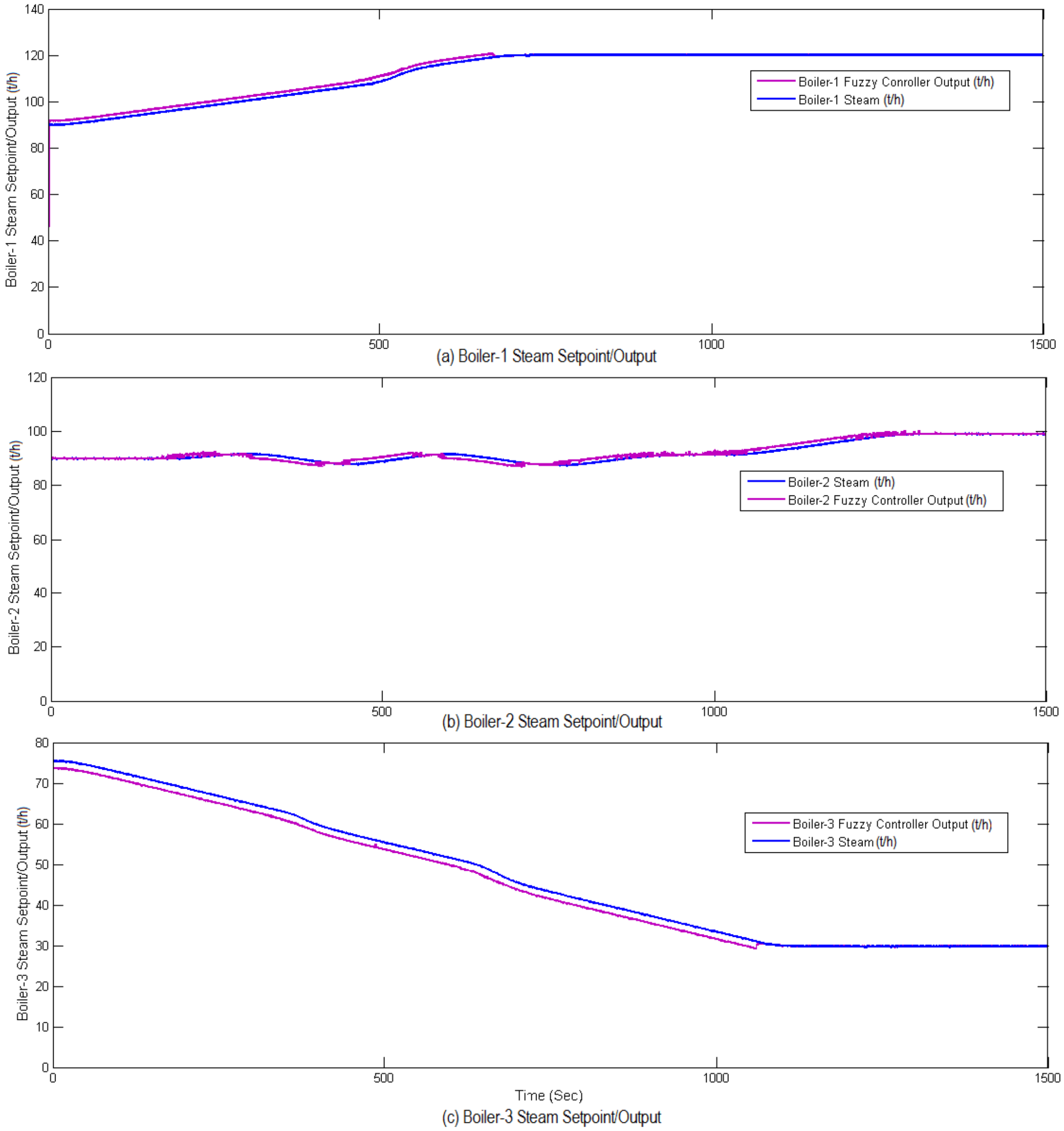


Figure 7.10 - Fuzzy Controller v/s Real Time Output (Boilers)

Figure 7.10 shows the Boiler fuzzy controller outputs and real time boiler output during the simulation studies. Here too the controllers provide well matched outputs. Minimal amount of control jitter can be found in Boiler-2 graph, but that is due to the fact that the header pressure measurement had a high level of Gaussian noise and the controller was designed with minimal dead bands.

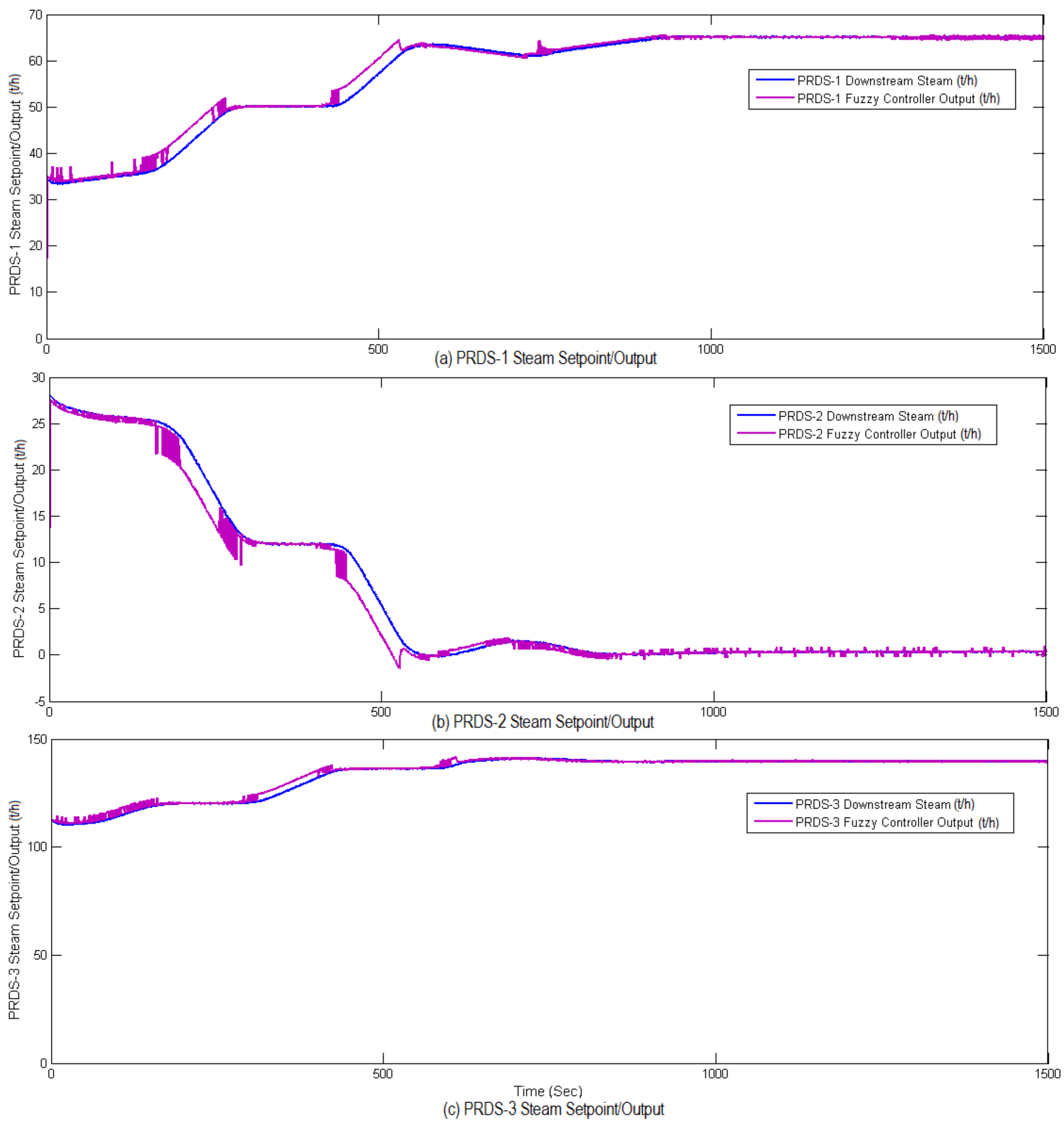
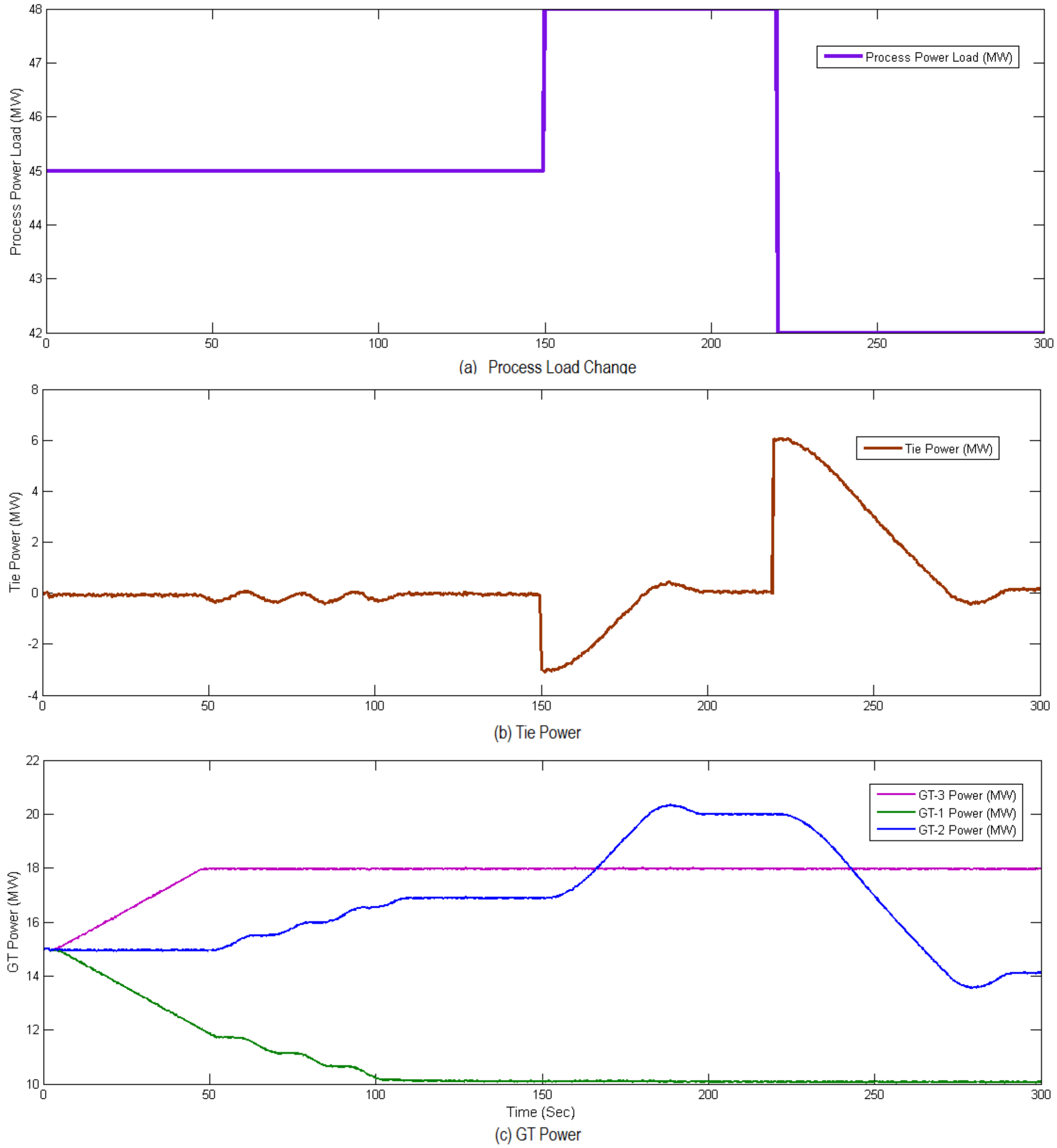


Figure 7.11 - Fuzzy Controller v/s Real Time Output (PRDS)

Figure 7.11 shows the PRDS fuzzy controllers' outputs and real time PRDS output during the simulation studies. Here too minimal amount of control jitter can be seen. This occurs due to the high measurement noise and controller working at a gradient change point in the control surface. This can be eliminated by using a rate limiter at the output, a better quality sensor or by filtering the measurement noise off the input.

7.3 Simulation with Changes in Power Loads

This section showcases the controller behaviour when a process power load change takes place in between optimization runs.



(a) Process Load Change

(b) Tie Power

(c) GT Power

Figure 7.12 - Simulation with Power Load Change

Figure 7.12 shows (a) changes in process load, (b) the tie power and (c) the GT power swing during the simulation with varying process power load. From the initial condition where the process load is 45 MW and all the GT's are loaded to 15 MW each, GT-1 reduces to 10 MW , GT-2 increases to 17 MW and GT-3 increases to 18 MW in accordance with the optimized set point values. In the simulation GT-1 and GT-3 operate in set point control mode while GT-2 is put in the tie power control mode. At time 150 s the process load suddenly increases to 48 MW; a process load increase of 3 MW and the tie power goes to an import of 3 MW. Now the GT-2 takes action and increase its power output from 17 MW to 20 MW, pushing the tie power back to the required operating value of 0 MW in about 40 s. At time 220 s the process load falls to 42 MW and the tie power goes to an export of 6 MW. Now GT-2 reduces its power to 14 MW to push the tie power to zero MW in about 70 s. During the load changes it is observed that the generators that are in the power set point control do not come into play and maintain their optimal operating point while GT-2 takes care of the load changes.

7.4 Simulation with Changes in Steam Loads

This section showcases the controller behaviour when a steam load change takes place in between optimization runs. In this simulation, it is assumed that HP and LP process steam loads remain constant while only the MP steam process load varies.

Figure 7.13 shows the (a) MP process steam load, (b) the MP header pressure, and (c) the boiler steam outputs. At 1250 s the steam load suddenly increases to 155 t/h. The header pressure begins to drop and falls to a maximum negative deviation of 0.8 kg/cm2. This time the Boiler-2 unit that is put in the header pressure control mode, increases its steam output and arrests the fall in header pressure. The header pressure starts increasing and the steam inflow into the header is increased. At time 1750 s the process load falls to 145 t/h and this causes the steam pressure to shoot-up. This causes Boiler-2 to reduce its steam load, the rise in steam pressure is arrested and starts reducing. Boiler 2 reduces its load until the steam pressure reaches its operating point of 14 kg/cm^2. Both Boiler-2 and steam pressure stabilize at around 2200 s.

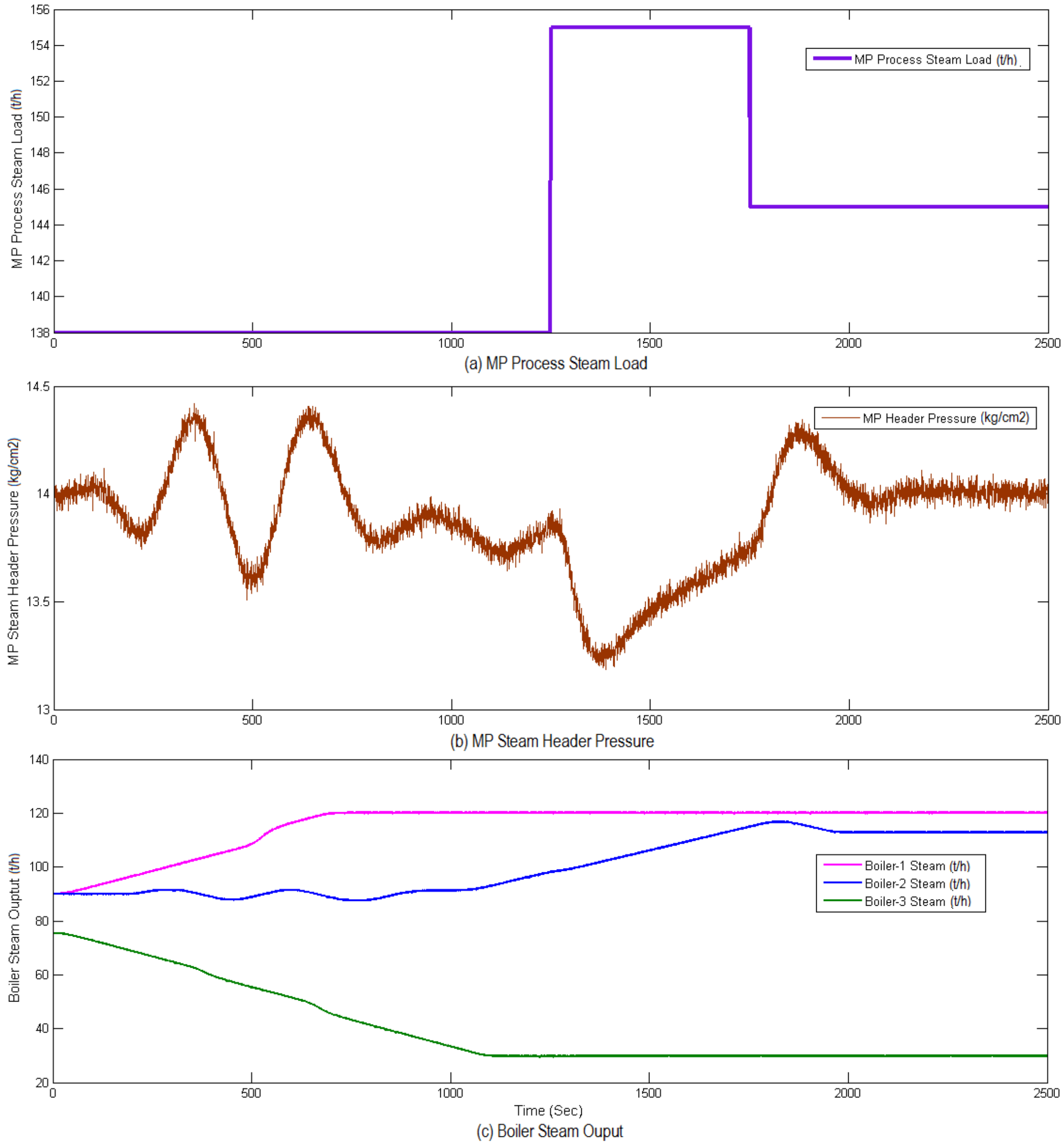


Figure 7.13 - Simulation with Changes in Steam Load

7.5 Analysis and Observations

It is seen from the simulation studies in section 7.1.1 , that a constant setpoint control does not provide adequate control and a full-fledged supervisory control may be required to keep the plant critical parameters safe under all operating condition. Simulation studies in section 7.1.2 show that the heuristic control is a step in the right direction as it uses expert knowledge to control plant. It settles the plant well and controls the plant parameters within limits but suffers from being difficult to work with as it requires considerable testing to get a correct gain setting. In addition it does not have the ability to handle plant disturbances. The fuzzy supervisory control eliminates the shortcomings of the heuristic controller by being easy to work with and handles plant disturbances. Salient features of the controller are explained below.

It is observed in the simulation run that the controller performs quite well when transferring the plant from the initial operating condition to optimal operating point. Though the generator and boiler output changes were significant, the controllers managed to stabilize the plant in about 20 m (1200 s). It was also seen that even though as many as nine different controllers in more than 5 different operational modes were used, the controllers worked in tandem to achieve the control objectives. Not only did the plant reach optimal operating conditions in a short period of time, the critical plant parameters were also maintained.

The generator set point controller takes the GTs to their operating point within 180 s and the tie line power controller aided by the generator set point controller keeps the maximum deviation in the tie line power to a bare minimum of -0.4 MW. The boiler set point controller took the boilers to their optimal operating point in about 1200 s; this 1200 s time interval is needed due to the large change in the steam outputs. Even with the huge steam inflow/outflow changes, the header pressure controller, used to control the MP header, works really well with a maximum positive and negative deviation of 0.4 kg/cm2; a 2.8% change from normal operating condition. The PRDS set point controller also does well both taking the PRDS to its operating point as well as maintaining the downstream header pressure. This can be seen in LP header where both PRDS-2 and PRDS-3 were controlling the header pressure. The LP header pressure had a maximum positive and negative deviation of 0.2 kg/cm2; a 4% change from nominal operating condition. It is also seen that the maximum positive and negative pressure deviation in the HP header was about 0.2 kg/cm2; a 0.4% change from nominal operating condition.

The controller also performed well in plant disturbance tests. When the plant power load was changed, the tie controller came into action and took the tie power from a positive or negative import condition to zero tie power level. Similarly when the process steam load was changed, the header pressure controller came into action and maintained the header pressure at its nominal operating value. Thus the control scheme maintained the plant critical parameters even during process load changes.

7.6 Software Architecture for Optimization and Control

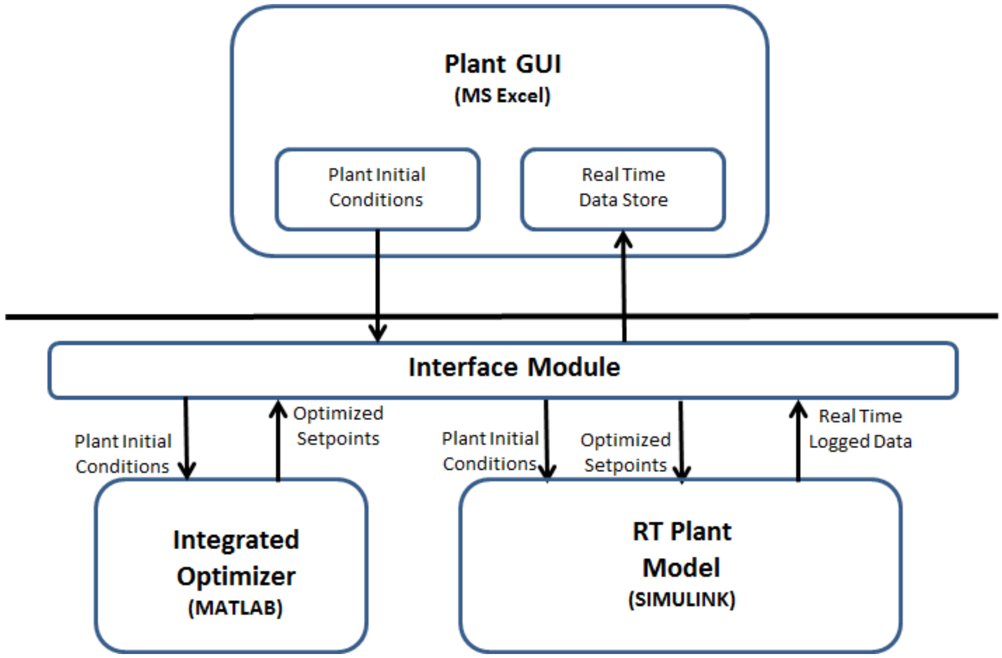


Figure 7.14 - Software Architecture for Optimization and Control

Figure 7.10 shows the architecture used for developing optimization and control software in this thesis. The software developed in this thesis can be divided into four main modules

- Plant Graphical User Interface (GUI)

  Plant Graphical User Interface (GUI) was developed in Excel/VB.Net. The GUI provides the user with a graphical overview of the plant operations. This module is designed similar to an operator interface in any modern process/power plant. The user can input plant initial conditions like steam load in headers and process power load through this GUI. In addition parameters used in optimization like Fuel cost and Water cost are also entered here. The Plant GUI also forms the repository for the logged plant data that comes from the real time simulation model.

- Integrated Optimizer

  The integrated optimizer runs the three stage optimization algorithm and provides the optimal operating points for the generators and boilers depending on the initial conditions and operating costs specified in the Plant GUI. The integrated optimizer is coded in MATLAB and uses MATLAB Optimization Toolbox.

- Real Time Plant Model

  The real time plant model is a SIMULINK model file in which the total plant is built and simulated. The fuzzy control of the plant is also implemented in SIMULINK. The fuzzy controller gets the optimal set points from the integrated optimizer and takes the plant from the initial operating condition specified in the Plant GUI to the optimal operating point. Once the simulation is complete it sends the real time logged data to be used in the Plant GUI.

- Interface Module

  The interface module is the link between the three modules. It is coded in MATLAB and controls the actions of the other software modules. It receives and transfers the plant initial conditions to the integrated optimizer. It obtains the optimal set points from the integrated optimizer and transfers them to the real time plant model in SIMULINK. Once the simulation is complete it accumulates the logged data and passes it to the Plant GUI.

Once the "START SIMULATION" button is pressed in the Plant GUI the following events occur.

- The interface module takes the plant initial conditions and transfers it on to the integrated optimizer and calls the optimization run.
- The integrated optimizer optimizes the plant operation and returns the Optimized set point to the Interface Module.
- The interface module transfers data regarding plant initial conditions and optimized set points for plant operation to the real time plant model and calls the simulation run.
- Real time plant model completes the simulation run and transfers the logged data back to the interface module.
- The interface module sends this logged data to the real time data repository in the plant GUI.

The real time data is displayed in the Plant GUI and is updated every two seconds.

Chapter Eight: Conclusions

Optimization of a cogeneration system is a complex mathematical and control problem. At the beginning of the research there were two main objectives

- Development of an integrated optimizer that will address the question of equipment selection and operation under specified constraints and diverse operational scenarios
- Development of a high level control system that will take the plant to the optimal operating point as specified by the optimizer.

In line with the objectives, the research work was divided into two parts. The optimization part of the research is focussed on the development of the optimizer while the control part is focussed on the development of a high level control system. During the course of the research all the initial objectives have been met.

8.1 Conclusions

Modelling of the cogeneration systems is accomplished, with detailed modelling and validation of ten types of plant equipment and headers. This is important as the models developed in the thesis can be reused in other research with changes only to the model parameters. In addition, the models give an insight into how models for other cogeneration equipment can be developed.

The framework developed for modelling is extended to support automated problem formation. This is a significant step in the development of commercial optimization software. Automated problem formulation helps the software to develop optimization equations from the plant configuration. Thus the end user need not have high level of knowledge on optimization, and only needs to know how the plant equipment and various modules are interconnected. This simplifies setting up of the optimization problem and facilitates rapid deployment.

The optimization is formulated as a mixed integer nonlinear programming problem. The problem of equipment selection is done using integer variables that represent running status. Continuous variables that represent generator and boiler load handle the problem of how to run the selected equipment. A novel three stage algorithm has been developed that solves the problem efficiently. Though the algorithm developed has a very specific usage and is valid only for the given problem, the results provide a strong support to the argument that it is often desirable to customize the MINLP solver for a specific application in order to achieve good computational performance.

The significance of the three stage algorithm lies not only in its speed but also in its ability to solve the optimization problem under various operating conditions. It not only solves for minimum cost but can also be extended to include power and steam reliability criterion. A natural extension of the three stage algorithm is able to handle multiple headers that solves the problem of islanded header operation. Thus an integrated optimizer that uses the three stage algorithm can not only be used as an online optimization tool due to its speed, but can also be used for offline operational analysis and maintenance planning as it is able to give solutions under various operating conditions.

The dynamic modelling of plant equipment and headers has also been accomplished to create a virtual cogeneration system that has been used to test the plant high level control. As with optimization models, the dynamic models developed here can be reused in future research work.

The supervisory fuzzy set point controller performed to expectations by successfully tracking the optimal set points in a complex cogeneration plant model. The control matched or exceeded the performance parameters and most importantly maintained critical plant parameters. The importance of the fuzzy controller lies in its ease of use and scalability. As the fuzzy controller works with predetermined rule base it does not need any system identification tests, thus significantly reducing deployment time and engineering costs. As the main supervisory control is made of independent sub-controllers with specific task and deployment equipment, the control can be easily scaled depending on the type and no of equipment present in the plant. In addition the fuzzy supervisory control being a set point controller can be easily integrated into the existing plant DCS or control equipment.

During the course of the project a comprehensive software library coded in MATLAB has been developed for performing the optimization. This formed the core of the integrated optimizer module that could handle any plant configuration with the supported plant equipment and headers. A specific data format that defines the plant equipment, headers and interconnections has also been developed, and is presently used by the software for automated constraint listing. The SIMULINK environment is used for dynamic simulations and testing the fuzzy supervisory controller. A graphical user interface is also created to simulate a real life operator station. The control and software architecture developed for this project can be easily applied to present day cogeneration systems.

8.2 Future Work

The project provides a good ground for future work. The equipment modeled forms only a part of the present cogeneration systems. Cogeneration equipment like back pressure steam turbines, extraction steam turbines, diesel engines, and auxiliary equipment, like cooling towers and DM plants, can also be modeled using the developed framework and easily incorporated into the existing software. In addition to this, the simulation platform developed for this project can serve to test new artificial intelligence and expert system algorithms used for process monitoring and control.

References

[1] Nag P.K; “Power Plant Engineering”, Tata McGraw Hill, New Delhi, 2007.

[2] Bogdan, Z.; Baburic, M.; Serman, N.; "Optimization model for EL-TO zagreb cogeneration plant," Proceedings of the 23rd International Conference on Information Technology Interfaces, vol. 1, 2001, pp. 287-295.

[3] Moslehi, K.; Khadem, N.; Bernal, R.; Hernandez, G.; "Optimization of multiplant cogeneration system operation including electric and steam networks," IEEE Transactions on Power Systems, vol. 6, 1991, pp. 484-490.

[4] Firestone, R.; Stadler, M.; Marnay, C.; "Integrated energy system dispatch optimization”, IEEE International Conference in Industrial Informatics, 2006, pp. 357-362.

[5] Garcia, C.E.; Prett, D.M.; Morari, M.; “Model Predictive Control: Theory and Practice - A Survey”, Automatica, vol. 25, 1989, pp. 335–348.

[6] Mayne, D.Q.; Rawlings, J.B.; Rao, C.V.; Scokaert, P.O.M.; “Constrained model predictive control: Stability and optimality”, Automatica, vol. 36, 2000, pp.789–814.

[7] Camacho, E.F,; Bordons, C.; “Model Predictive Control in the Process Industry”, Springer: Berlin, 1999.

[8] Maciejowski, J.; “Predictive Control with Constraints”, Prentice-Hall: Englewood Cliffs, NJ, 2001.

[9] Qin, S.J.; Badgwell, T.A.; “An overview of industrial model predictive control technology”, Fifth International Conference on Chemical Process Control, 1996, pp. 232–256.

[10] Lu, S.; Hogg, B.W.; “Predictive co-ordinated control for power-plant steam pressure and power output”, Control Engineering Practice, vol. 5(1), 1997, pp.79–84.

[11] Andrzej Kr´olikowski,; “Adaptive Generalized Predictive Control Subject to Input Constraints”, Proceedings of the 7th Mediterranean Conference on Control and Automation 1999, pp.54-61.

[12] Andrzej Krolikowski; Damian Jerzy; “Self-Tuning Generelised Predictive Control With Input Constraints”, Int. J. Appl. Math. Comput. Sci., vol. 11(2), 2001, pp. 459-470.

[13] Prasad, G.; Swidenbank. E.; Hogg, B.W.; “A neural net model-based multivariable long-range predictive control strategy applied in thermal power plant control”, IEEE Transactions on Energy Conversion, vol. 13(2), 1997, pp.176–182.

[14] Prasad, G.; Irwin, G.W.; Swidenbank, E.; Hogg, B.W.; “Plant-wide predictive control for a thermal power plant based on a physical plant model”, IEEE Proceedings on Control Theory Applications vol. 147(5), 2000, 523–537.

[15] Yukita, K.; Goto, Y.; Mizuno, K.; Miyafuji, T.; Ichiyanagi, K.; Mizutani, Y.; , "Study of load frequency control using fuzzy theory by combined cycle power plant," *Power Engineering Society Winter Meeting, 2000. IEEE* , vol.1, 2000, pp.422-427.

[16] Pruessmann, D.; Krause, B.; von Altrock, C.; "Fuzzy logic supervisory control for coal power plant," *Fuzzy Systems, 1997, Proceedings of the Sixth IEEE International Conference on*, vol.2, 1997, pp.921-925.

[17] Garduno-Ramirez, R.; Lee, K.Y.; "Wide range operation of a power unit via feedforward fuzzy control [thermal power plants]," *IEEE Transactions on Energy Conversion*, vol. 15(4), 2000, pp.421-426.

[18] Garduno-Ramirez, R.; Lee, K.Y.; "Fuzzy scheduling control of a power plant," Power Engineering Society Winter Meeting, 2000. IEEE, vol.1, 2000, pp.441-445.

[19] Brown, M.; Harris, C.J.; "Neurofuzzy Adaptive Modelling and Control". Englewood Cliffs, NJ: Prentice-Hall, 1994.

[20] Liu, X.J.; Lara-Rosano, F.; Chan, C.W.; "Neurofuzzy network modelling and control of steam pressure in 300MW steam-boiler system". Engineering Applications of Artificial Intelligence, vol. 16(5), 2003, pp. 431-440.

[21] Sean Goodhart; Jun Nishizawa; Kunihisa Yano; Hiroyuki Yada; "Advanced control in cogeneration utility management",Computing and Computer Engineering Journal, vol. 1, 2000, pp.273-282.

[22] Duran, M.A.; Grossmann, P.D.; "An Outer-Approximation Algorithm for a Class of Mixed-integer Nonlinear Programs", Math. Program., vol. 36,1986, pp.307–339.

[23] Fletcher, R.; Leyffer; S.; "Solving Mixed Integer Programs by Outer Approximation", Math. Program., vol. 66, 1994, pp. 327–349.

[24] Gupta, O.K.; Ravindran, A.; "Branch and Bound Experiments in Convex Nonlinear Integer Programming", Manage Sci., vol. 31(12), 1986, pp. 1533–1546.

[25] Quesada, I.; Grossmann, I.E.; "An LP/NLP Based Branch and Bound Algorithm for Convex MINLP Optimization Problems", Computers Chem. Eng., vol. 16(10/11), 1992, pp. 937–947.

[26] Westerlund, T.; Petersson, F.; "A Cutting Plane Method for Solving Convex MINLP Problems", Computers Chem. Eng., vol. 19, 1995, pp. 131–136.

[27] Geoffrion, A.M.; “A Generalized Benders Decomposition”, J. Optim. Theory Appl., vol.10 (4), 1972, pp. 237–260.

[28] Bragalli C.; D’Ambrosio, C.; Lee, J.; Lodi, A.; Toth, P.; “Practical water network design by MINLPs”, Technical Report OR-09-21, University of Bologna, 2009.

[29] Bussieck, M.R.; “Introduction to GAMS Branch-and-Cut Facility”, Technical report, GAMS Development Corp., 2003.

[30] Kirschen, D.S.; "Do investments prevent blackouts?" , IEEE Power Engineering Society General Meeting, 2007, pp. 1-5.

[31] Schlueter, M.; Gerdts, M.; Rückmann J.J.; “A Numerical Study of MIDACO on 100 MINLP”, Benchmarks Optimization, Taylor & Francis, Vol 61(7), 2012, pp.873-900.

[32] Schlueter, M.; Gerdts, M.; “The oracle penalty method”, Journal of Global Optimization, Vol. 47(2), 2012, pp.293-325.

[33] Schlueter, M.; Egea, J.A.; Banga, J.R.; “Extended ant colony optimization for non-convex mixed integer nonlinear programming”, Computers & Operations Research, Vol. 36(7), 2009, pp.2217-2229.

[34] Tony Kealy; Aidan O’Dwyer; “Closed Loop Identification Of A First Order Plus Dead Time Process Model Under PI Control”, Proceedings of the Irish Signals and Systems Conference, University College, 2002, pp. 9-14.

APPENDIX A: REFINERY COGENERATION UNIT

The Refinery Cogeneration unit equipment and header data are given in tables A.1 and A.2, while Figures A.1 and A.2 show the plant configuration.

Table A. 1 - Refinery Cogeneration Unit Equipment List

EQUIPMENT DATA			
Sl No	**Name**	**Manufacturer**	**Comments**
1	Gas Turbine - 1	GE/BHEL	Type : Frame 5EA unit Fuel : Duel Fuel (Naptha/HSD) Rating : 22 MW at 35 deg C
2	Gas Turbine - 2	GE/BHEL	
3	Gas Turbine - 3	GE/BHEL	
4	HRSG - 1	BHEL	Type : Twin Drum Heat recovery unit Rating (Drum-1) : 56.9 kg/cm^2, 438 deg C, 46 t/h Rating (Drum-2) : 20 kg/cm^2, 238 deg, 7 t/h
5	HRSG - 2	BHEL	
6	HRSG - 3	BHEL	
7	Boiler - 1	BHPV	Type : Oil/Gas fired boiler unit Fuel : Refinery Fuel oil Rating : 18 kg/cm^2, 252 deg C, 120 t/h
8	Boiler - 2	BHPV	
9	Boiler - 3	BHPV	
10	Deaerator - 1	BHEL	Rating : 3.5 kg/cm^2, 120 dec C, 40 m3
11	Deaerator - 2	BHEL	
12	PRDS - 1	Emerson	Type : Fisher D CL 600
13	PRDS - 2	Emerson	Type : Fisher D CL 150
14	PRDS - 3	Emerson	

Table A. 2 - Refinery Cogeneration Unit Header List

HEADER DATA		
Sl no	**Name**	**Comments**
1	Water Header	DM water at 5.5 kg/cm^2
2	BFW Header - 1	Boiler feed water at 80 kg/cm^2
3	BFW Header - 2	Boiler feed water at 36 kg/cm^2
4	BFW Header - 3	Boiler feed water at 42 kg/cm^2
5	Fuel Header - 1	Naptha at 5 kg/cm^2
6	Fuel Header - 2	Refinery fuel oil at 7 kg/cm^2
7	Steam Header - 1	HP steam at 50 kg/cm^2 & 440 dec C
8	Steam Header - 2	MP steam at 14 kg/cm^2 & 240 dec C
9	Steam Header - 3	LP steam at 5 kg/cm^2 & 190 dec C

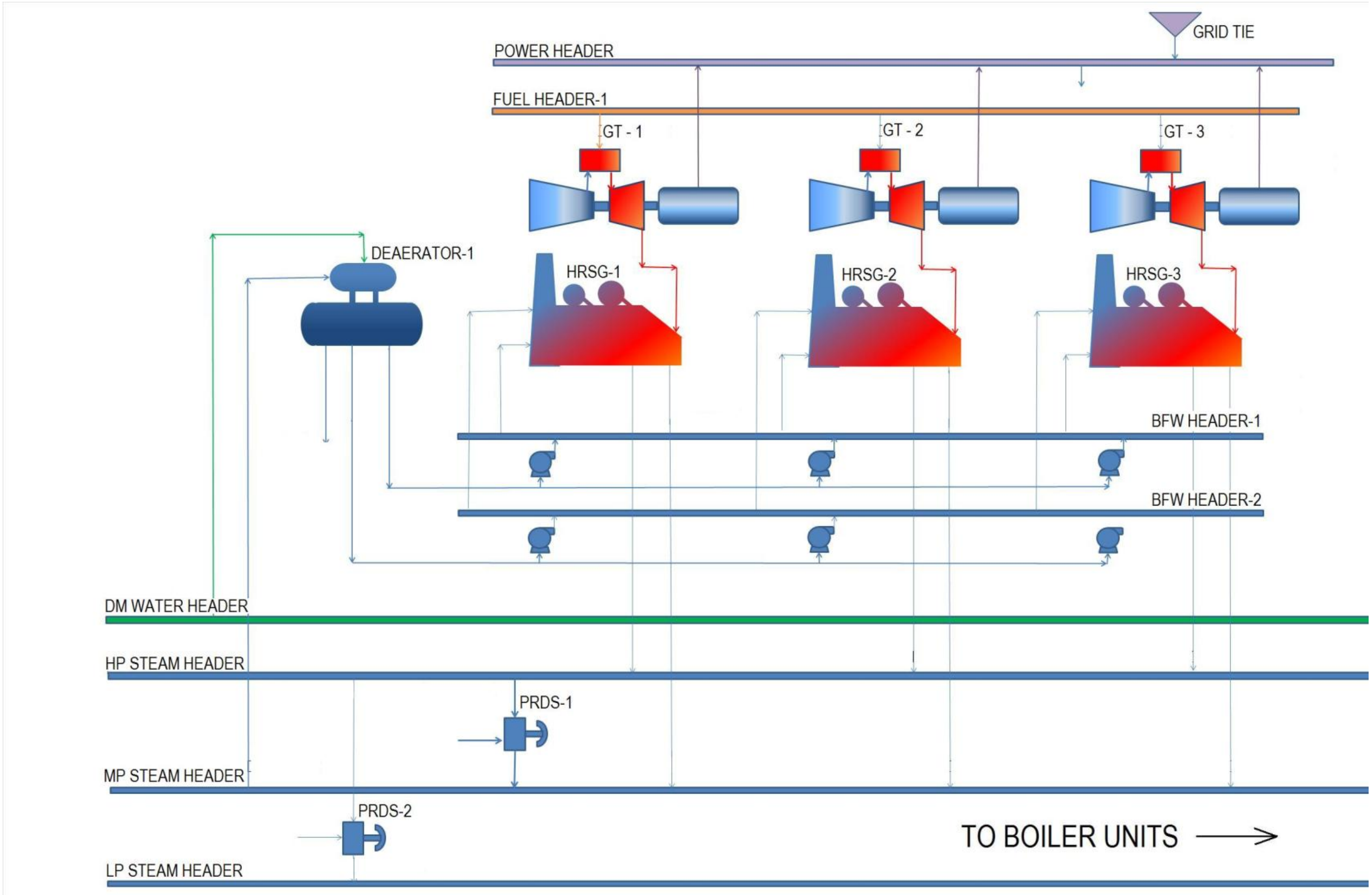


Figure A. 1 - Refinery Cogeneration Unit (Generation Block)

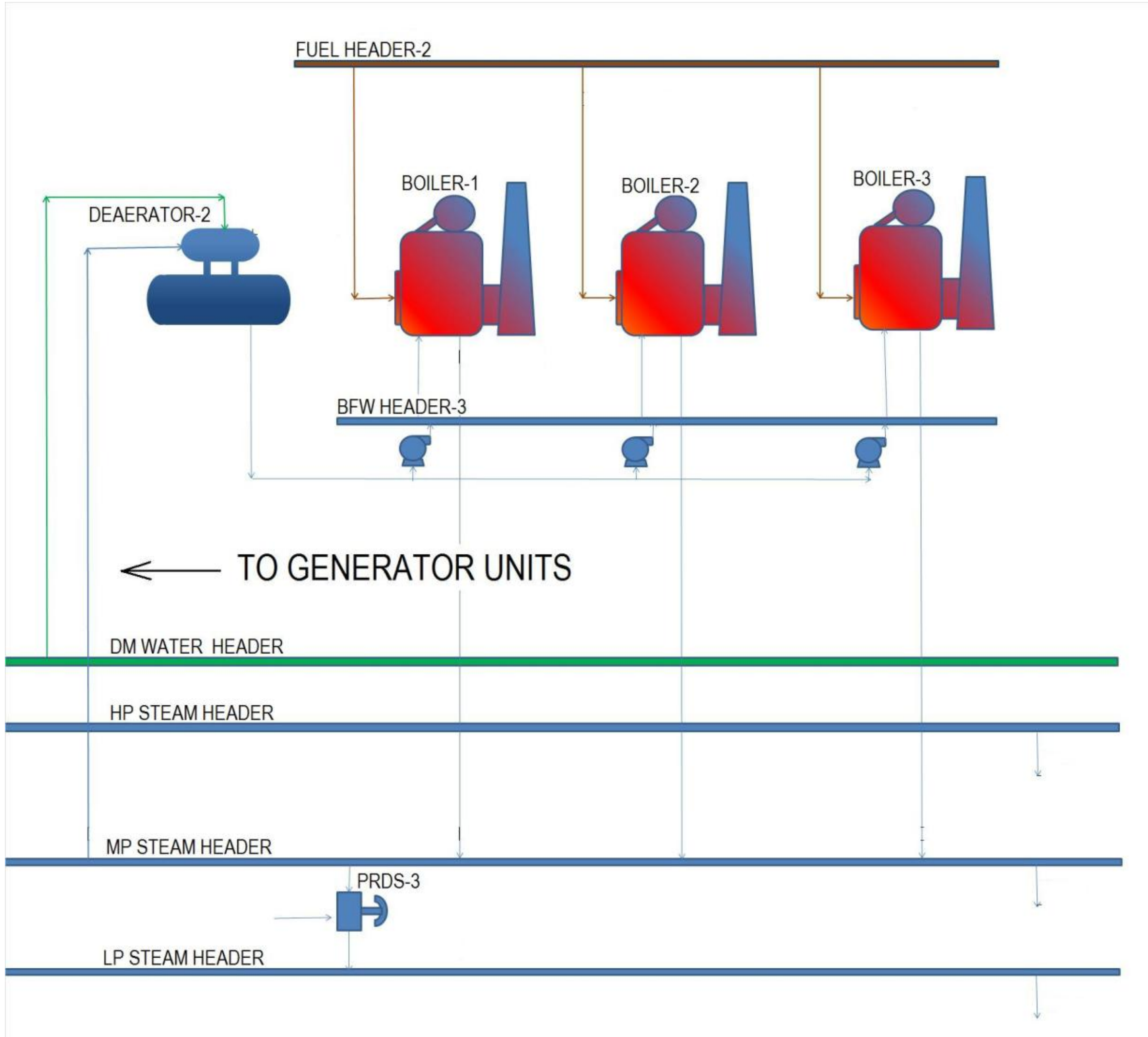


Figure A. 2 - Refinery Cogeneration Unit (Boiler Battery)

APPENDIX B: DISTRICT HEAT AND POWER SYSTEM

The district heat and power system consists of five identical power plants. Details of one such plant are given below. The equipment and header details are given in Tables B.1 and B.2, while Figures B.1 and B.2 show the plant configuration.

Table B. 1 - District Heat and Power Unit Equipment List

EQUIPMENT DATA			
Sl No	Name	Manufacturer	Comments
1	Gas Turbine - 1	GE/BHEL	Type : Frame 6FA unit Fuel : Duel Fuel (Gas/Naptha) Rating : 35 MW at 35 deg C
2	Gas Turbine - 2	GE/BHEL	
3	Gas Turbine - 3	GE/BHEL	
4	Gas Turbine - 4	GE/BHEL	
5	Gas Turbine - 5	GE/BHEL	
6	Gas Turbine - 6	GE/BHEL	
7	HRSG - 1	BHEL	Type : Single drum aux fired heat recovery unit Rating : 19.6 kg/cm^2, 264 deg C, 100 t/h
8	HRSG - 2	BHEL	
9	HRSG - 3	BHEL	
10	HRSG - 4	BHEL	
11	HRSG - 5	BHEL	
12	HRSG - 6	BHEL	
13	Boiler - 1	IJT	Type : Oil/Gas fired boiler unit Fuel : Refinery Fuel oil Rating : 57.5 kg/cm^2, 454 deg C, 90 t/h
14	Boiler - 2	BHPV	Type : Oil/Gas fired boiler unit Fuel : Refinery Fuel oil Rating : 18 kg/cm^2, 252 deg C, 140 t/h
15	Boiler - 3	BHPV	
16	Boiler - 4	BHPV	
17	Boiler - 5	BHPV	
18	Boiler - 6	BHPV	
19	Boiler - 7	BHPV	
20	Deaerator - 1	BHEL	Rating : 3.5 kg/cm^2, 120 dec C, 40 m3
21	Deaerator - 2	BHEL	
22	PRDS - 1	Emerson	Type : Fisher D CL 600
23	PRDS - 2	Emerson	Type : Fisher D CL 150
24	PRDS - 3	Emerson	

Table B. 2 - District Heat and Power Unit Header List

HEADER DATA		
Sl no	Name	Comments
1	Water Header	DM water at 5.5 kg/cm^2
2	BFW Header - 1	Boiler feed water at 36 kg/cm^2
3	BFW Header - 2	Boiler feed water at 80 kg/cm^2
4	BFW Header - 3	Boiler feed water at 42 kg/cm^2
5	Fuel Header - 1	Naptha at 5 kg/cm^2; Gas at 12 kg/cm^2
6	Fuel Header - 2	Refinery fuel oil at 7 kg/cm^2
7	Steam Header - 1	HP steam at 50 kg/cm^2 & 440 dec C
8	Steam Header - 2	MP steam at 14 kg/cm^2 & 240 dec C
9	Steam Header - 3	LP steam at 5 kg/cm^2 & 190 dec C

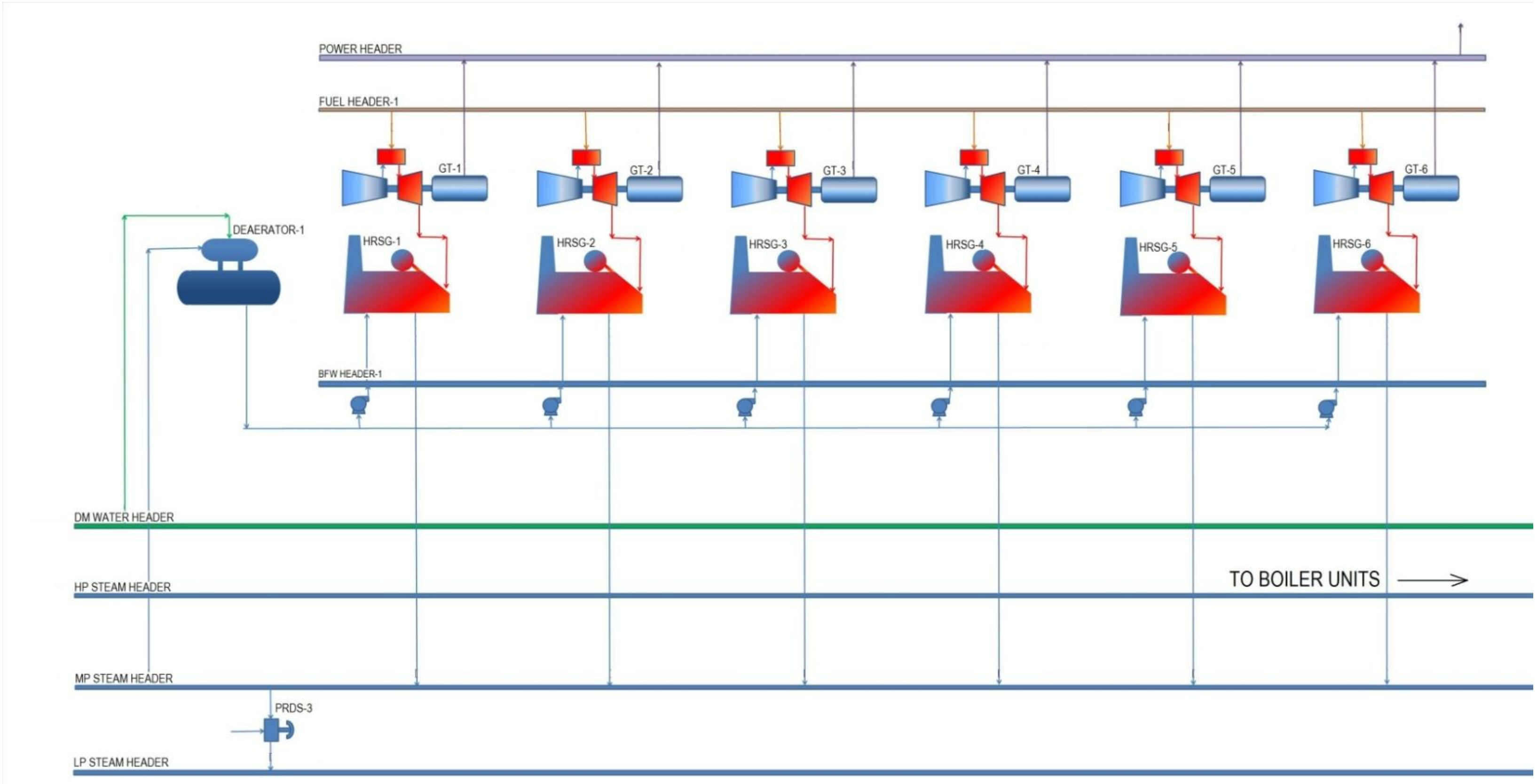


Figure B. 1 - District Heat and Power Unit (Generation Block)

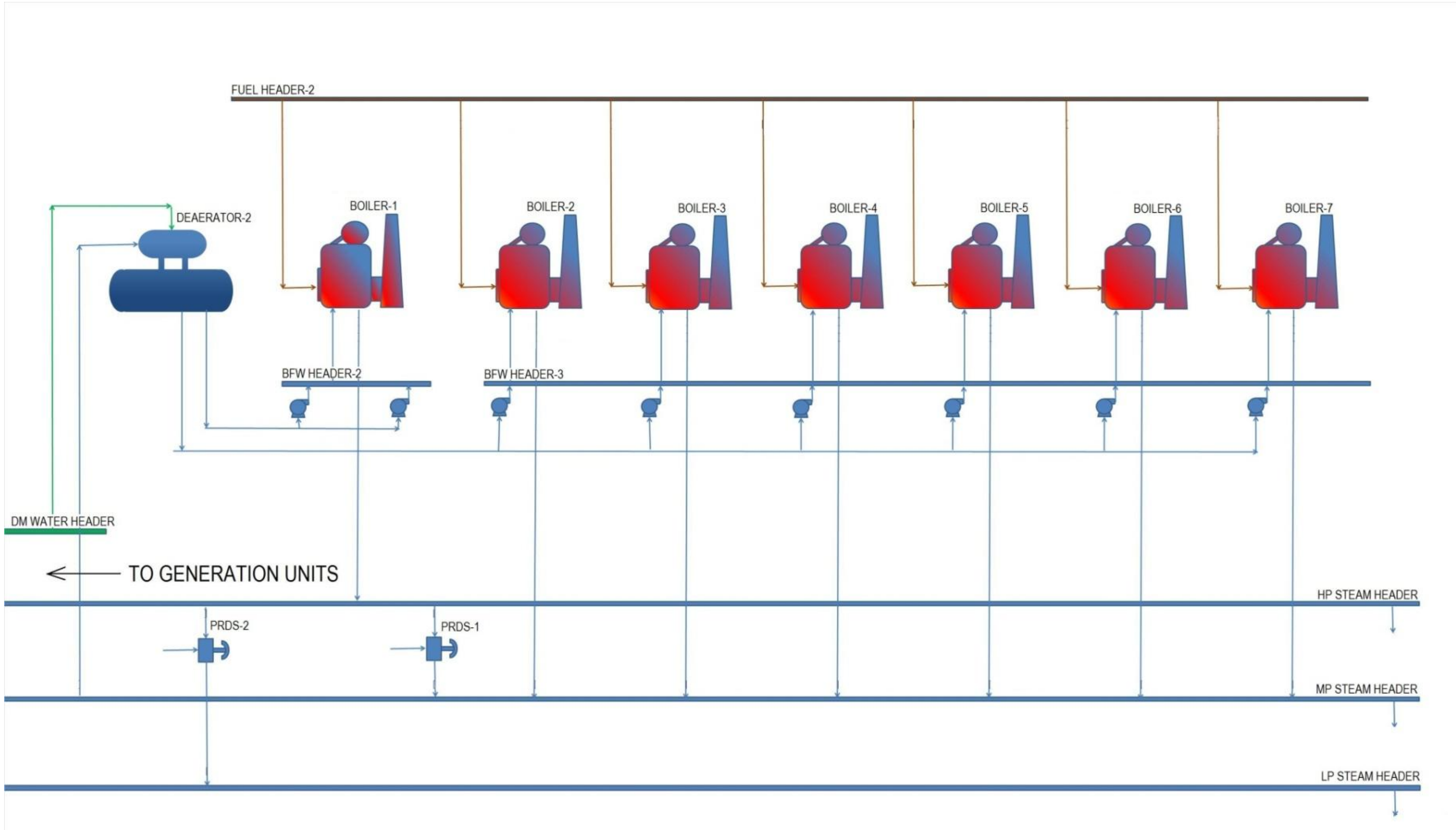


Figure B. 2 - District Heat and Power Unit (Boiler Battery)

APPENDIX C: OPTIMIZATION MODEL (GT-HRSG)

The steady state optimization model for GT-HRSG is described below. The plant data is taken from a Frame 5 Distillate fired GT and the HRSG is a twin drum HP/MP unit.

Table C.1 shows the variation of the GT fuel input with change in GT power. Regression analysis is used to fit the plant data (GT-CPP Plant 230, CPCL) into the equation of form ***"GT fuel = 0.06*GT Power + 0.52".*** Comparison between the GT fuel obtained from the plant data and the one from the derived empirical formula is shown in Fig. C.1.

Table C. 1 - GT Fuel Consumption Chart

GT Power	GT Fuel Input (Plant Data)	GT Fuel Input (Empirical Formula)	Error
(MW)	(kg/s)	(kg/s)	(%)
10	1.12	1.12	0
12	1.25	1.24	-0.81
14	1.37	1.36	-0.73
16	1.48	1.48	0
18	1.61	1.6	-0.62
20	1.71	1.72	0.58
22	1.84	1.84	0

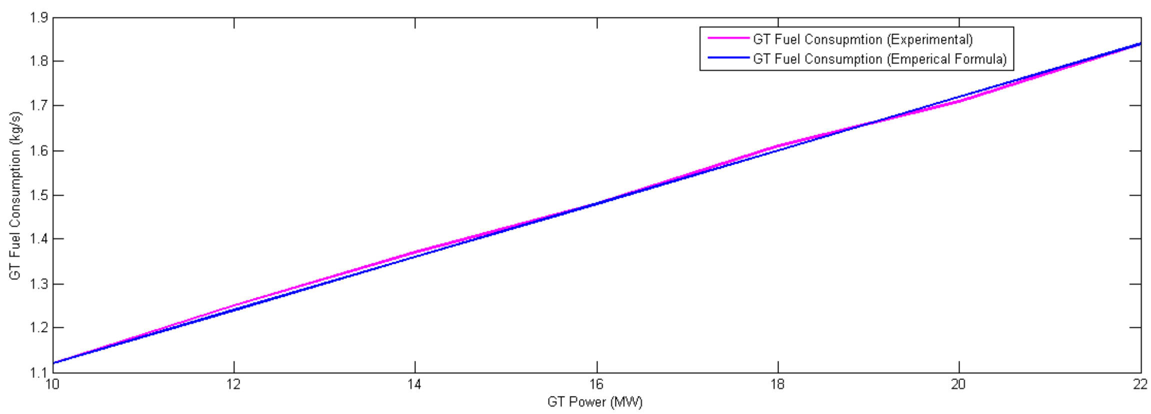


Figure C. 1 - GT Fuel (Plant Data Vs Formula)

Table C.2 shows the variation of the HRSG HP and MP steam with change in GT power. Regression analysis is used to fit the plant data (GT-CPP Plant 230, CPCL) into the equation of form ***"HRSG HP Steam = 1.5*GT Power + 9"*** and ***"HRSG MP Steam = 0.1*GT Power + 3".*** Comparison between the HRSG steam obtained from the plant data and the one from the derived empirical formula is shown in Fig. C.2.

Table C. 2 - HRSG Steam Flow Chart

GT-1 Power	HRSG HP Steam (Plant Data)	HRSG HP Steam (Empirical Formula)	Error	HRSG MP Steam (Plant Data)	HRSG MP Steam (Empirical Formula)	Error
(MW)	(kg/s)	(kg/s)	(%)	(kg/s)	(kg/s)	(%)
10	24	24	0	4	4	0
12	27.2	27	-0.7407	4.1	4.2	2.381
14	30	30	0	4.3	4.4	2.2727
16	32.4	33	1.81818	4.5	4.6	2.1739
18	36	36	0	4.6	4.8	4.1667
20	39	39	0	5	5	0
22	42.5	42	-1.1904	5.2	5.2	0

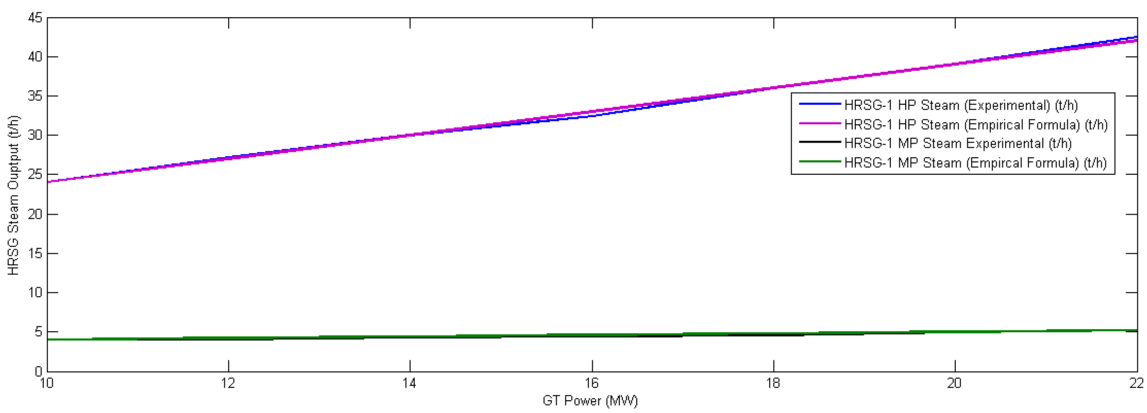


Figure C. 2 - HRSG Steam (Plant Data Vs Formula)

APPENDIX D: DYNAMIC CONTROL MODEL (GT-HRSG)

The dynamic control model for GT-HRSG is described below. The plant data is taken from a Frame 5 Distillate fired GT and the HRSG is a twin drum HP/MP unit.

Table D.1 shows the GT dynamic model parameters used in simulation. The model parameters were chosen to fit the plant dynamic data (GT-CPP Plant 230, CPCL). The comparison between the real time plant data and the simulation when the GT power is increased from 16MW to 19 MW is shown in Figure D.1.

Table D. 1 - GT Model Parameters

GT Model Parameter	Parameter Value
Setpoint Rate Limiter	0.2
Saturation lower limit	0
Saturation Higher limit	25
First Order Parameter (s,c)	10,1
Dead time	0.5
White Noise Variance (Power)	0.01

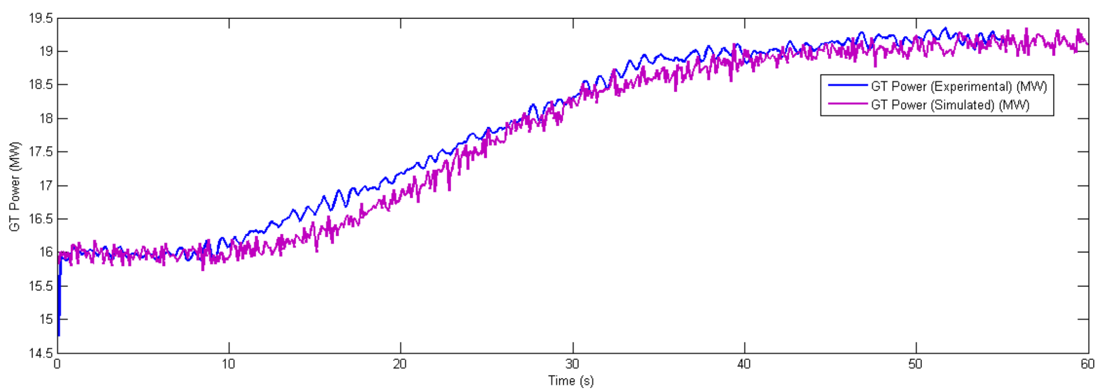


Figure D. 1 - GT Dynamic Model (Plant Data Vs Simulated)

Table D.2 shows the HRSG dynamic model parameters used in simulation. The model parameters were chosen to fit the plant dynamic data (GT-CPP Plant 230, CPCL). Comparison between the real time plant data and the simulation where the HRSG steam increases from 25 t/h to 33 t/h is shown in Figure D.2.

Table D. 2 - HRSG Model Parameters

HRSG Model Parameter s	Parameter Value
Damper angle	90
Saturation lower limit	0
Saturation Higher limit	50
Second Order Parameter (s2,s,c)	1250,50,1
Dead time	4
White Noise Variance (Steam flow)	0.05

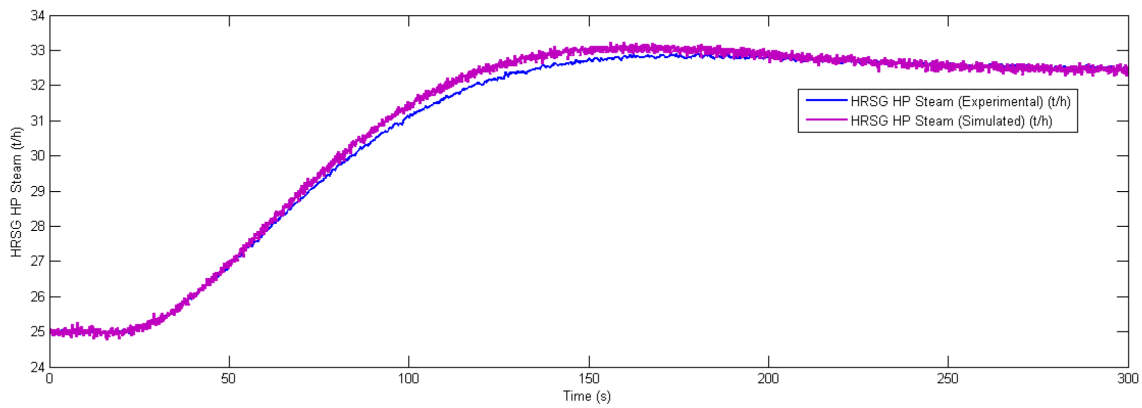


Figure D. 2 - HRSG Dynamic Model (Plant Data Vs Simulated)