AN INVESTIGATION INTO A GENERALLY APPLICABLE PLANT PERFORMANCE INDEX

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AN INVESTIGATION INTO A GENERALLY APPLICABLE PLANT PERFORMANCE INDEX

by

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SYNOPSIS

It is important to develop methods that are capable of successfully determining plant performance. The method used should be based on the ability to determine the performance of each of the various unit operations within the plant. This in turn will assist with the correct decision as to which unit in the plant should be improved first. The performance of the various units can be accumulated to give a representation of the performance of the entire plant.

A plant-wide performance monitoring method has been developed to do just this. Originally it was developed for a specific unit operation. It has now been verified that this method is applicable to different unit operations. The method employed to determine this plant-wide performance is by evaluating how close the plant is to its inherent optimum. Where applicable, this inherent optimum can also be replaced with a user specified optimum. When an optimum is specified there is a possibility of oscillations around this "optimum" and the effects of this on the performance number are eliminated to give a more general plant-wide performance number for each unit operation.

In addition to the "optimum" value selection the addition of performance weights to specific focus areas (utility usage or product quality) in the performance calculation will also improve the comparative nature of the plant-wide index for different unit operations.

The scope of this investigation is limited to the experimental test rigs that were available in the Process Control Laboratory at the University of Pretoria. The methods that were used to determine the single loop performance of each of the different control loops are:

- Minimum variance
- Generalised minimum variance
- Integral of the Absolute Error (IAE)
- Integral of the Square Error (ISE)

The single loop performance methods are required to determine how effectively the plant-wide performance index evaluates the plant, since these are existing means of determining how well a plant is operating, but these become impractical due to excessive amounts of information needing evaluation.

KEYWORDS: performance monitoring, plant, controller, oscillation detection, plant-wide performance, controller optimisation, time-based, frequency-based, minimum variance, diagnosis.

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NOMENCLATURE

a	amplitude
C_0	controller model
CPM	Control Performance Monitoring
CPA	Control Performance Assessment
e	error
E	mean of
EHPI	Extended Horizon Performance Index
f	resolution
F_c	control signal weighting
GMV	Generalised Minimum Variance
GUI	Graphical User Interface
HIS	Historical data benchmark
i	limited time period
ISE	Integral of the Square Error
IAE	Integral of the absolute of the error
ITAE	Integral of the time-weighted absolute error
LQG	Linear-Quadratic Gaussian
m	number of products
MPC	Model Predictive Control
MVI	Minimum Variance Index
n	size of data set
n_{lim}	limited number of disturbances
OPID	Optimal PI(D)
p	number of feeds
P_c	control error weighting
PI	Proportional Integral Controller
PID	Proportional Integral Derivative Controller

P_{yy}	power density
PWI	Plant-Wide Performance Index
P&ID	Piping and Instrumentation diagram
q	number of utilities
r_0	The value of r_{uy} at lag 0
r_{max}	The maximum value of r_{uy}
r_{uy}	Cross correlation function
s	sensitivity factor
S_{cap}	capable standard deviation
S_{fbc}	Fellner's formula
S_{real}	real standard deviation
SLP	single loop performance
t	time
$t_{a,b}$	time of analysis
T_{sup}	evaluation time
T_u	ultimate oscillation period
UPI	unit operation performance index
u	control action
u(t)	input signal
VI	variance index
w	plant model
w_x	weight
w_d	disturbance model
x	input
y	output
y(t)	output signal
z	number of sign changes
α	upper bound on controller output variance
ϕ_0	Optimisation for generalised minimum variance

ζ	process disturbance
λ	constant related to the evaluation time
μ	mean
$\hat{\mu}$	Gaussian mean
σ	standard deviation
$-\tau_l$	The first zero crossing for negative lags
τ_r	The first zero crossing for positive lags
ω	Frequency
ω_d	Deadtime Frequency
ω_u	ultimate frequency
ω_i	estimate of ultimate frequency
Δt	sampling period
ΔQ	Phase shift

1 INTRODUCTION

1.1 Background

It is a well known fact that control loop performance will deteriorate over time as a result of process changes and equipment deterioration (Ghraizi et.al., 2006).

It has been found that some advanced control applications, specifically Model Predictive Controllers (MPC) do not perform as well as was originally intended or designed (Gao et.al., 2003; Julien, Foley & Cluett, 2004). Also, some baselayer control systems do not perform adequately to optimally support advanced control (Desborough & Miller, 2001; Hugo, 2002).

There are many methods available for the development and implementation of control loops, but there are comparatively few methods available to determine the performance of these control loops. Many of these fall short of recognising the difference between acceptable performance and good control (Harris, Seppala, Jofriet & Surgenor, 1996).

Research relating to control loop performance monitoring and diagnosis has been an active research area (Huang, Ding & Thornhill, 2005). Extending this to plant-wide performance monitoring has now become an area of interest.

1.2   Problem Statement

Performance monitoring techniques are not generic, and reports on performance are generally given in a graphic format, which is not easily interpreted or comparable for different control loops. These graphical methods include control charts, cumulative sum charts and statistical distributions. A generic and easily understandable performance index is required for purposes of quantifying performance in a simple, yet soundly-based manner, but leaving the graphical representation accessible for a more detailed analysis of problem loops, should it be necessary.

It is the purpose of this investigation to establish a method which will be used to determine the plant performance of different unit operations in such a manner that the result for each unit operation will be comparable with that of various unit operations and should thus be as generic as possible. This performance number would ideally indicate how close the plant is performing to a desired optimum.

This technique should therefore be universally applicable and easily adaptable to different types of loops (i.e. level, flow, composition and temperature), unit operations (i.e. distillation columns, heat exchanges and reactors) and plants (i.e. petrochemical, minerals processing).

1.3 Objectives

In order to achieve a plant-wide performance index that is comparable for different unit operations it will be required to investigate the following:

- Search for and evaluate existing methods and derive a plant-wide performance monitoring method that will be applied to different unit operations which contain different types of controllers.
- Evaluate different methods for determining if oscillations are present and determining the source of such oscillations.
- Determine if there is a relationship between controller performance and plant performance. This will require the investigation of various single loop performance methodologies.

1.4 Method

In order to achieve the stated objectives different methods will have to be investigated and tested on the available equipment. To facilitate this investigation a software interface developed by a previous investigator (du Toit, 2005), will be used and expanded as required. Further evaluation of the method used during the development of this software will be applied to different available installations. Furthermore, the software will also have to be adapted to achieve the stated objectives and to incorporate other appropriate methods for single loop evaluation and oscillation detection, as well as to expand upon the existing plant-wide performance monitoring method.

This software interface will be used as a means to evaluate the data obtained from different installations. The original interface was developed for the evaluation of a distillation column. This unit will be re-evaluated and compared to the plant-wide performance index obtained for the other available installations, which include a smaller distillation column, a highly interactive level and flow rig and a temperature control rig which has variable deadtime. For each of these units different controller settings will be used as a means of worsening or improving the performance of that specific unit. The results obtained from each unit can then be compared with that of the other units evaluated.

2 LITERATURE SURVEY AND THEORETICAL BACKGROUND

Each piece of equipment has an inherent maximum throughput and operating efficiency. This inherent nature cannot be changed by the control system. The purpose of the control system is to drive the plant towards this optimum while maintaining operational stability (Blevins, McMillan, Wojsznis and Brown, 2003).

The deviation of the plant from this optimum is used to calculate the performance of that plant. Performance is therefore an indication of how close the plant is operating to its inherent maximum.

When applying controllers, a control engineer is faced with a trade-off between performance and stability, where performance typically requires fast responses and aggressive control action and stability requires slower responses and conservative control actions. Advanced control design has the main objective of maximising performance for the process while still ensuring stability (Romagnoli & Palazoglu, 2005: 10).

After implementation of a control loop it is therefore required to determine the performance of that loop to see if it is giving the performance it was designed for (Kamrunnahar, Fisher & Huang, 2004), and then implement whatever corrective action is required if the loop is not performing as desired.

2.1   Reasons for Poor Performance

Previous studies show that between 66% and 80% of controllers do not perform optimally (Desborough & Miller, 2001; Hugo, 2002) and that large numbers of advanced controllers like Model Predictive Controllers (MPC) do not perform as well as is desired (Gao et.al., 2003; Julien, Foley & Cluett, 2004).

One of the main reasons for this low level of performance is the large number of control loops in a processing plant, and the fact that larger sections are being placed under the responsibility of fewer supervisors (Hägglund, 1999). This means that no one has the time to investigate these poor performing loops and implement solutions to improve the performance. The performance of control loops also degrades over time and unless continual monitoring and improvement is implemented it is not possible to achieve optimally performing controllers.

In many cases these poor performing loops are not identified as such by the responsible personnel and no further investigation into the underlying cause of the poor performance can be initiated until these loops have been identified as poor performers (Harris, Seppala & Desborough, 1999). The main reason for this is that just viewing the raw data is often not sufficient to differentiate between normal operating conditions and abnormal situations (Rengaswamy & Venkatasubramanian, 1995). These data trends often result in complicated response patterns as they are influenced by noise and disturbances (Qiang & Shaoyuan, 2006).

Industrial interest resulted in research into performance monitoring tools that can assist operators to identify problematic loops (Hägglund, 2005). These methods fall under the banner of any of the following (Jelali, 2005):

- Control performance monitoring (CPM)
- Control performance assessment (CPA)
- Control loop management
- Loop auditing

Possible sources of poor controller performance within industry include the following (Jelali, 2005):

- Inadequate controller tuning
- Operating conditions that change over time
- Lack of maintenance resulting in equipment malfunction
- Poor plant and controller design
- Poor feedforward compensation, where applicable
- Oscillatory manipulated variables
- Abnormal situations
- Feedstock variations
- Process nonlinearity
- Seasonal influences (Schäfer & Cinar, 2004)
- Control loops in manual mode (Paulonis & Cox, 2003).

When using a model-based controller there are still more causes of poor performance that need to be considered, all of which are due to uncertainty in the models (Patwardhan & Shah, 2002), namely:

- Model uncertainty

- Disturbance uncertainty

- Time delay uncertainty

- Noise model uncertainty

Uncertainty is also the fundamental cause of the need for feedback when implementing model based controllers. Without feedback it would not be possible to achieve a good performance unless the model was precise, which is hardly ever the case.

Origins of uncertainty include (Skogestad & Postlethwaite, 2004: 255):

1. approximations or errors in model parameters
2. parameters that vary due to nonlinearities or changes in operating conditions
3. measurement device imperfections (lack of calibration or nonlinearities)
4. using simpler, lower order models (i.e. creating model uncertainty)

2.2   Benefits of Monitoring Performance

Since we know that there are factors which can degrade the performance of a controller on a continual basis it is required to continually observe the actual performance of the controller. Thus when any changes in performance are detected it will be possible to investigate possible solutions in a timely fashion.

The control system can drive the plant closer to the inherent optimum of the plant resulting in (Blevins et.al 2003):

- better product quality

- increased throughput

- reduced waste products

- reduced energy usage

A more efficient plant maintenance plan can be incorporated allowing (Thornhill, Oettinger & Fedenczuk 1999):

- better scheduled maintenance

- reduced workloads due to reduced troubleshooting of problematic loops

- prevent operator information overload

A plant that operates under an effective performance monitoring structure leads to a more profitable, more efficient and safer plant (Rengaswamy et.al., 2001).

2.3   Controller Performance Monitoring Methods

The performance monitoring criteria can be split into steady-state performance criteria and dynamic response performance criteria. These performance criteria developed for single loop controllers are (Marlin, 2000: 219-220):

- Error at steady state

- Manipulated Variable Overshoot (C/D)

- Controlled Variable Overshoot (A/E)

- Rise time (T_r)

- Settling time (time to reach ±5% of the desired value)

- Decay ratio (B/A)

- Period of oscillation (P)

These criteria are illustrated in **Figure 2-1**.
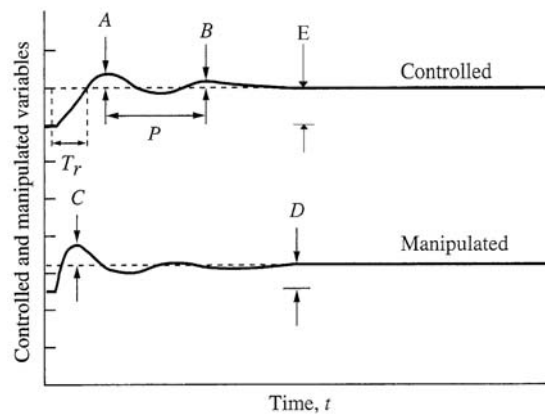
Figure 2-1: Setpoint change response for a typical feedback controlled system (Marlin, 2000: 219)

A further criterion used to evaluate single loop controllers in the Laplace domain is the observation of root location in the s plane (Luyben, 1990: 351) Once a base-case (a period of operation deemed acceptable by the control engineer) has been established, these performance measures could be used for early performance degradation detection (Romagnoli & Palazoglu, 2005: 475). These methods are however more applicable to a system undergoing setpoint changes (Jämsä-Jounela, Poikonen, Vatanski & Rantala, 2003).

The more traditional methods used for early performance degradation detection, depending on the desired outcome, are the minimisation of one of the following (Marlin, 2000: 219):

- Integral of the square error (ISE)
- Integral of the absolute value of the error (IAE)
- Integral of the time-weighted absolute error (ITAE)

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Each of these has a specific application namely (Marlin, 2000: 219):

- ISE, calculated using equation **(2.01),** is used when large errors are being investigated as the cause of performance degradation.
- The IAE, calculated using equation **(2.02),** is used for systems where small errors occur as the source of performance degradation.
- When errors persist for long periods of time the ITAE, calculated using equation **(2.03),** is used as performance measure.

$$ISE = \int_0^\infty \left[SP(t) - CV(t) \right]^2 dt \qquad \textbf{(2.01)}$$

$$IAE = \int_0^\infty \left| SP(t) - CV(t) \right| dt \qquad \textbf{(2.02)}$$

$$ITAE = \int_0^\infty t \left| SP(t) - CV(t) \right| dt \qquad \textbf{(2.03)}$$

Also included in these traditional methods but becoming more complex and requiring plant models to perform any analysis are the items discussed in the following sections.

2.3.1 Input-Output Controllability

The controllability objective of the plant is to achieve acceptable control performance in spite of disturbances. This analysis is independent of the controller; therefore it is a property of the plant alone. If a plant is deemed uncontrollable it would be necessary to change the design of the plant itself. These changes could include any of the following (Skogestad & Postlethwaite, 2004: 160):

- changing the type, size, etc, of the apparatus itself
- relocating sensors and actuators
- adding new equipment to dampen disturbances
- adding extra sensors
- adding extra actuators
- changing the control objective
- changing the configuration of the lower layers of control already in place

2.3.2 H$_\infty$ norm

The H$_\infty$ norm is determined by finding the peak value of a stable scalar transfer function f(s), as a function of frequency, ω, using equation **(2.04)** (Skogestad & Postlethwaite, 2004: 55).

$$\|f(s)\|_\infty = \max_\omega |f(j\omega)| \tag{2.04}$$

This method is used during the design phase. The lower these peaks are the more stable the plant will be and therefore the better the controller performance is when implemented. The H$_\infty$ norm is therefore used to select the most appropriate combination of transfer functions for the control system that will result in an operable plant but at the same time have the lowest peak which will result in a more stable plant.

The H$_\infty$ norm can also be used after commissioning of the plant as it can be used to determine the most appropriate changes that can be made to the control system that will result in a more controllable and better performing plant.

2.3.3 Robustness

Robust performance is observed in the Nyquist plot of the system. Ensuring that the Nyquist plot of the system does not encircle the -1 point on the real axis will ensure stability while maintaining a suitable distance from this point will ensure robust performance. These methods should ideally be applied during the design phase of the unit to achieve a plant that will be controllable and can achieve acceptable performance (Skogestad & Postlethwaite, 2004: 159-252).

All the methods discussed thus far are relatively simple and have possibly already been used during the development of the control systems of most processing plants. These methods do however still serve as a platform for controller performance. These methods can also be used to identify the causes of or reasons for the poor performance on a plant.

2.4   Time-based Controller Performance Monitoring Methods

The basic procedure to be followed during control performance monitoring is shown in **Figure 2-2**. This includes analysing the current control loop and comparing that to a benchmark. When the current loop is not performing as desired then a diagnosis of the underlying cause of the deviation is done. The results of this comparison are then used to determine the necessary action.

```
                    ┌────────────────────────────┐
                    │  Computation of current    │◄───────┐
          ┌────────►│  controller performance    │        │
          │         └────────────────────────────┘        │
          │                      ▲ u                       │
          │         ┌────────────────────────────┐         │
   r      │         │  Control loop              │    y    │
  ───────►│         │  to be assessed            │────────►│───►
          │         └────────────────────────────┘         │
          │                      ▼ u                        │
          │         ┌────────────────────────────┐         │
          └────────►│  Model/parameter           │◄────────┘
                    │  identification            │
                    └────────────────────────────┘
                                 │
   Constraints       ┌────────────────────────────┐
  ──────────────────►│  Design of ideal/optimal   │
                     │  controller (Benchmark)    │
                     └────────────────────────────┘
                                 │
                     ┌───────────────────┐      ┌──────────────────────┐
                     │  Comparison/      │      │ * Control Retuning   │
  ──────────────────►│  Diagnosis/       │─────►│ * Control Redesign   │
                     │  Decision         │      │ * Plant Redesign     │
                     └───────────────────┘      └──────────────────────┘
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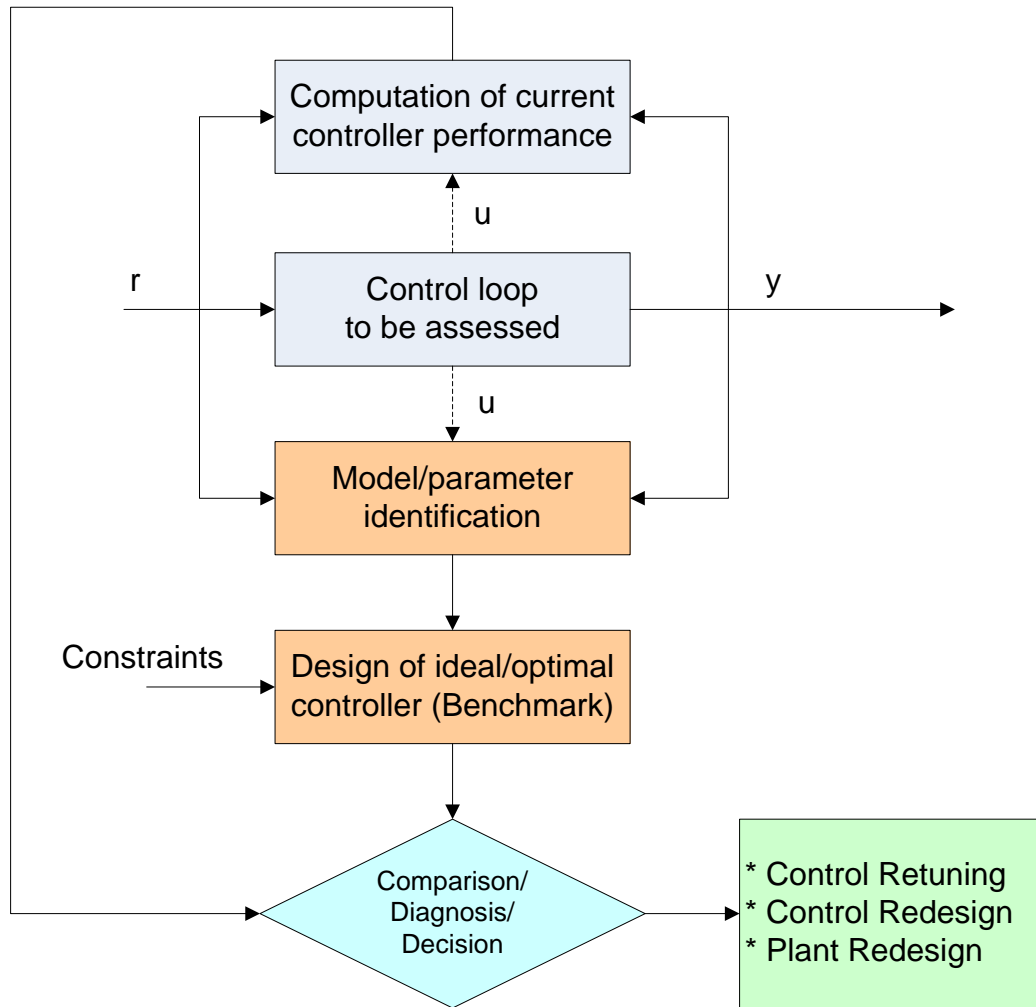
Figure 2-2: Procedure for controller performance monitoring, (Jelali, 2005)

(Jelali, 2005) gives an extensive summary of methods currently availably for performance monitoring of control loops. What follows is a short description of each method.

2.4.1 Minimum Variance Index (MVI)

The benchmark for a minimum variance controller is estimated using routine operating data. The method is non-intrusive which is why this method is suitable for industry (Huang, Ding & Thornhill, 2006). This method was first introduced by (Harris, 1989). If this method is used to optimise the controller, it effectively results in the placement of all the closed-loop poles in the origin (Horch & Isaksson, 1999). This controller is not feasible since it will be very slow, therefore having the lowest possible variance. It is however not necessary to design the controller based on this method.

The minimum variance index (MVI) can however be a useful method to analyse the performance of a loop since it is merely a comparison of the variance of the controller under investigation with that of the estimated minimum variance controller (MVC). The practical application of this method follows (Clegg, Xia & Uduehi, 2006):

When applying the Fourier transform to the normal output of a control loop a power spectral density plot is achieved. In this plot a point is reached where the frequency of changes of the control loop exceeds that of the deadtime frequency (ω_d), which is the maximum frequency of response achievable by the controller. This is illustrated in **Figure 2-3**.
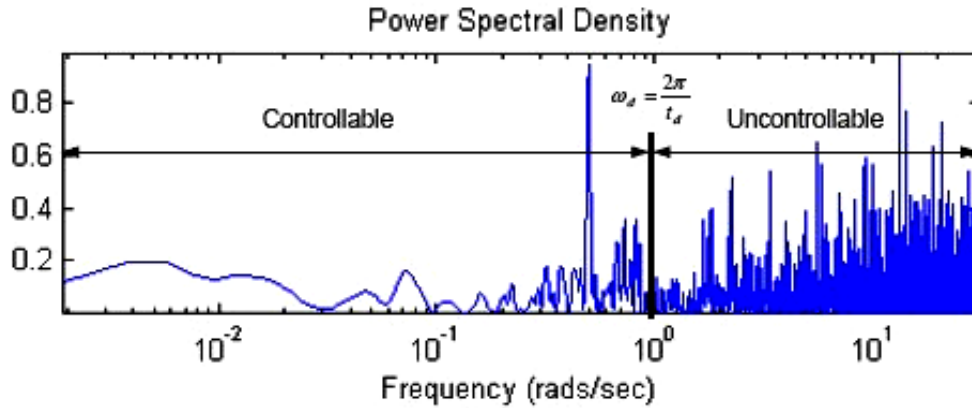
Figure 2-3: Power spectral density to illustrate the controllable and uncontrollable regions (Clegg et.al., 2006)

Beyond this point (ω_d), changes cannot be compensated for by the controller. This is known as the uncontrollable variance, since the controller can only respond to a disturbance within a time that exceeds the time interval between the application of the disturbance and the measurement of the effect of that disturbance.

Making use of an increment of 1 (therefore comparing the current and previous data point), to simplify calculations, an estimate of the minimum variance can be obtained using equation **(2.05).**

$$S_{cap} = \sqrt{\frac{\sum_{i=2}^{n}(x_i - x_{i-1})^2}{2(n-1)}}$$

(2.05)

S_{cap} is defined as the capable standard deviation and n is the size of the data set. S_{cap} is an indication of the variance in the input, x, which cannot be compensated for by the controller.

For comparison to this capable standard deviation, the real standard deviation of the process can be calculated using equation **(2.06)**, with respect to the output, y.

$$S_{real} = \sqrt{\frac{\sum_{i=1}^{n}(y_i - \bar{y})^2}{n-1}}$$

(2.06)

The real standard deviation and the capable standard deviation can now be compared using Fellner's formula as shown in equation **(2.07)**. Since the square root of variance is defined as standard deviation.

$$S_{fbc} = S_{cap}\sqrt{2 - \left[\frac{S_{cap}}{S_{real}}\right]^2}$$

(2.07)

S_{fbc}, is an indication of the standard deviation from minimum variance control.

The percentage of the loop's potential that can be achieved can be determined using equation **(2.08)**.

$$VI = 100\frac{S_{fbc} + s}{S_{real} + s}$$

(2.08)

Here s is a sensitivity factor usually in the order of 0.1. The closer the variance index, VI, is to 100% the closer the loop is to a minimum variance controller, which implies that no other controller can result in a lower process variance (Huang & Shaw, 1999: 99).

The minimum variance method derived by (Harris, 1989) is a frequency based method; the procedure discussed above is a time-based determination of the variance of a control loop, which results in a percentage of the minimum achievable variance for the control loop in question. It is this percentage that will be used as the minimum variance index (MVI).

2.4.2 Historical benchmarking (HIS)

HIS uses historical data as a benchmark; the data is obtained when the plant is defined as "doing well". The current plant operation is then compared against this control/maintenance engineer-determined optimum, and the plant must have already achieved this "optimal" state. This optimum could be determined by simply examining the raw data. This optimum state could also be determined by observing the ISE, IAE or ITAE of the process over a desired interval. It is also possible to specify the desired closed loop response and then comparing the plant to this desired response.

2.4.3 Generalised Minimum Variance (GMV)

This method is an extension of the MVI, the difference being that a control action penalty is applied. This is achieved by using the same method as that for the minimum variance index, but the output/control error as used in the calculation is replaced by a new signal, which is the weighted sum of the output/control error and the control effort (manipulated variable movement) as seen in **Figure 2-4** (Jelali, 2005).

Since the MVI usually results in an aggressively tuned controller, it is required to penalise the movement of the manipulated variable, and any unrealistic control errors. This will then result in a more practical performance index (Grimble, 2002). A univariate control feedback controller with the GMV performance assessment method in place is illustrated in **Figure 2-4**.

In **Figure 2-4** P_c is the control error weighting and F_c is the control signal weighting. These weights are used to lessen the aggressiveness of the controller by increasing the weighting F_c. The error (e) and control action (u) are used in conjunction with the weighted values, F_c and P_c, to determine the controller performance. The process disturbance is defined as ζ, C_0 is the model for the controller while, w is the plant model and w_d is the disturbance model.
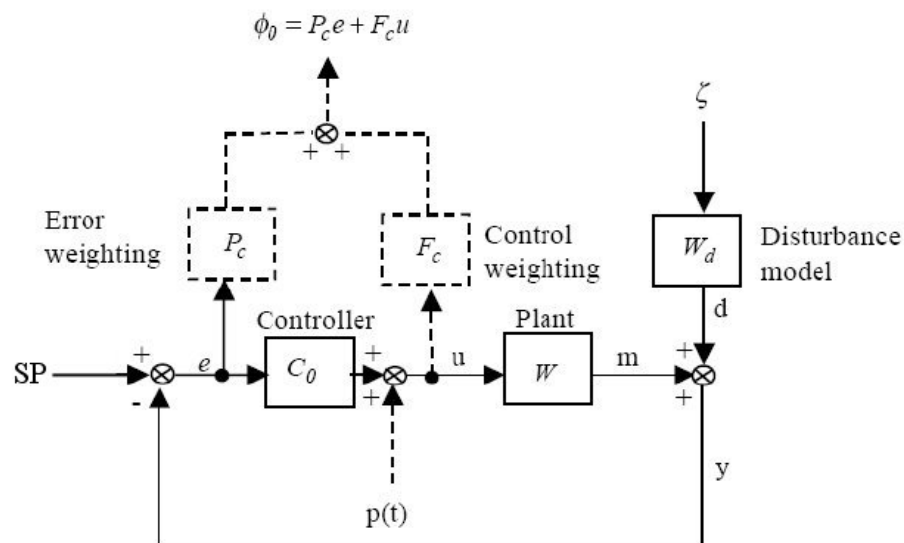

Figure 2-4: Method for generalised minimum variance (Grimble, 2002)

ϕ_0 is then used as the variable to be minimised in order to achieve an optimal controller, or in this case as the value that can be used as the controllers signal and by applying the MVI to determine the performance of the controller.

2.4.4 Extended Horizon Performance Index (EHPI)

EHPI is used when time delays are unknown, for systems where it may be difficult or expensive to determine the deadtime of the process. The minimum variance index (MVI) is determined for each time delay over a range of time delays. This range is decided upon, based on the expected value of the actual time delay of the process. The calculated MVI values are then plotted over the range of time delays chosen. The region on the plot where the MVI does not vary rapidly is considered a good choice for the prediction horizon as this is an estimate of the time delay of the process (Ingimundarson & Hägglund, 2004). This method saves time in avoiding the time consuming determination of time delays by performing step tests on the plant.

2.4.5 Linear-Quadratic Gaussian (LQG)

This method also has a penalisation of the manipulated variable inherently built into the method. The benchmark for LQG uses a trade-off curve which plots variance in controlled variables vs. variance in manipulated variables, as illustrated in **Figure 2-5**. This curve is generated by varying λ, which is a value used to span the region, in equation **(2.09)**, which is the LQG objective function with E being the mean of the process input, u, related to the CV and the process output, y, related to the MV.

$$J(\lambda) = E\left[(CV - SP)_t^2\right] + \lambda E\left[(MV - MV_{best})_t^2\right] \qquad \textbf{(2.09)}$$
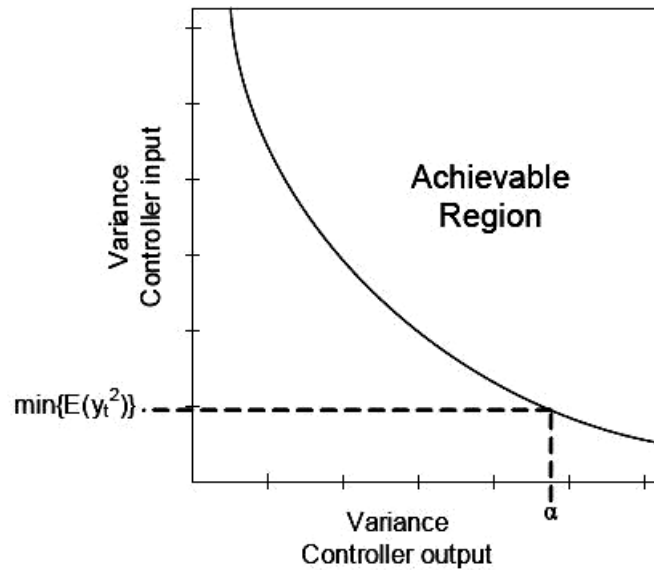
Figure 2-5: Trade-off curve for LQG (Huang & Shah, 1999: 154)

In **Figure 2-5** the achievable performance is limited by the curve. By placing an upper bound, α, on the variance of the controller output, the curve then gives a realistic minimum variance for the process. This value, α, is chosen by the control engineer and is based on the personal opinion as to how much controller variance will be acceptable for a specific controller.

2.4.6 Optimal PI(D) Control (OPID)

OPID compares the response of the current controller to the best possible achievable PI(D) controller response for that loop. The optimal controller settings that will result in the best possible achievable PI(D) response can be determined off-line using models of both the process and the disturbance and evaluating the response with any acceptable method namely, ISE, IAE or overshoot and settling time, whichever the person responsible for determining these settings chooses based on their applicability. If these models are however not accurate, the method would not be very successful.

2.4.7 Summary of Performance Monitoring Methods

A summary of the requirements and computational burden for the determination of plant performance for the various methods discussed is given in **Table 2-1**.

Table 2-1: Summary of requirements for various methods, (Jelali, 2005)

Method	HIS	EHPI	MV	GMV	LQG	OPID
Parameters/data required						
Time delay			√	√	√	√
Control error		√	√	√	√	√
Control input	(√)			√		
Process/disturbance model					√	√
Controller structure						√
Benefits						
Control benchmark	√	√	√	√	√	√
Limitation of actuator energy				√	√	(√)
Reflection of controller structure	(√)	(√)				√
Provision of optimal controller setting						√
Required computational burden (Low: * - High: ***)**	*	**	**	***	*****	****

2.5 Oscillation Detection Methods

Surveys have shown that almost 30% of control loops are oscillating (Singhal & Salsbury, 2005). Oscillation detection complements the evaluation of control loop performance (Thornhill & Hägglund, 1997). The detection and diagnosis of undesired oscillations within a chemical process is important since profit is impacted by process variability (Thornhill, Huang & Zhang, 2003). This is especially true when recycle streams are present which propagate the oscillation throughout the plant (Thornhill, Cox & Paulonis, 2003).
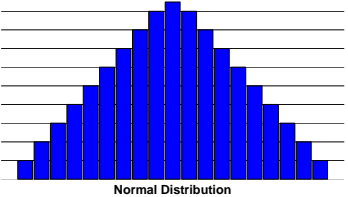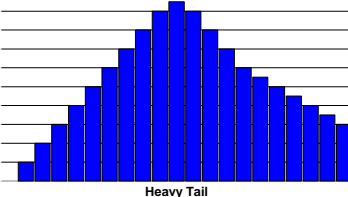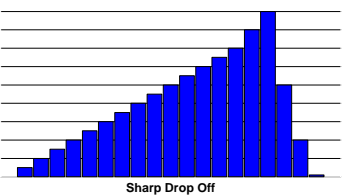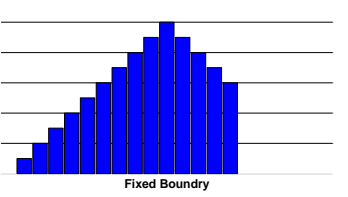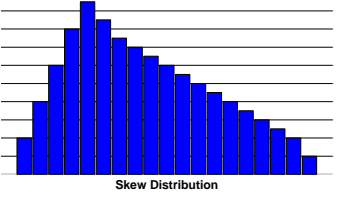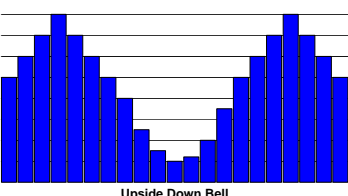
The origins of oscillations in a plant include internal or external disturbances, aggressive control action or non-linearity (stiction, dead zone and hysteresis) (Jelali, 2005). If left unchecked, oscillatory behaviour can lead to degradation and eventually failure of control valves. These oscillations therefore lead to increased maintenance costs and also increased energy costs (Salsbury, 1999).

Several different techniques can be used to identify the presence and origin of the oscillatory behaviour, these will be discussed in the following sections.

2.5.1 Histograms

A histogram is generated by subdividing the data, gathered from an operating unit, into error bins. Here the deviation of the controlled variable from setpoint is defined as the error. These bins are then plotted and the frequency of a certain error can be seen. This data can be used to visually determine if the system is oscillatory or not (Shunta, 1995). For investigation of the histograms some of the following non-normal distributions could occur as seen in **Table 2-2.**

Table 2-2: Histograms for determination of the source of oscillations

Origin of Non-Gaussian distribution	Appearance	Solution	Example
Normal	Gaussian	N/A	 Normal Distribution
Outliers	Heavier Tail	Incidental outliers can be filtered out Real outliers need further investigation	 Heavy Tail
Measuring Elements	Sharp drop off on one side	Recalibration of measuring elements	 Sharp Drop Off
Variable Characteristics	Fixed boundary on one side	Physical constraint on system	 Fixed Boundry
Nonlinearity	Skew distribution	Invert non-linear effect through measuring elements	 Skew Distribution
Final Control Elements	Upside down bell or two maxima	Valve Maintenance or Re-tuning	 Upside Down Bell

It is generally accepted that if the histogram has two maxima the oscillation is caused by stiction but it is also possible to get a histogram with two maxima for a system that is tightly tuned (Jones, 2005). With no visual difference between the two, other methods will be required to determine the source of such an oscillation.

2.5.2 Normality Index

The process of determining the cause of the oscillation can be automated by normalising the deviation from setpoint and calculating a normality index. This is done using the following algorithm as summarised by (Jones, 2005).

1. Calculate the mean (μ) and the standard deviation (σ) from the data.
2. Calculate the distribution of the error, and save the number of occurrences of each error in an error range (this data is used to generate the histogram of the data).
3. Calculate the normal distribution f(x) characterised by the same information determined for the system, i.e. μ and σ. Minimise the sum of the square of the errors between function f(x) and the histogram, using a single multiplier K, which is a result of the difference between the actual and the normal distribution.
4. Based on the difference between the Gaussian curve (normal distribution) and the histogram, use one of the following measures to determine the normality of the data:

   - Calculate the mean square error (MSE) between the cumulative sum of the histogram and the Gaussian curve. If MSE > 0.01, flag as oscillatory data.

- If $\dfrac{\left| \mu_{data} - \hat{\mu} \right|}{\sigma_{data}} > 10$ then flag data as oscillatory. This effectively calculates the accuracy of fit between the histogram and the Gaussian curve as $\hat{\mu}$ is the mean of the fitted Gaussian curve.

This method accurately determines if the system is oscillatory or not, but to determine the cause of the oscillation, requires further analysis.

2.5.3 Frequency of Control Error Sign Changes

A robust and easy way of determining the oscillatory nature of a normal regulatory controller is suggested by (Hägglund, 1995) as the analysis of the frequency of the control error over a sign change.

Firstly it is required to determine whether a load disturbance or setpoint change has taken place. This is done by determining the integral of the absolute error (IAE) using equation **(2.10)**, which is similar to equation **(2.02)** except that it is over a limited time period, i.

$$IAE = \int_{t_{i=1}}^{t_i} |e(t)| \, dt \qquad\qquad \textbf{(2.10)}$$

This IAE is compared to a limit value to determine whether a load disturbance has occurred. This limit value is given by equation **(2.11)**.

$$IAE_{\lim} \leq \frac{2a}{\omega_u} \qquad\qquad \textbf{(2.11)}$$

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When the ultimate frequency (ω_u) is not known it can be estimated using the integral time constant of the controller giving $w_i = 2\pi / t_i$. An amplitude (a) of 1% was suggested by (Hägglund, 1995), which gives an indication of the acceptable amount of noise present in the data before it will be detected as a load disturbance.

Now it is required to implement this load disturbance detection into an oscillation detection method. By defining the evaluation time (T_{sup}) with a limited number of disturbances (n_{lim}) which can be detected before an oscillation will be deemed present, this has been selected as $n_{lim}=10$, this value can be altered by the control engineer, but a value of 10 has become the norm. This is determined using equation **(2.12).**

$$T_{sup} \geq n_{lim} \frac{T_u}{2} \tag{2.12}$$

Here T_u is the ultimate oscillation period calculated as ($T_u = \frac{1}{w_u}$).

Equation **(2.13)** is used to determine the number of sign changes (z) in the error.

$$z = \lambda z + load \tag{2.13}$$

Here λ is a constant related to the evaluation time as given in equation **(2.14)** and load is a value determined using the logic of *if IAE < IAE$_{lim}$ then load=1 else load=0.*

$$\lambda = 1 - \frac{\Delta t}{T_{sup}} \tag{2.14}$$

Here Δt is the sampling period of the signal.

An oscillation is present if equation **(2.15)** is true, which occurs when the number of sign changes (z) exceeds the acceptable number of sign changes (n_{lim}) in the time being analysed.

$$z > n_{lim} \tag{2.15}$$

This method merely identifies the presence of an oscillation and is unable to determine the source of the oscillation.

The overall algorithm for oscillation detection is given by **Figure 2-6**.
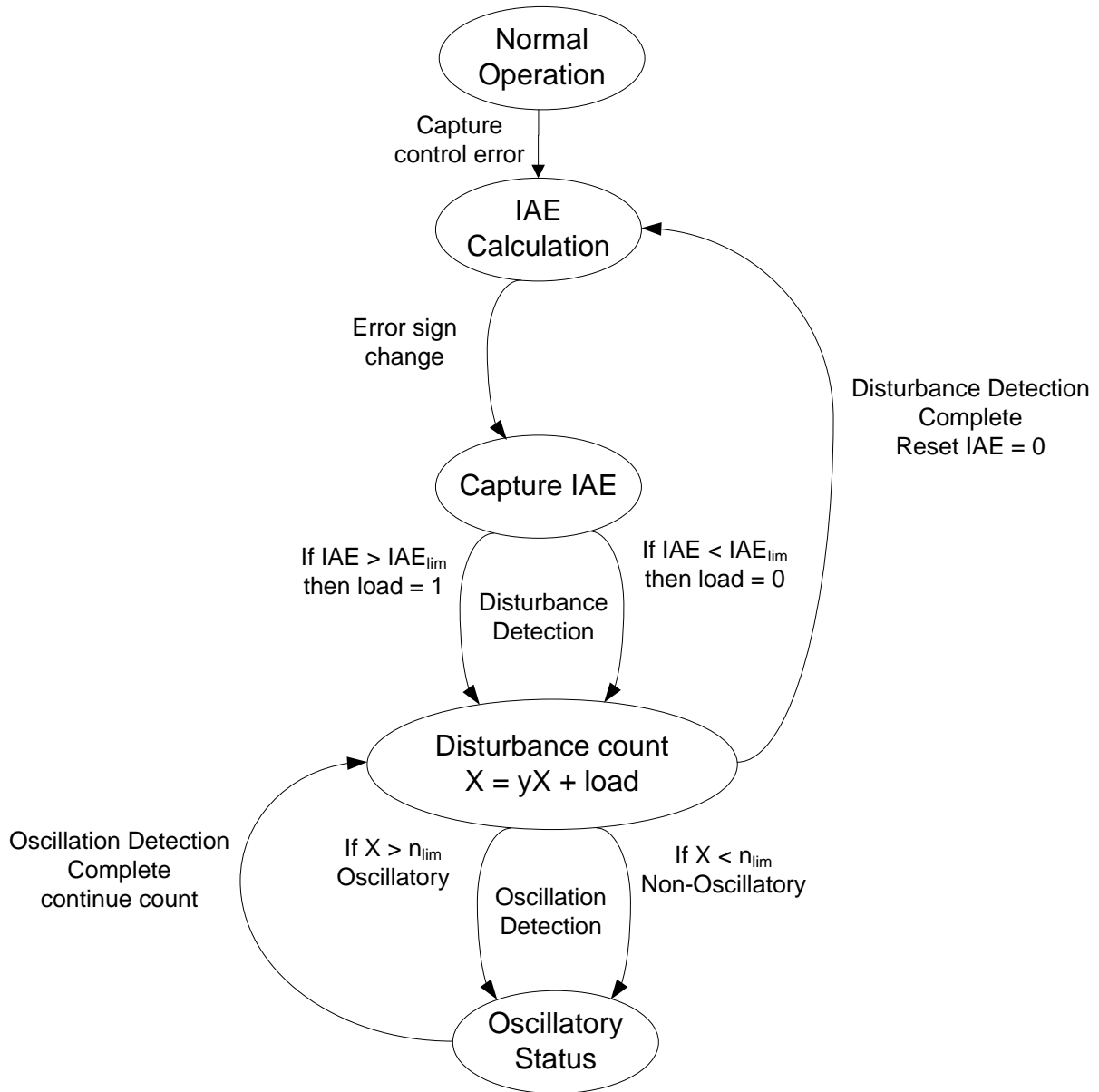
Figure 2-6: Oscillation detection algorithm (Hägglund, 1995)

2.5.4 Cross Correlation

To determine the source of the oscillation the cross correlation method can be used. The input signal, u(t), and the output signal, y(t) are normalised and then evaluated. Any of the following is possible results as shown in **Table 2-3** (Jones, 2005) are possible, the shift between u(t) and y(t) is the phase difference between the input data and the output data.

Table 2-3: Cross correlation evaluation

Shift between u(t) and y(t)	Source of oscillation
$\pi/2$	Valve Problems
π	Disturbance based

The cross correlation between the input signal, u(t), and the output signal, y(t), can be determined using equation **(2.16)** (Horch, 1999).

$$r_{uy}(\tau) = \lim_{T \to \infty} \frac{1}{2T} \int_{-T}^{T} u(t+\tau)y(t)dt \qquad (2.16)$$

Where r_{uy} – Cross correlation function

Equation **(2.16)** is used to determine the elements of equation **(2.17)**.

$$\Delta Q = \frac{|r_0 - r_{max}|}{|r_0 + r_{max}|} \qquad (2.17)$$

Where ΔQ – Phase shift (Radians)

r_0 – The value of $r_{uy}(\tau)$ at lag 0

r_{max} – The maximum value of $r_{uy}(\tau)$ in $[-\tau_l, \tau_r]$

$-\tau_l$ – The first zero crossing for negative lags

τ_r – The first zero crossing for positive lags

This value ΔQ is then used in a decision making process as shown in **Figure 2-7**. It is also clear from the figure that a safety band of $\pi/6$ is used in the decision making process.
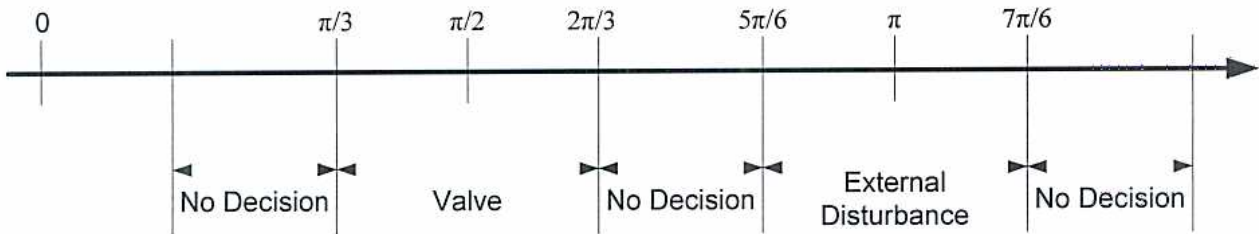
Figure 2-7: Cross correlation decision making (Jones, 2005)

This method gives a good indication as to whether the oscillation is caused by disturbances to the system or by valve problems such as stiction or tuning, but is unable to indicate whether the valve oscillation is caused by stiction or by tuning.

2.5.5 Input-Output Pattern Recognition

If the input variable is plotted against the output variable and anything other than a linear relationship is seen, then there is a possibility of an oscillation within the system. **Figure 2-8a** shows a typical industrial linear response and **Figure 2-8b** shows an example of an abnormal response that can be expected. The sharp edges denote the presence of oscillations due to stiction.

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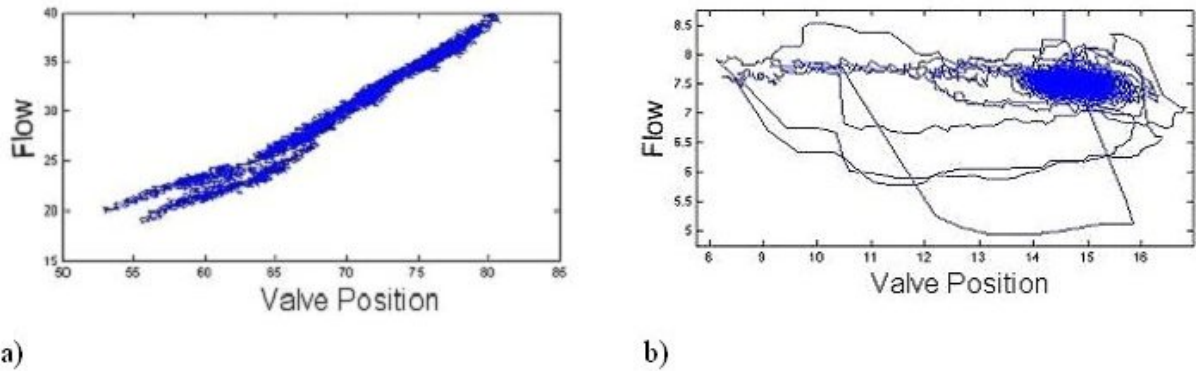
Figure 2-8: Possible patterns achievable (Jones, 2005)

2.5.6 Power Spectrum

The time domain has its benefits with respect to the analysis of controller performance, but if converted to the frequency domain, data can be interpreted with different tools. Frequency domain analysis can reveal periodicities and nonlinearities. (Choudhury, Shah & Thornhill, 2004)

The power spectrum is however only sufficient at describing linear processes (Choudhury, Shah & Thornhill, 2004). The power spectrum can be determined by taking the Fourier transform of the data using equation **(2.18)** (Jones, 2005).

$$X(f) = \sum_{n=1}^{N} x(n)\omega_N^{(f-1)(n-1)} \qquad \textbf{(2.18)}$$

Where $\qquad \omega_N = e^{\frac{-2\pi i}{N}}$ which is the N^{th} root.

$\qquad$ f = The resolution of the plot (a range of frequencies with a specific interval)

A power spectral density plot can then be obtained by plotting the power density P_{yy} obtained in equation **(2.19)** against frequency. Here the oscillation dominant period can be seen as a peak in the figure that results.

$$P_{yy} = \frac{X \times conjugate(X)}{f}$$

(2.19)

In equation **(2.19)** f, the resolution, is the same as that used for the Fourier transform in equation **(2.18)**.

This method should theoretically be able to identify the frequency of the oscillation which can then be used to identify the cause of the oscillation, this is achieved by observing the plot obtained using the power spectral density, and the peak value would indicate the frequency of oscillation present at that value, an example of this is shown in **Figure 2-9**. There are however cases where the power spectral density plot does not accurately indicate the frequency of the oscillation. This occurs particularly when both oscillations due to stiction and oscillations due to the over tuning of the control loop are present, but can still be a good indication as to the presence of possible oscillations. This is shown in **Figure 2-10**.
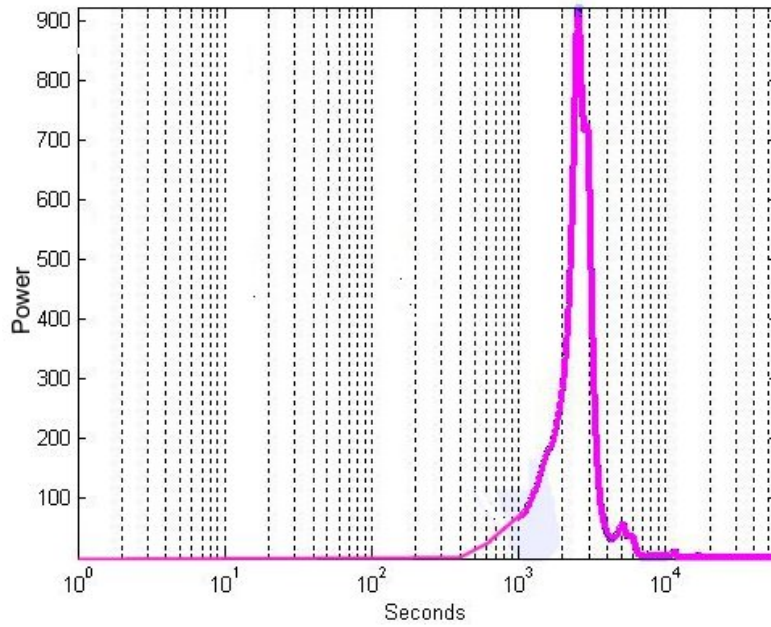
Figure 2-9: Power spectral density plot indicating an oscillation
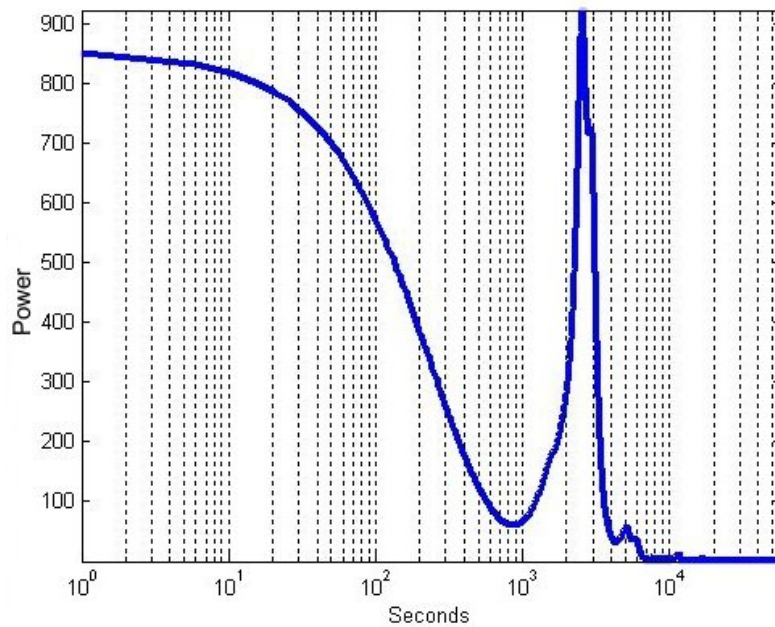
Figure 2-10: Power spectral density plot with both oscillations due to over tuning and stiction

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2.5.7 Summary of Oscillation Detection Methods

Both the normality index and the Frequency of control error sign changes methods are calculations which accurately determine the presence of an oscillation. The Histogram method is a visual interpretation of the distribution of the data. This method can also detect other problems present in control loops as well as the presence of oscillations, but is unable to determine the cause of the oscillation

The Cross Correlation function gives an indication as to whether the oscillation is due to valve problems (stiction or over tuning) or due to disturbances. If the type of valve problem needs to be identified more accurately the input-output pattern recognition can be used to determine if stiction is present as indicated in **Figure 2-8b**. The Power Spectrum analysis can be used to determine the frequency of the oscillation which can then be used to infer the type of oscillation, if due to a stiction or tuning problem, since tuning problems have a higher frequency of oscillation than stiction.

2.6   Plant-Wide Performance Index (PWI)

All the methods discussed thus far have been single loop assessment tools. The desire for a single-valued plant-wide performance index (PWI) has become evident as it is virtually impossible to evaluate each and every loop in order to determine the efficiency of the plant. The PWI is an indication of how well the plant is performing, expressed as a percentage of its inherent "optimum". This is achieved in various ways, but the general method is to define an optimum for specific variables on the plant and compare those to the values actually achieved for that plant. These variables are not necessarily controlled variables.

The requirements for a desirable performance assessment are (Salsbury, 2005):

- Non-intrusive
- No training period or historical data requirement
- Simple to implement.

The main value factors for any processing plant are (du Toit, 2005):

- Quantity of feed entering the plant
- Quantity of products leaving the plant
- Quality of products (or inferred quality often using temperature)
- Utilities and other processing costs

These factors are therefore part of a plant optimisation exercise and using the comparison between the actual and the optimal condition would result in a generic number that can be used as an indication of how close the plant is operating to its inherent optimum.

With this in mind and the main value factors, a weighted expression, equation **(2.20)** was developed (du Toit, 2005), which is constrained by equation **(2.21)**.

$$PWI = 100 \left[\begin{array}{c} w_1 \sum_{i=1}^{m} \dfrac{\int_{t_a}^{t_b} Product_{i_{actual}} dt}{\int_{t_a}^{t_b} Product_{i_{optimal}} dt} + w_2 \sum_{i=1}^{m} \dfrac{Quality_{i_{actual}}}{Quality_{i_{optimal}}} + \\[4ex] w_3 \sum_{i=1}^{p} \dfrac{\int_{t_a}^{t_b} Feed_{i_{actuall}} dt}{\int_{t_a}^{t_b} Feed_{i_{optimal}} dt} + w_4 \sum_{i=1}^{q} \dfrac{\int_{t_a}^{t_b} Utilities_{i_{optimal}} dt}{\int_{t_a}^{t_b} Utilities_{i_{actual}} dt} \end{array} \right]$$ **(2.20)**

$$w_1 + w_2 + w_3 + w_4 = 1$$ **(2.21)**

From examination of equation **(2.20)** it is clear that this formula is dependent on the control engineer's choice of the weights, w_i, and on the choice of the optimal operating conditions. While this is not a design limitation, this very fact lends the equation to the versatility of being applied to and refined for various production facilities.

Since the optimal values used in the denominators and one numerator of the weighted terms in equation **(2.20)** are not always a plant constraint but a realistic achievable optimum, it may be possible for the plant to be operating better than what is defined as the optimum for a certain period. This would then result in an improved performance because this period of better than "optimal" results in a cancelling effect for the periods where the plant is performing less optimally. This can be rectified in one of three ways.

- Firstly the "optimal" point could be seen in the same light as a setpoint, in that the plant should be driving toward this defined optimum. The effects of the better than "optimal" performance can then be removed by using

an absolute of the error approach, this would then result in a poor performance being defined for these periods of better than the "optimal".

- Secondly if the plant operates at better than the "optimal" defined condition, the effects of this improved performance period can simply be ignored by making all the periods that are better than "optimal", equal to the optimal condition. This would then not result in a cancelling of the poor performance areas.

- Thirdly, this does however only occur when the "optimal" condition that is specified by the control engineer is not the plant inherent optimum performance. The best way to overcome this problem would be to redefine the optimum behaviour of the plant so that exceeding of the optimum is not possible. It must however still be an achievable optimum that is defined, as this would allow for different unit operations to be compared on the basis of how close to the inherent optimum the unit is operating. Changing the optimum is however not the obvious answer since sensor errors could push the actual value to a position only marginally over the optimum and then one of the other two methods discussed would be a more appropriate correction for exceeding the defined optimum state.

To keep to a generic solution an indication of this "optimum" violation can be incorporated in the calculations and the user can be prompted to alter the optimum variables such that this is no longer the case, in other words the third option will be used as this is the most generic method.

The plant-wide performance method as described in equation **(2.20)** can also be used to evaluate specific unit operations within a plant. This will then result in a unit operation performance index (UPI). A PWI can then also be determined by taking a weighted average of each of these UPI's.

3 METHOD

There are a number of different methods available for determining single loop performance. These should be used in conjunction with oscillation detection methods since oscillations are one of the main causes of reduced performance mentioned in **Section 2.1**.

The performance method requiring the least prior knowledge of the system, which is also the least computationally intensive and generally accepted method for determining performance, is the minimum variance method with any of the adaptations applicable to the method.

The oscillation detection methods need to be used in tandem to obtain a desired result. Generally the frequency of control error sign change method can be used to identify the presence of oscillations on an on-line process.

Further investigation into the source of the oscillation would be done using the cross correlation function. This would then identify whether a disturbance or a valve problem is causing the oscillation as discussed in **Section 2.5.4**. If a valve problem is the cause it would be required to further investigate the source of the oscillation using either power spectrum analysis or an input-output pattern recognition method to determine whether the valve problem detected is due to stiction or to incorrect tuning parameters applied to the control loop.

The plant-wide performance index can be used to compare different unit operations. It can also be used as a basis for the comparison of the change in performance of single loops to that of the change in the plant-wide index as will be illustrated in this investigation.

3.1 Relationship between Plant and Controller Performance

Once the method used to determine the plant performance has been established, it can be used to compare the change seen in the plant-wide performance relative to that of the changes in performance of the individual controllers that resulted for a specific and repeatable set of operating conditions. In this case the tuning parameters of each of the controllers will in turn be changed and the effect of this change on the plant performance and on the performance of the controller will be compared.

Ideally it would be desirable to rank the controllers within a plant in the order in which changes to the loops would result in the largest change in the performance of the plant. Unfortunately most of the single loop assessment methods available do not achieve this. Possible ways of overcoming this problem would be to use any of the single loop assessment methods and then to calculate a ranking based on any of the following factors:

- One possible factor could be based on an indication of whether the loop is oscillatory or not. This is however not applicable to loops that are specifically designed to allow for oscillations, such as buffer tanks. This factor would then automatically define loops that are oscillatory as being a large source of the degradation of the plant performance and will allow for these loops to be dealt with swiftly.

- A further factor that can be used to identify problematic loops is whether they are in a normal operating mode or not. Loops that are not in a normal mode would then require investigation as to the reason for the controller being turned off, keeping in mind that when a controller is turned off, the

controller loses it ability to contribute towards driving the plant to its inherent optimum.

- Yet another possibility would be to add a factor which ranks the loops based on the number of alarms associated with each loop. This would not necessarily result in a performance improvement by dealing with the loops, but would certainly reduce the number of alarms that an operator is faced with and would then allow the operator the time to deal with the loop alarms that are in fact affecting the performance of the plant.

- The loops can also be sorted according to the performance that is calculated for each loop. Methods that can be used for single loop performance include:

  - Minimum variance
  - Generalised minimum variance
  - ISE, IAE or ITAE

From the above factors the ranked single loop performance values would then be defined by equation **(3.01)**.

$$ SLP = MVI \left(1 - \frac{\left(Oscillation + \dfrac{time\ in\ AUTO}{Total\ time} + \dfrac{Alarms}{Total\ Alarms} \right)}{Total\ Number\ of\ Factors} \right) \qquad \textbf{(3.01)} $$

In equation **(3.01)** Oscillation is either a 1 or a 0, since the loop is either oscillatory or not. The single loop performance (SLP) can be based on any of the single loop performance criteria. In this investigation the minimum variance index (MVI) has been used.

The effect of this sorting of loops will be investigated by improving the controller parameters for the loops in question and then observing the resulting effect on the performance of the plant. From this it can be seen if the loops with the largest potential for improvement do indeed result in the largest improvement in plant performance.

3.2   Software Development

A software interface developed by a previous investigator (du Toit, 2005), was used for the evaluation of the performance evaluation method discussed. As part of this investigation, the software was adapted to include the proposed methods for single loop evaluation, oscillation detection and improvements to the plant-wide performance monitoring method.

This software was developed in the Matlab® environment to allow for easy programming and reprogramming as required. The software makes use of a graphic user interface to allow for the evaluation of various data sets.

Some minor software changes to make the interface more user-friendly were also implemented. These changes include:

- Replacement of the user input to prompting the user to input the required data sequentially. This prevents the program from giving errors and helps the user to make the required inputs.
- The program has been improved to allow the user to load data from any directory.
- The original program has been corrected to determine the menu variables from the data supplied, enabling the assessment of different unit operations.

More to the point are the changes that were made in the form of performance evaluation tools.

- Firstly, it was required to adapt the program to accept information in a different format. The original program was designed to operate by obtaining data from a *.olf file. The program has now been modified to also accept *.mat files. This improvement allows the program to evaluate the different installations in the laboratory, as their output data format is *.mat.

- The performance definition for each plant is defined separately in different files allowing for each unit to be tested with the same method which is specific to each unique performance requirement for the different units. This improvement was brought about to make the program applicable to more than one unit operation.

- Additional performance criteria that were added to the controller performance assessment include ISE, IAE, ITAE and the generalised minimum variance (GMV).

- Oscillation detection methods added to the single loop assessment portion include the cross correlation function and input-output pattern recognition. The addition of these methods allows for the determination of the source of an oscillation in a control loop.

- The identification of incorrect "optimum" values in the performance statement will be done to ensure the most appropriate calculation of the plant-wide index (PWI).

All the files required for the evaluation of both plant performance and single loop performance are included on a CD which accompanies this dissertation. The specific purpose of each file is summarised in **Appendix B**.

3.3 Software Interface

Figure 3-1 shows the Matlab® graphic user interface which was developed for the evaluation of different unit operations. This is the single loop evaluation interface, which will be used to evaluate the source of oscillations and the different single loop performance methods available.
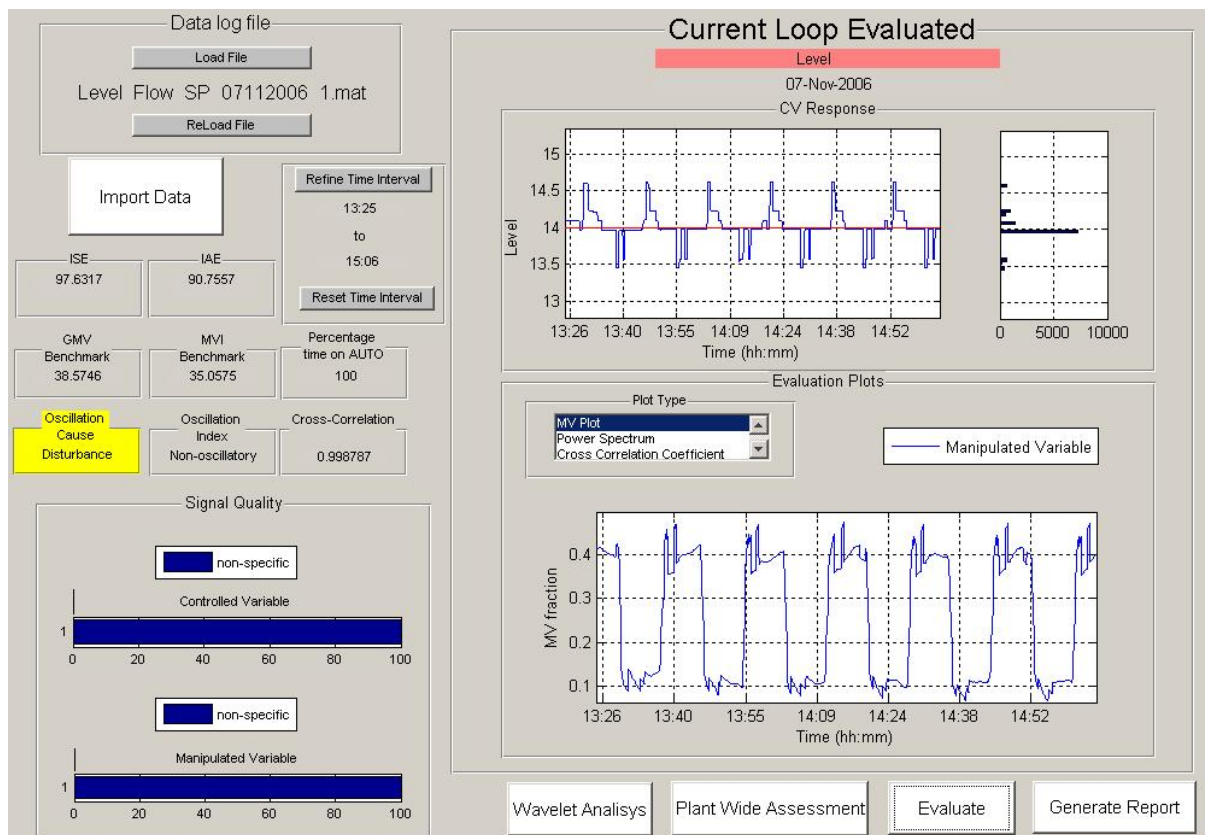
Figure 3-1: Single loop assessment interface

Having a closer look at the left hand side of **Figure 3-1**, **Figure 3-2** is obtained.

45
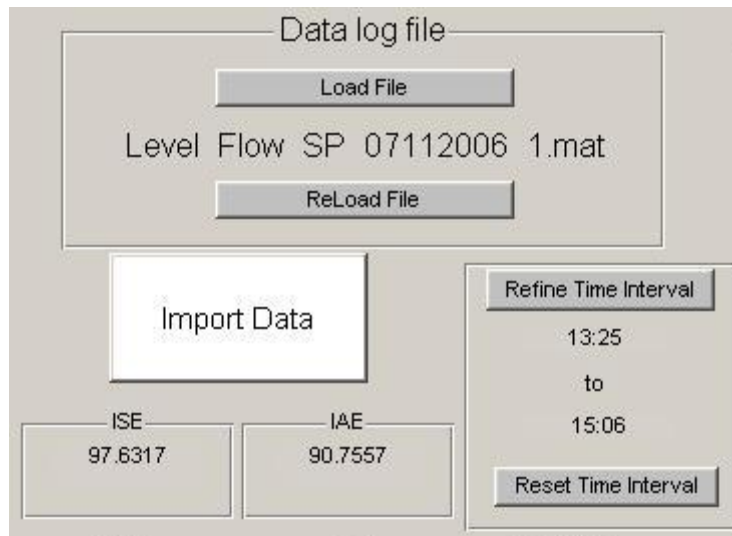
Figure 3-2: Single loop assessment – Top left

Figure 3-2 is the top left portion of **Figure 3-1**. This is the user interface to load the data files that have been generated on different units, the Import Data, button is used to select between the different single loop controllers available within the data set and to the right of that is a refine time portion which is used to zoom in on a specific portion of the data that is of interest.

Figure 3-3 is the middle left portion of **Figure 3-1**, this is for the evaluation of different single loop performance methods and oscillation detection methods.

Figure 3-3: Single loop assessment – Middle left

In **Figure 3-3** the **ISE** value is calculated by taking the sum of the error (deviation from setpoint) for the controller in question and over the specific period of time and comparing that to a maximum allowable deviation from setpoint defined as 5%. This then results in a percentage which specifies how well the controller is operating. The **IAE** uses the same method of calculation as the **ISE** except that it is done using the absolute value of the error. **GMV*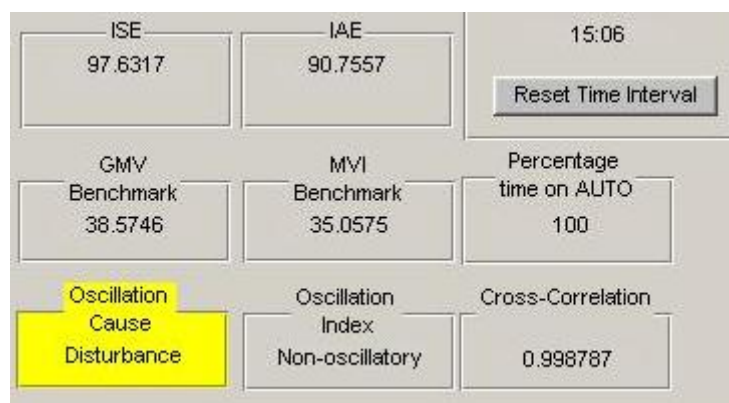* and **MVI** are the generalised minimum variance and the minimum variance index, these are calculated using the methods discussed in Chapter 2.

In **Figure 3-3** the **percentage time on AUTO** gives the user an indication as to how much time the controller was in the AUTO mode, if the controller was not in AUTO for a sufficient period then evaluation of the controller from the data available will not be possible, since the evaluation methods rely on the controller actually being in control of the controlled variable.

The bottom row of values in **Figure 3-3** are the methods used for detection of oscillations. **Oscillation Cause** is calculated using the method described for the Cross-correlation method in Chapter 2. This value will be either, "disturbance", "valve", or "no decision", depending on the value calculated using the Cross-correlation method and thereby indicating the source of the oscillation. **Oscillation index** is calculated using the frequency of control error sign changes method is discussed in Chapter 2. This indicated whether the loop being assessed has any oscillations present and if further evaluation of the data to determine the source of this oscillation will be required, the results given in **Oscillation Cause** then become of interest. The **Cross-Correlation** is a built in Matlab® function that is used to determine a correlation between the manipulated variable and the controlled variable, this value should always be close to 1, to indicate that the controller is indeed having a large influence on the controlled variable.

The bottom portion of **Figure 3-1** gives the signal quality for the period of evaluation. This gives the user an indication as to how reliable the data being evaluated is.

Figure 3-4 is the right had portion of **Figure 3-1**. This is used to do graphical analysis of the data.
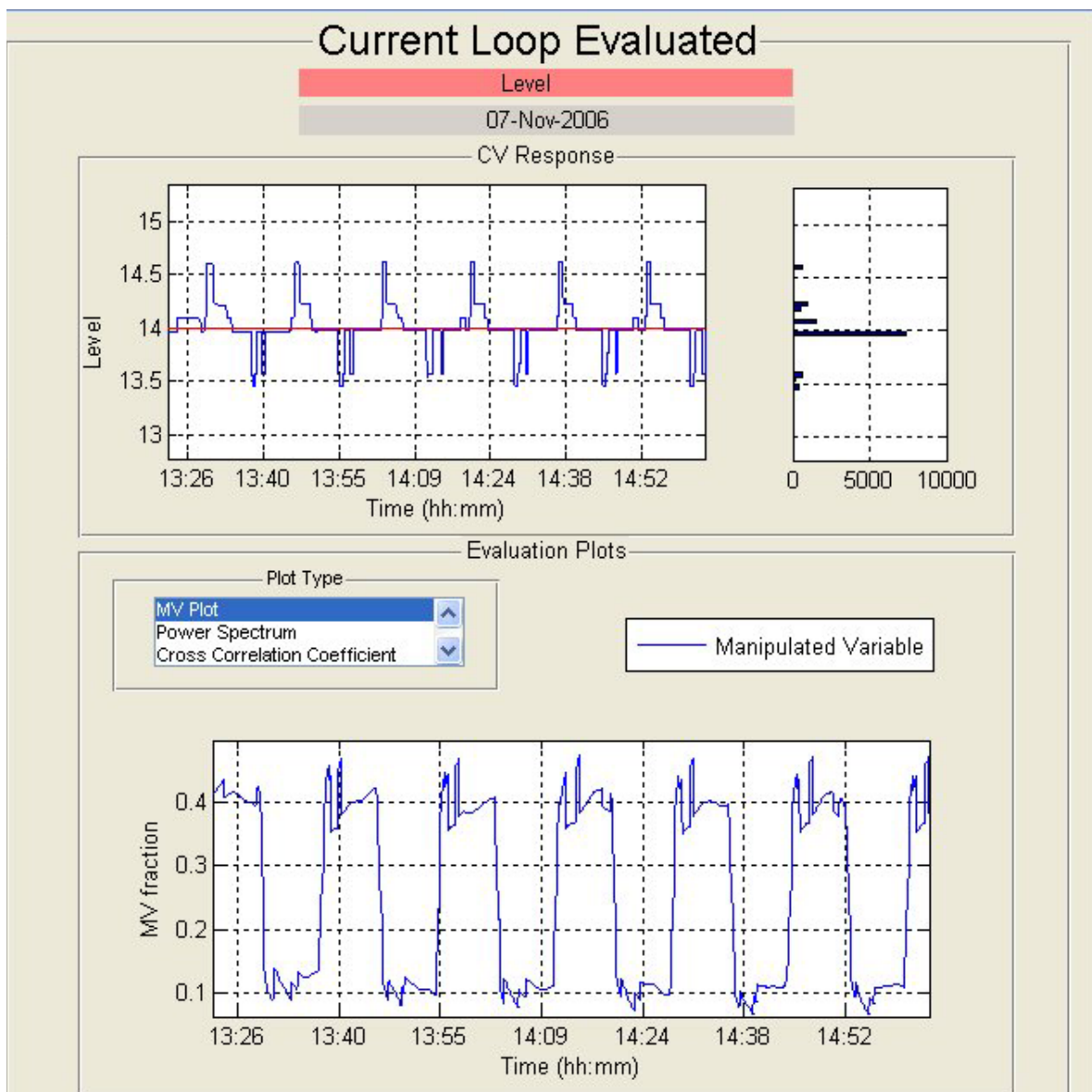

Figure 3-4: Single loop assessment – Right

In **Figure 3-4**, the top portion gives information about the control loop being evaluated. The top figure shows the setpoint and the process variable for the specific control loop being evaluated, and a sideways histogram of the process variable is shown on the right of the figure, this is used to analyse the data, as discussed previously.

The lower graph in **Figure 3-4**, can be changed to one of many different graphical analyses. One is the manipulated variable plot for the controller. Another is the power spectrum plot of the data being analysed. A cross correlation plot over the period of analysis can also be generated. The fourth plot available is the input-output pattern recognition plot, which is generated by plotting the manipulated variable against the controlled variable. These different figures are selected by the user depending on what graphical representation of the data is required for visual analysis of a specific control loop.

Figure 3-1 to **Figure 3-4** is the interface that is used to evaluate all the different unit operations, on a single loop basis. These are used for single loop performance evaluation and source of oscillation detection. The plant wide evaluation interface used for the evaluation of the different units is shown in **Figure 3-5**.
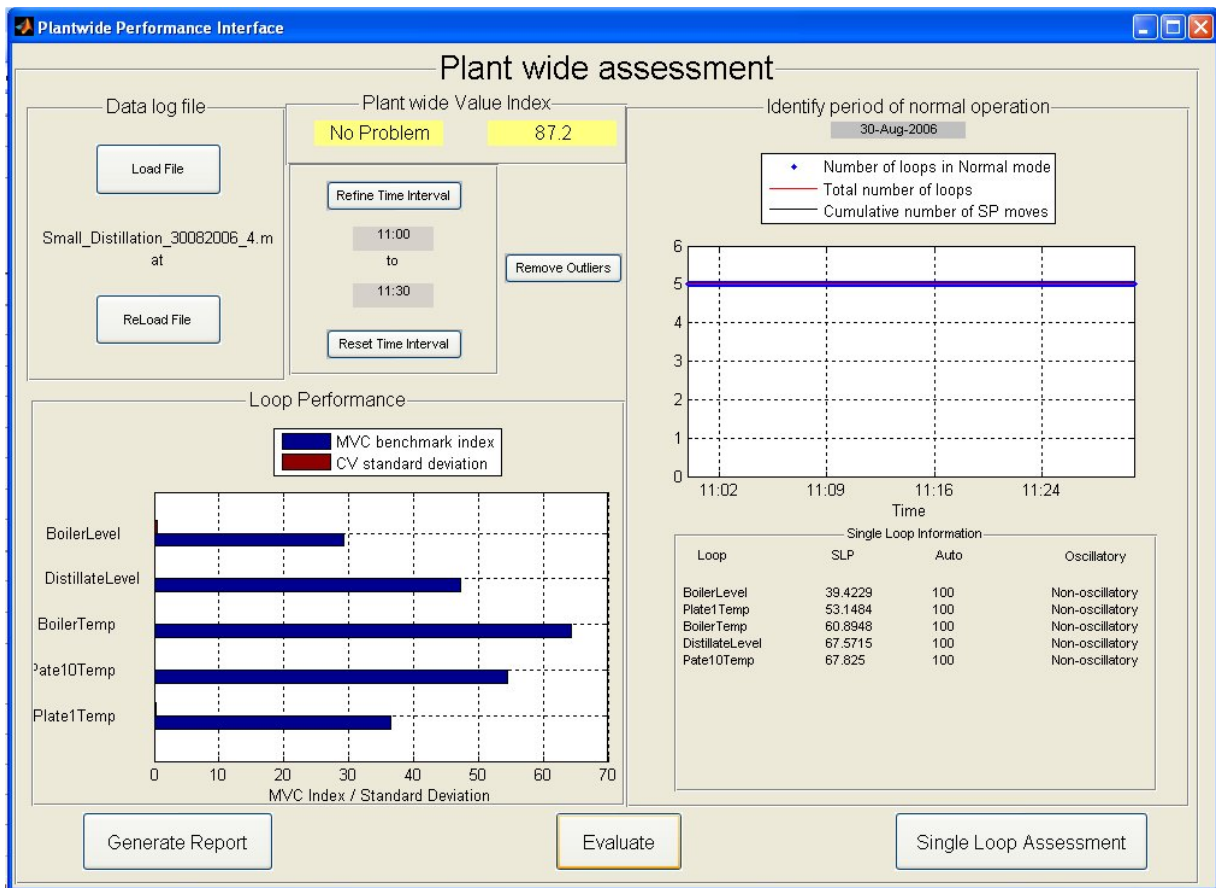
Figure 3-5: Plant wide evaluation interface

Figure 3-5 has a summary of the single loop assessments given in the lower portion of the figure, but the value of importance is the **Plant wide value index** given in the top middle of the figure. The actual value is calculated using **Equation 2.20**. The **No Problem** indicated to the left of the PWI is given to inform the user whether the "optimum" defined values used to calculate the PWI are acceptable, if not this would prompt the user to select more appropriate or realistic values for the PWI calculation.

3.4 Software Usage

After the addition of the different assessment methods to the software package the different unit operations available will be evaluated using this newly developed tool. This will be discussed in chapter 4.

Visual interpretation of the manipulated variables which are available for each data set will be done. This will assist in the interpretation of the other more abstract figures and results.

Comparisons can be drawn between different single loop assessment methods. The different ways of determining the presence and the cause of oscillations will be evaluated.

The plant-wide performance index will be compared for all the different unit operations available, this comparison will then determine how applicable this method is. The plant-wide index of one unit operation can also be used in conjunction with the single loop performances of that same unit to evaluate if any relationship can be found to relate controller performance to plant-wide performance.

4 EXPERIMENTAL

A Matlab Graphical User Interface (GUI) which was developed and was tested on numerous installations as will be described in this chapter. Different installations have been chosen to determine to what extent the plant-wide performance conforms to the proposed method and to confirm the general applicability of the method. This plant-wide performance will be validated by evaluating the single loop performance of all of the controllers in the unit and by degrading or improving the performance of these single loops to observe the changes with regards to plant-wide performance.

There are four different test rigs available in the Process Control Laboratory which will facilitate the evaluation of this method. These four units are: a level and flow rig, a temperature rig with variable deadtime and two distillation columns.

Each unit operation has specific elements which are of interest for this investigation. These are as follows:

- The level and flow rig has a control valve which is suspected to exhibit stiction. The control loops can also be tuned to oscillate in order to observe the effects and results obtainable from the different oscillation detection methods.
- The temperature rig was developed with a PI controller with dead-time compensation (Smith predictor) as an optional controller. The plant-wide performance index can be used to determine which type of controller is a better choice for the rig; that of the PI controller with dead-time compensation or the simple PI controller option.

- The two distillation columns can be used to compare the plant-wide index for two similar unit operations, but with largely different control structures and hardware. The smaller of the two will also be used to examine the relationship between plant performance and controller performance.

The reason for using multiple units is for comparison purposes. In this way the PWI determined for each unit can be compared for relative differences in performance, to see if the index ranks the different units appropriately in terms of worst performing units giving the lowest PWI. This evaluation is required since all the units have different objectives which are evaluated. The use of a similar unit allows for an easy comparison of at least these two units.

4.1 Large Distillation Column

This is the installation that was used for the initial development of the software (du Toit, 2005).

The Piping and Instrumentation Diagram (P&ID) for this installation is shown in the Appendix, **Figure A-1**.

The control for this unit is done in such a way that the top plate temperature is controlled by manipulating the reflux to the column. For inventory purposes the level in the distillate drum and the boiler are controlled; the distillate level is controlled by manipulating the product flow rate out of the column, while the boiler level is controlled by manipulating the bottoms product flow rate from the unit. The temperature in the boiler is controlled by manipulating steam flow rate through the reboiler and the feed rate to the column is kept constant with a flow controller.

This installation was used to verify the results obtained by (du Toit, 2005), also to compare the plant performance index derived for this distillation column with that of other unit operations and to determine successful tuning parameters for some of the loops as some of the current controllers oscillate, due to improper tuning.

4.2   Level and Flow Rig

This unit has a single input stream (two pumps in series) which splits into two streams, these two streams then come together again in the tank. The level in the tank is a controlled variable and the flow through one of the two input streams is another controlled variable. This setup has a large amount of inherent interaction.

The controllers for this system are simple PI controllers. The P&ID for this installation is shown in the Appendix, **Figure A-2**.

Flow control is achieved with the control valve in the same line as the flow measuring element and level is controlled with the control valve in the other line. Since the delivery lines to the two controllers have a single source, manipulation of one stream directly influences the flow through the other.

This specific installation was chosen as it was suspected that the control valves on this installation are prone to stiction. The presence of stiction enables the testing of various oscillation detection methods and to determine which methods can best identify the source of the oscillations. The tuning parameters of the controllers will be adjusted to obtain an oscillatory response. This result can then also be examined with the different oscillation detection methods.

The level control loop for this rig has much slower dynamics than the flow control loop. This difference in dynamics can help determine the most appropriate single loop performance criteria, for easy comparison of different control loops.

The objective of the system is to achieve a desired level in the tank while maintaining a desired flow rate in another leg entering this tank. This system is therefore operated under the same fixed setpoint change patterns for each run. This will allow for a period where the level loop has a constant setpoint, while the flow loop has a step pattern for the setpoint, followed by a period where the flow loop has a constant setpoint, while the level loop has a step pattern for the setpoint.

Since the system is interactive the setpoint changes in the flow loop will serve as disturbances to the level loop and setpoint changes to the level loop will consequently serve as disturbances to the flow loop. In this way the system can be operated with exactly the same disturbances for each of the different runs. This rig is therefore operated with the controllers in auto mode at all times.

4.3 Temperature Rig

This rig was chosen for performance monitoring as it has variable deadtime elements, which depend of the flow rate and different flow paths change the distance between the point of measurement and the point of control. This rig has both a PI controller and a PI controller with dead-time compensation as possible temperature controller configurations.

The P&ID for this installation is shown in the Appendix, **Figure A-3**.

The unit only has one control loop. The temperature at the measuring point is controlled by manipulating the energy flow to the heating elements in a down stream tank. This variation is achieved by making use of a thyristor.

Variable deadtime for this rig is achieved by the appropriate selection of valves HV002, HV003, and HV004 seen in Appendix, **Figure A-3**. The performance of the controller for two of these loops will be determined.

This rig can also be controlled by a PI controller with dead-time compensation so that the performance of a model based controller can be examined with the performance monitoring tool and compared with that of a system without the dead-time compensation.

4.4 Small Distillation Column

This installation is similar to the Large Distillation Column, which was the original system that the evaluation software was developed for, in the sense that this distillation column also has a feed of an ethanol-water mixture, but the control method, size of column and availability of measurements are all different. A similar performance objective will be used for this column and will thus be used to evaluate if the performance monitoring tool is comparable for similar units.

The P&ID for this installation is shown in the Appendix, **Figure A-4**.

This distillation column has a different control configuration to that of the large distillation column discussed earlier. The top plate temperature is controlled by manipulating the reflux flow. For inventory reasons the level of the distillate drum is controlled with the outflow of product, while the bottoms level is controlled by manipulating the bottoms flow. The difference between the two columns arises because this unit achieves temperature control of the boiler with an electrical unit and not steam. Secondly, the feed flow rate on this column becomes a manipulated variable for the control of the bottom plate temperature.

The desired experimental outcomes for this installation are:

- The plant-wide performance index can be examined for the principle of application to distillation columns.
- A comparison between plant and controller performance can be examined.

In order to achieve the objectives, a control philosophy had to be derived and then implemented. The controllers then required tuning.

In order to obtain the required information, the "steady state" data obtained under specific controller settings were compared to those of the "steady state" data obtained after a change in the controller settings was brought about. Changing the controller settings was the easiest and most repeatable way of degrading the performance of the controller and then comparing that with the performance obtained for the plant previously.

5 RESULTS AND DISCUSSION

This section gives the results obtained from the experiments that were discussed in the previous section as well as the actual experiments with the different controller settings or algorithms and specific disturbances or changes to each unit that were brought about. The various control structures are shown in the P&ID's for each rig given in Appendix, **Figures A1-4.**

For consistency the results of each of the different units will be presented in the same manner. The figures give the setpoint and present operating variable (a), the manipulated variable (b) and the input-output pattern recognition (c). All these plots are generated from the same data set and over the same time period, so the patterns observed in plot (c) are the patterns that arise for the entire period.

These specific plots are given so that the time response and the required change due to the manipulated variable can be observed. The input-output pattern recognition is given to observe the presence of oscillations and to determine the source of these oscillations. In most cases the oscillations are clearly visible in the time based setpoint and PV plot but the underlying cause will be identified from plot (c).

For all these figures plot (a) and (b) originate from the data that is being evaluated and plot (c) is obtained by plotting the input (from plot (b)) against the output (from plot (a)), which then results in the input-output patterns recognition plot that needs to be observed in order to determine the cause of an oscillation.

5.1   Large Distillation Column

The data obtained for this column is the "steady state" data (setpoint, process variable and valve position) for each control loop which is then compared to the data obtained for the retuned controllers. There is no disturbance specifically applied to this unit, but the steady state values are evaluated after a successful start up of the column. The difference between data sets will be controller tuning parameters. In this way it is possible to evaluate the plant performance improvement achieved by improving the individual controllers, and the extent to which this is possible.

Figures 5-1 – 5-5 are the results for each of the controllers associated with the distillation column with poorly tuned controllers. The minimum variance (MVI) for each loop is also indicated with the figures.
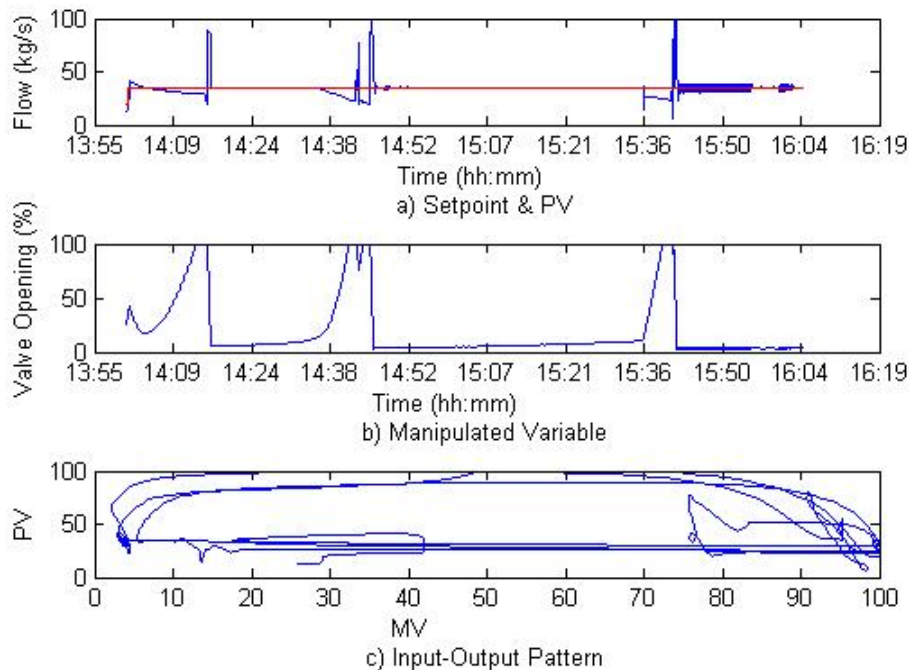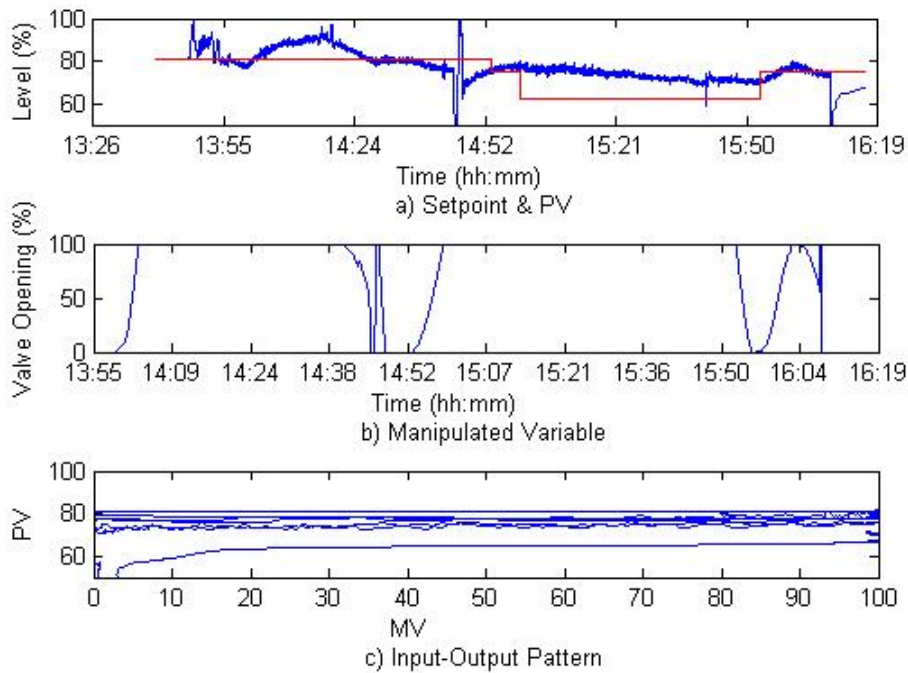

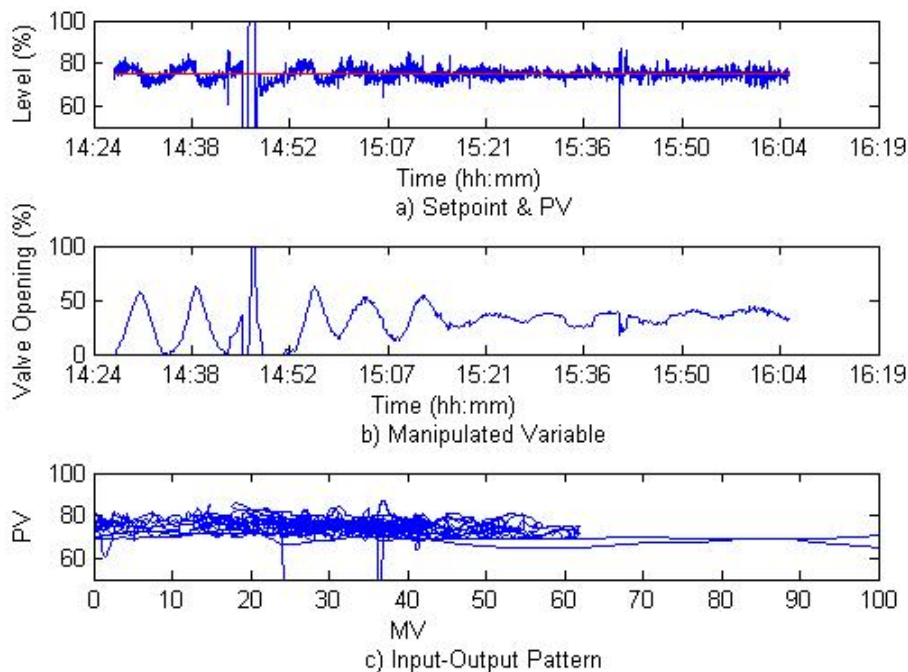**Figure 5-1: F001 data - MVI = 8.82**

Figure 5-2: L001 data - MVI = 8.89
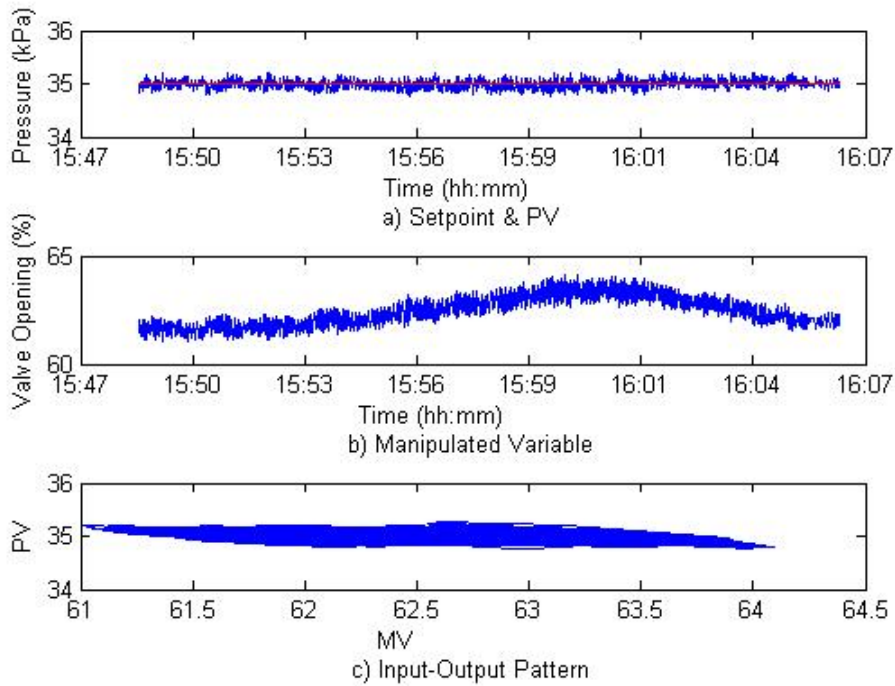

Figure 5-3: L002 data - MVI = 14.56

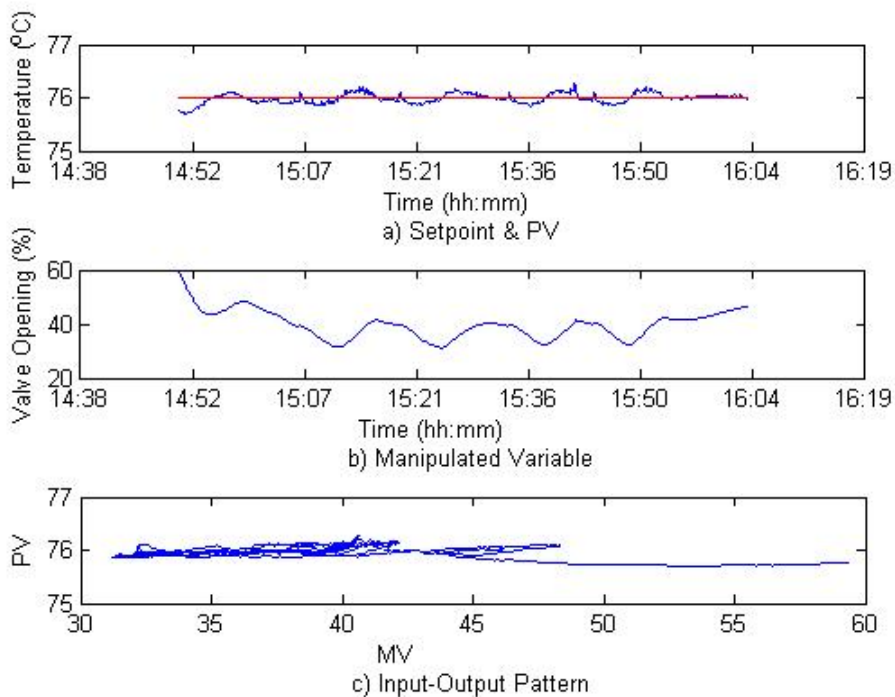**Figure 5-4: P001 data - MVI = 90.05**

Figure 5-5: T001 data - MVI = 56.69

The a) portion of the figures gives the setpoint and the process variable. This can be used to observe how close to setpoint the controller is able to keep the process variable. The controllers of **Figures 5-1 – 5-3** are clearly not achieving the desired output, as the variables are only drifting in the vicinity of the setpoint, and the slightly better **Figure 5-3** has large magnitude oscillations around the setpoint. **Figure 5-4** is tightly tuned and **Figure 5-5** shows better control but with some oscillations around setpoint which is a knock on effect observed from the distillate drum level control L002, being so poorly tuned.

In the b) portion of **Figures 5-1 – 5-3** clear valve saturation is seen. This is a good indication that the controllers for these loops are performing sub-optimally and will require re-tuning to improve their performance. **Figure 5-4** has very small magnitude changes, and **Figure 5-5** has an oscillatory response but this is likely the knock on effect seen from the L002 controller.

If the c) portion of each figure is examined, the circular pattern indicates clear oscillation and the cause of these oscillations is assumed to be incorrect tuning parameters for **Figures 5-1 – 5-3**. No oscillations are present in **Figures 5-4 & 5-5** as the patterns are linear, the linear pattern see in **Figure 5-5** confirms that it is not the controller that is causing the minor oscillations seen in the a) and b) portions of the figure but indeed as assumed a knock on effect from other controllers.

In **Figure 5-1 – 5-3** it is clear that the controllers have very little control over the desired variable as the manipulated variable is saturated for extended periods and are poorly tuned as oscillations are present. The MVI of each of these control loops is also very low, therefore this data represents a rather poorly controlled unit.

Making use of equation **(2.20)** with equal weights the PWI for this unit is given by equation **(5.01)**. The optimal terms being defined as:

- Possible Product – calculated from the actual ethanol in the feed
- Optimal Ethanol Quality – taken as pure ethanol for this calculation
- Optimal Water Quality – taken as pure water for this calculation
- Optimal Feed – taken as an operating state previously achieved
- Optimal Cooling Water – Theoretical requirement
- Optimal Steam Usage – Theoretical requirement

$$\text{PWI} = \frac{100}{6} * \left[\begin{array}{c} \dfrac{\text{Actual Product}}{\text{Possible Product}} + \dfrac{\text{Ethanol Quality}}{\text{Optimal Ethanol Quality}} + \dfrac{\text{Water Quality}}{\text{Optimal Water Quality}} \\[2ex] + \dfrac{\text{Actual Feed}}{\text{Optimal Feed}} + \dfrac{\text{Optimal Cooling Water}}{\text{Cooling Water}} + \dfrac{\text{Optimal Steam Usage}}{\text{Steam Usage}} \end{array} \right] \quad \textbf{(5.01)}$$

Using equation **(5.01)** on the data shown in **Figure 5-1 – 5-5** a PWI of 64.36 is obtained. An example of how this is actually calculated follows, in this calculation no instantaneous rates are used but the totals achieved over the period, as well as an average rate achieved over the evaluation period.

The possible mass product is calculated from the feed rate and the expected composition of the feed, which is used to determine how much product, is actually entering the unit. A value of 129.9 kg ethanol for the total time evaluated entered the unit and the actual product that left the top of the column is then measured as 71.5 kg for the same time period. The difference in product entering the column and product leaving the column would imply that a large amount of ethanol is leaving the column in the bottom product; this is undesirable and thus results in a poor performance value.

The quality of the top and bottom product are inferred from a temperature reading and compared with that of the pure component boiling point at the pressure in question, thus for ethanol 76 $^{\circ}$C is used and for water 100 $^{\circ}$C is used. The actual values are then the time-weighted average of the temperatures for the top and bottom product over the period being evaluated. This calculation results in a value which indicates how close the products are to the desired product streams, being pure ethanol and pure water. The azeotropic nature of the mixture limits the ethanol product quality and could possibly be refined to a more feasible solution with boiling points representing actual achievable product qualities, but for these calculations the pure component boiling points were used as optimal values.

The optimal feed amount is the setpoint for the feed controller which is multiplied by the evaluation time. The actual feed amount is then measured and summed over the period. This results in a total possible feed of 260 kg during the evaluated time, while only achieving a total feed of 147.7 kg. This lower than desired feed can be observed in **Figure 5-1** where there are periods of lower than desired feed rate, which has a net effect of lowering the total feed amount for the period of evaluation.

The optimal utilities are the theoretical energy required to operate this distillation column and achieve the desired separation. This indicates how much energy is required for the specific mass of feed entering the column. Any differences in actual to optimal imply inefficiencies or excessive reflux flow. These values could also be refined to a more accurate representation of what is physically possible for the column in question, but for these calculations the theoretical optimum values will be used, this is a value of 194.7 kg cooling water optimally required while 516.4 kg of cooling water was used to achieve the results shown.

The optimal steam requirement is determined as 20 kPa, while a 33.5 kPa is the average pressure that was required to achieve the separation.

The actual and optimal values are used in equation **(5.02)** to obtain the PWI for the experiment in question.

$$PWI = \frac{100}{6} * \left[\frac{71.5}{129.9} + \frac{72.3}{76} + \frac{82}{100} + \frac{147.4}{260} + \frac{194.7}{516.4} + \frac{20}{33.5} \right]$$

(5.02)

$$PWI = 64.36$$

The PWI for plant when operating with poorly tuned control loops is 64.36, which is effectively a percentage of the best possible performance for that plant. This value is not thought of as being too high as the controlled variable does at least attempt to remain within the vicinity of the setpoint. In this case the cause of the poor performance is apparent as some of the loops are oscillatory and have very low values for the MVI; this is observed in **Figures 5-1 – 5-3** and retuning of the loops in question is a definite requirement.

After visual observation of **Figures 5-1 – 5-5** it was clear that loops F001, L001 and L002 do not perform well. This is confirmed by low MVI values for these loops. After retuning loops F001, L001 and L002 **Figures 5-6 – 5-8** results. The other control loops P001 and T001 were left unchanged.
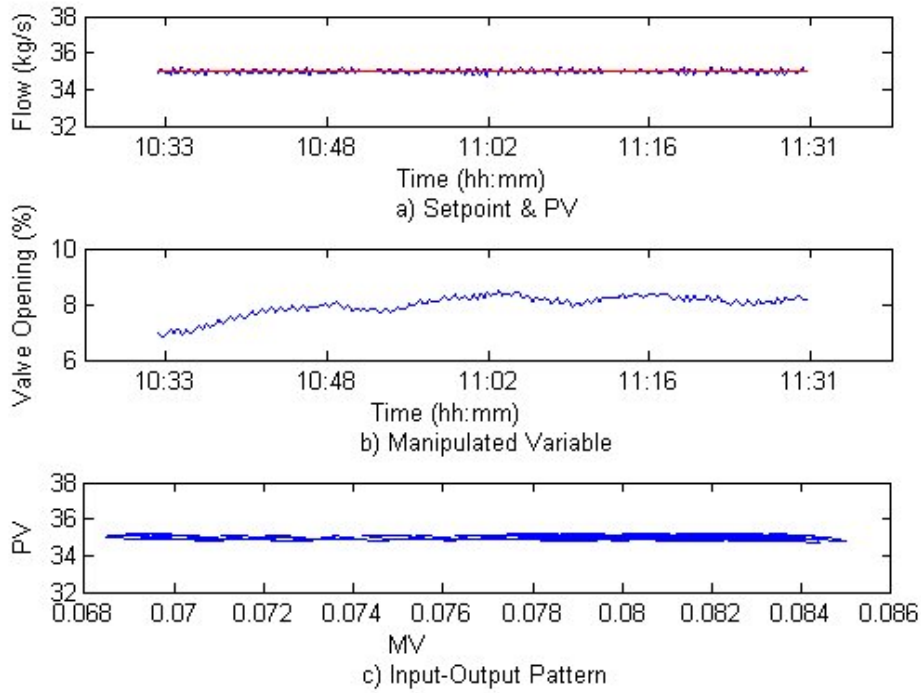
Figure 5-6: F001_Retune data - MVI = 84.16
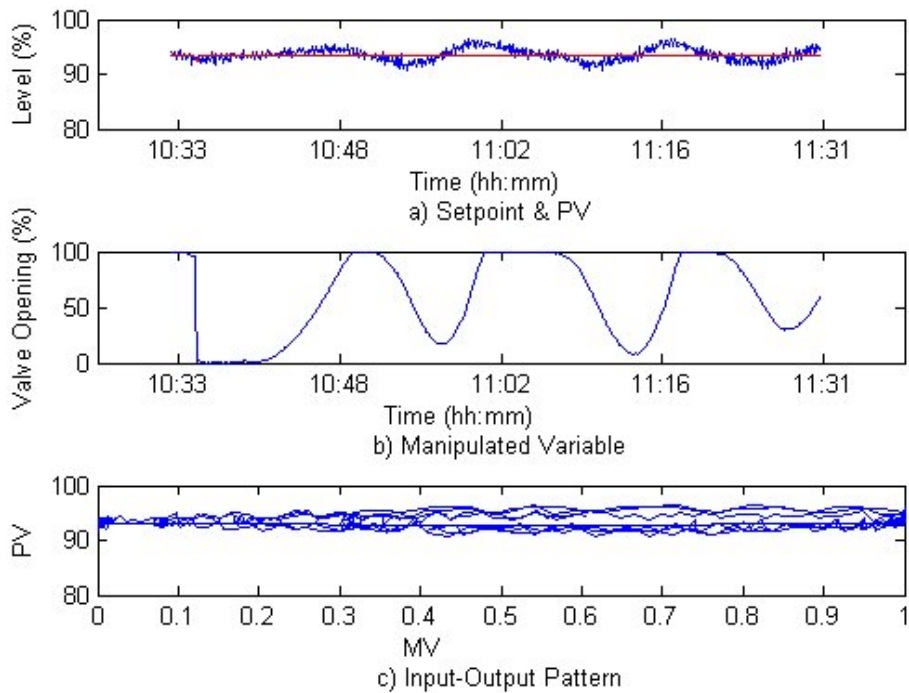
Figure 5-7: L001_Retune data - MVI = 66.36
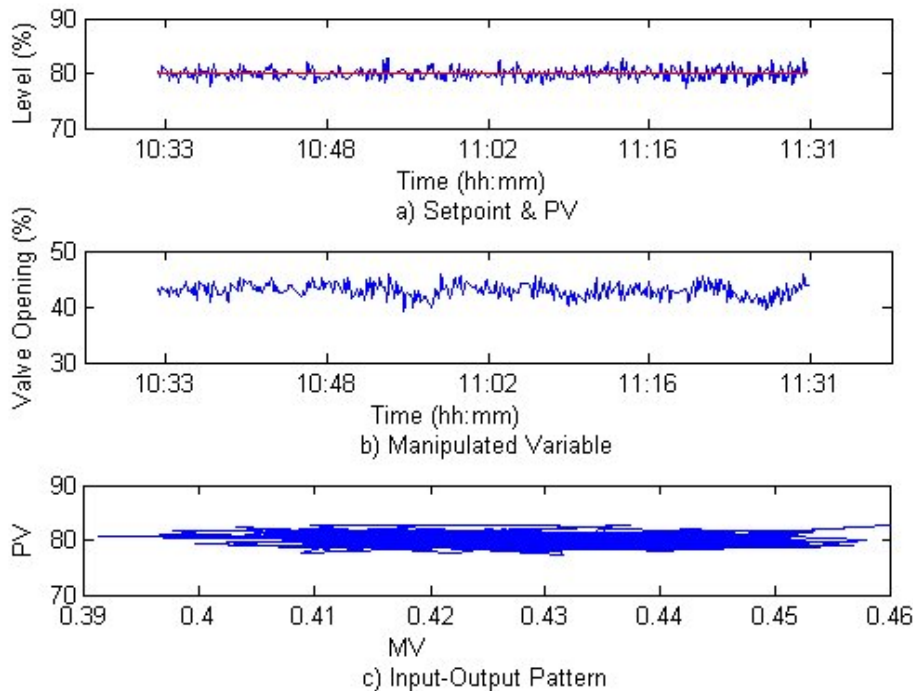
Figure 5-8: L002_Retune data - MVI = 83.51

In f**igures 5-6 – 5-8** it can be seen that there was a large increase in the MVI value as a result of the retuning these control loops. F001 showed an increase from 8.82 to 84.16 for the MVI value, L001 increased from 8.89 to 66.36 for the MVI value and L002 showed an increase from 14.56 to 83.51 for the MVI value.

From observation of **Figure 5-7** it is evident that further retuning is still required for this loop L001 as the output variable is still oscillatory as seen in the a) and b) portions and a circular pattern in the c) portion is still visible, but at least the level is being kept within proximity of the desired setpoint, the controller is also a proportional only controller, which makes the slight deviations from setpoint acceptable. The MVI value for L001 gives a good indication that further tuning is required and possible for this loop.

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The actual controller settings that were changed and the MVI values obtained for the retuned loops are all shown in **Table 5-1**. The plant-wide index for each of these data sets is also shown.

Table 5-1: Experimental information for the large distillation column

		F001			L001			L002		
	PWI	Gain	Tau	MVI	Gain	Tau	MVI	Gain	Tau	MVI
Poorly Tuned	64.36	2	8	8.82	5	0	8.89	5	2	14.56
Retuned	74.06	0.3	5	84.16	2	0	66.36	18	6	83.51

The PWI for the plant with retuned loops is 74.06 (**Figures 5-6 – 5-8**), which is a marked increase from that of the previous PWI of 64.36 (**Figures 5-1 – 5-5**).

5.2   Level and Flow Rig

The experiments that were performed on the level and flow rig are summarised in **Table 5-2** which gives the tuning parameters used to evaluate the Level and Flow rig. Each of these runs was performed with setpoint changes, at the same time and of the same magnitude. The only difference between the runs is the controller parameters as indicated.

Table 5-2: Level and Flow data

	Level Loop		Flow Loop	
Run	Gain (K_c)	Tau (τ_i)	Gain (K_c)	Tau (τ_i)
1	0.46	0.0022	0.1	0.1
2	0.3	0.0022	0.15	0.15
3	0.75	0.01	0.8	0.95
4	0.46	0.0022	0.075	0.075

In **Table 5-2** run 1 is a controller with normal controller settings, run 2 is with a controller that has improved controller settings, run 3 is an over-tuned controller to observe the oscillation patterns achieved and run 4 has a fixed period of oscillation disturbance affecting the level loop, applied in the form of setpoint changes in the flow loop. Run 3 and run 4 are done to verify the source of oscillations due to disturbances by making use of the input-output pattern recognition and the cross correlation method.

The PWI calculation for this installation had to be simplified since the main objective of this rig is to follow the desired setpoint. The ISE method was used to calculate the performance of the system. The actual calculation for the PWI of the level and flow controllers is given in equation **(5.03)**.

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$$PWI = \frac{100}{2} * \left[\left[1 - \frac{(\text{Actual level} - \text{Desired level})^2}{(\text{Desired level})^2} \right] + \left[1 - \frac{(\text{Actual flow} - \text{Desired flow})^2}{(\text{Desired flow})^2} \right] \right] \quad \textbf{(5.03)}$$

The results for run 1 in **Table 5-2** are shown in **Figure 5-9** for the Level data and **Figure 5-10** for the Flow data.
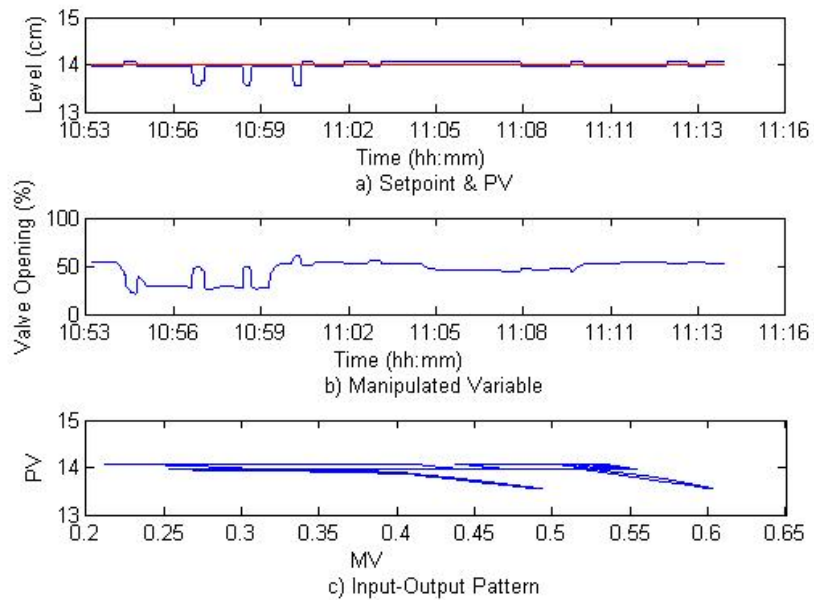

Figure 5-9: Level data for run 1

Figure 5-10: Flow data for run 1

The results for run 3 in **Table 5-2** are shown in **Figure 5-11** for the Level data and **Figure 5-12** for the Flow data.
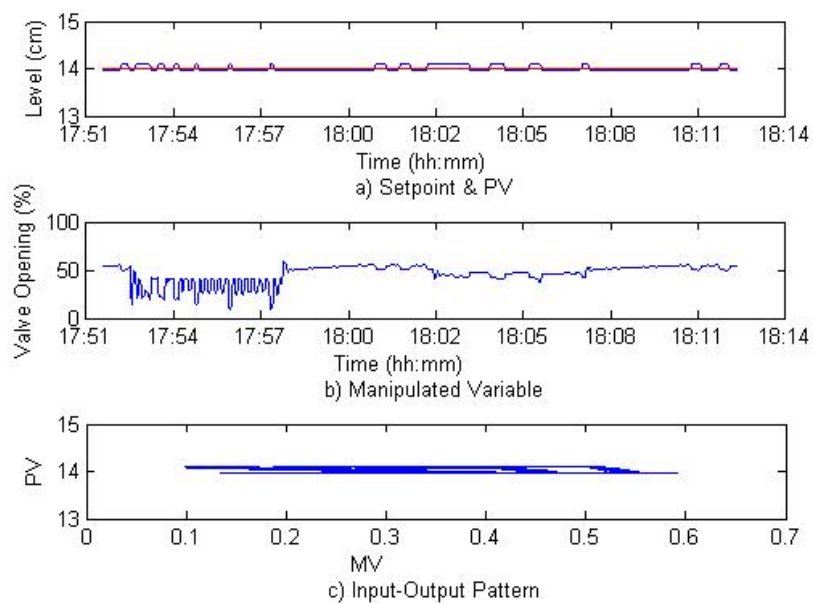
Figure 5-11: Level data for run 3
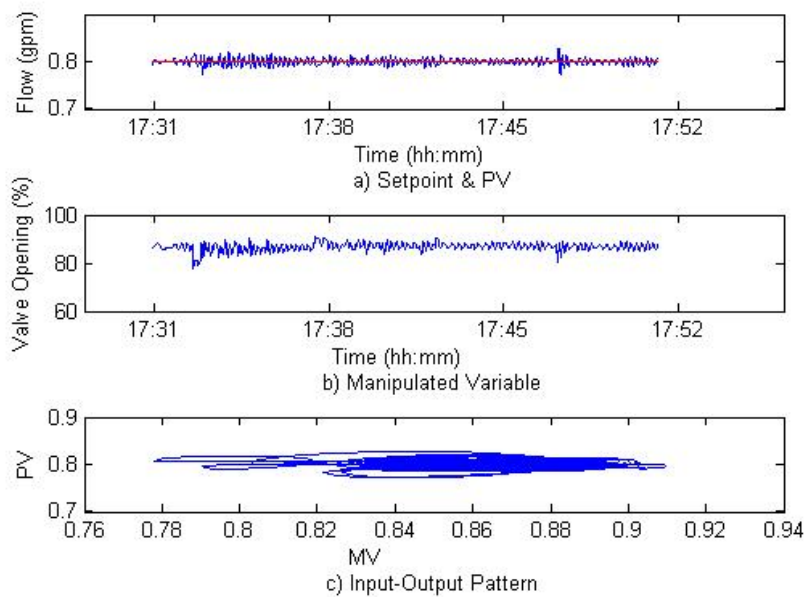
Figure 5-12: Flow data for run 3

The presence of oscillations is indicated by the sharpness of the edges of the input-output plot (c) in **Figure 5-9 – 5-11**, but **Figure 5-12** shows very smooth edges and a tightly wound pattern, except for a single sharper edge to the left of the figure. This smoothing of the edges is the direct result of over-tuning which results in excessive oscillations and thus a rounding of the pattern. If the manipulated variable plot (b) in **Figure 5-12** is observed it is also clear that an oscillation is present and since the only change between the two runs is the tuning parameters it is clear that a tightly wound pattern is the result of incorrect tuning parameters. **Figure 5-11** does have oscillations in the valve output for the first portion of the test but in the overall picture these are seen as oscillations due to stiction and not due to incorrect tuning parameters. The stiction on the level control valve is magnified by the incorrect tuning parameters used for the flow control valve because of the interactive nature of the system.

This confirms that input-output pattern recognition identifies oscillations due to stiction with sharp edges in the pattern and oscillations due to over-tuning with smooth tightly wound patterns. The presence of both over-tuning and stiction, which is the case in **Figure 5-12** is however not evident from the patterns as it appears that one of the two patterns can dominate. This would only be the case if the constant changes in the manipulated variable are relatively large which then minimise the effect of stiction as the actuator is supplied very quickly with enough energy to overcome the stiction.

The data for the control loops for run 4 as described in **Table 5-2** is shown in **Figure 5-13** for the level and **Figure 5-14** for the flow.
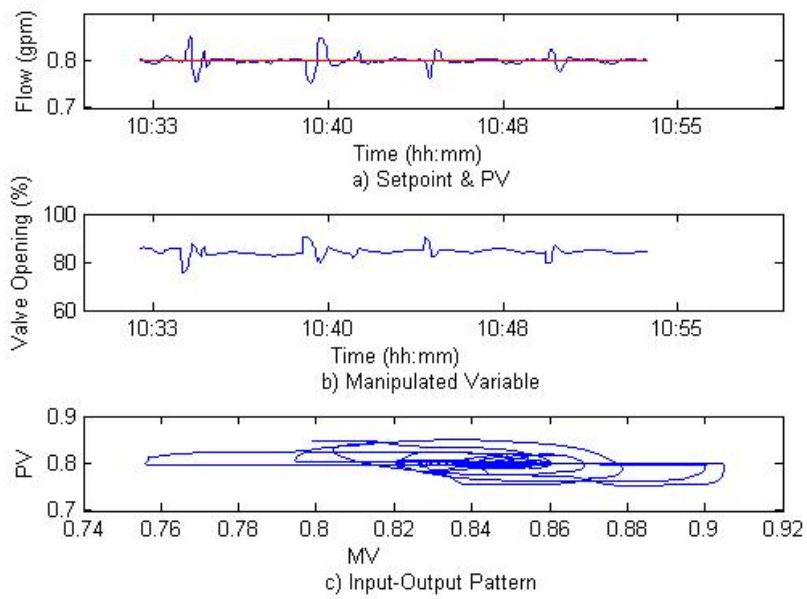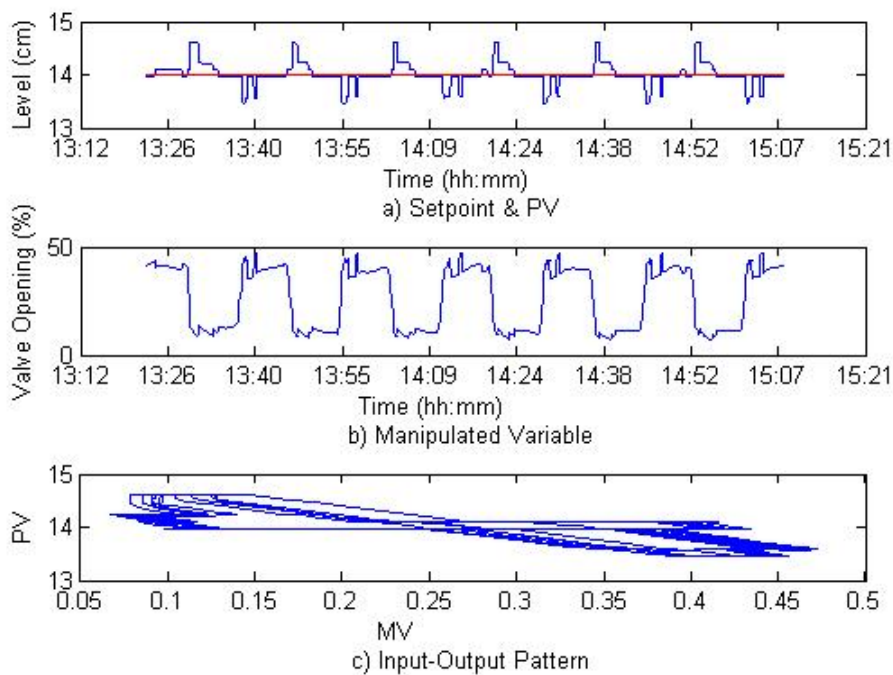

Figure 5-13: Level data for run 4
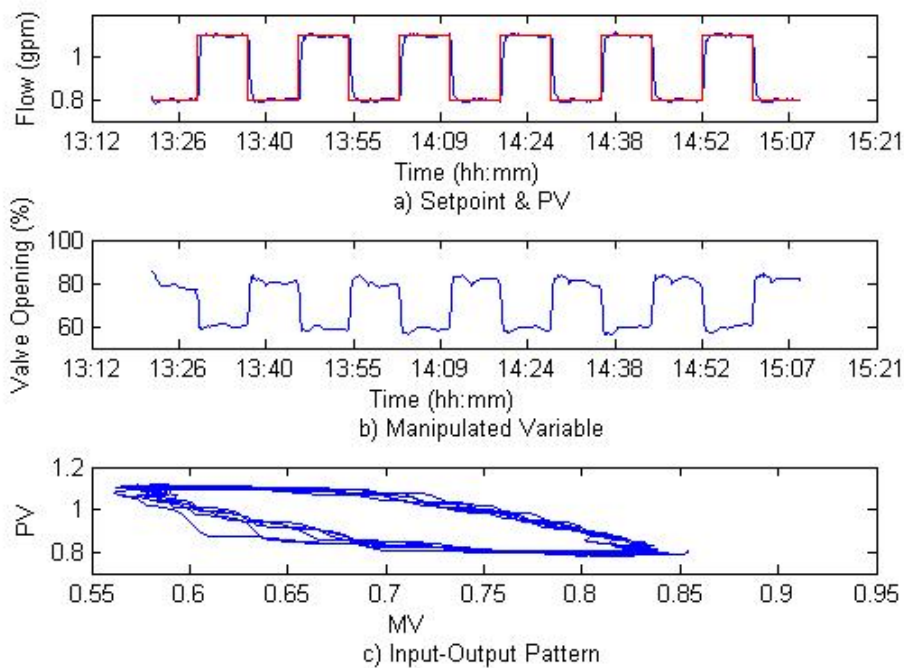
Figure 5-14: Flow data for run 4

The c) portion of **Figure 5-13** has a figure of eight pattern, which is seen due to the disturbance being applied to this control loop from an external source. In **Figure 5-14** the c) potion is only given to observe the circular pattern seen due to the cyclic setpoint changes that are applied. These oscillations are applied in the form of setpoint changes to this loop, but it can be seen how this method of input-output pattern recognition does indeed generate a circular pattern for oscillatory data.

Figure 5-15 is the view of the single loop performance interface, generated using the data for run 4, which is shown in **Figure 5-13 & 5-14**. From this figure it can be seen that the cause of the oscillation is due to a disturbance, as seen under the heading "oscillation cause" to the left of the figure which has been highlighted.
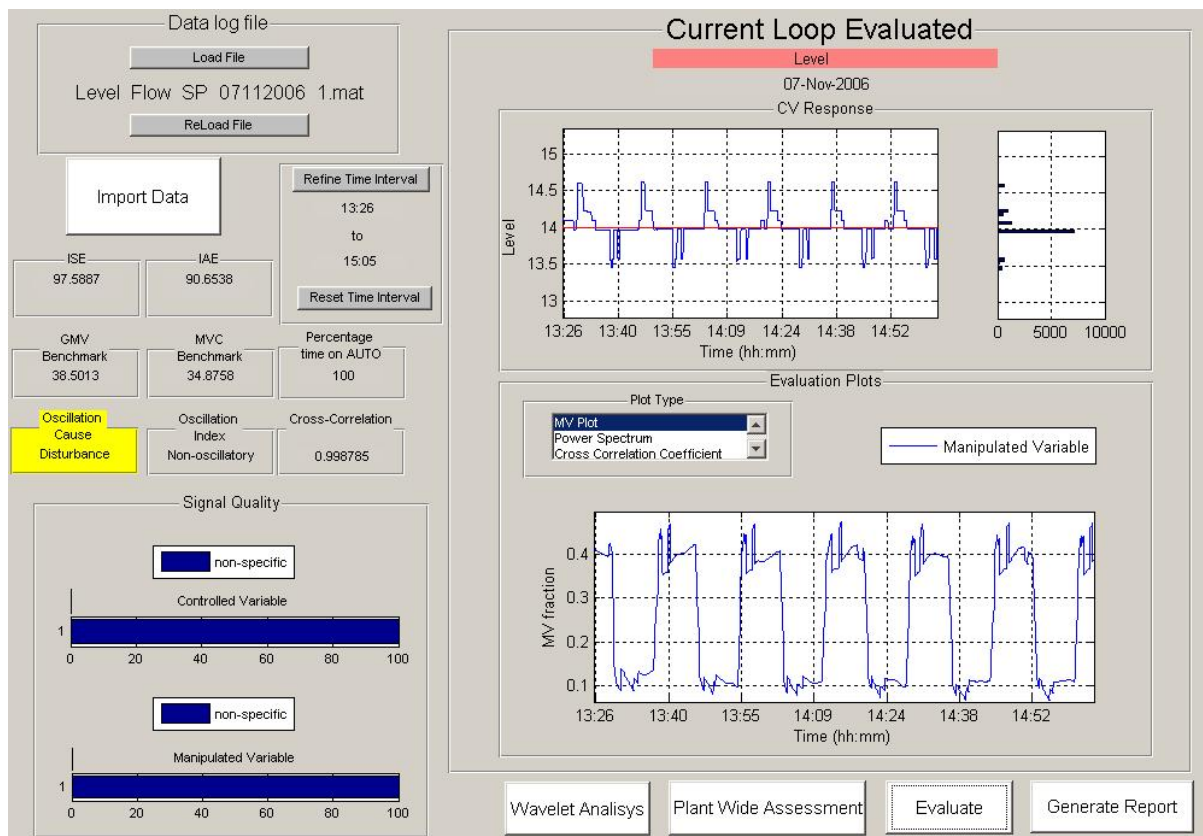
Figure 5-15: Level data for run 4

The cross correlation determination of the cause of the oscillation detected by the frequency of control error sign changes (Oscillation index in **Figure 5-15**) is given in **Table 5-3**.

Table 5-3: Detection of oscillations and the source

	Level Loop		Flow Loop	
Run	Status	Cause	Status	Cause
1	Non-Oscillatory	No Decision	Non-Oscillatory	Valve
2	Non-Oscillatory	No Decision	Non-Oscillatory	Valve
3	Non-Oscillatory	No Decision	Oscillatory	Valve
4	Non-Oscillatory	Disturbance	Non-Oscillatory	No Decision

The cross correlation function accurately determined that run 4 had oscillations present as a result of a disturbance, as indicated in **Figure 5-13**. The cross correlation function also accurately determined that valve problems are evident in the flow loop. Using the input-output pattern recognition it is possible to attribute these valve problems to stiction in the control valve for run 1, seen in **Figure 5-10**, and to incorrect tuning for run 3, seen in **Figure 5-12**.

The frequency of control error sign changes used to determine if oscillations are present as is evident from **Table 5-3**, only detects the oscillations that occur as a result of the over-tuning but is unable to determine the oscillations due to stiction. This would be because the magnitude of the oscillation that results in this case is less than that which is detected as a disturbance in the frequency of control error sign change method.

Table 5-4 gives the generalised minimum variance (GMV), the minimum variance (MVI), the integral of the square error (ISE) and the integral of the absolute error (IAE), for each of the loops. As well as the overall PWI achieved for each run, as defined previously in **Table 5-2**.

Table 5-4: Performance results for the Level and Flow rig

Run	PWI	Level Loop				Flow Loop			
		GMV	MVI	ISE	IAE	GMV	MVI	ISE	IAE
1	72.5	42.03	31.09	80.88	87.72	71.44	77.06	81.23	86.69
2	75.59	42.72	32.08	82.82	87.19	72.06	79.15	85.49	88.71
3	68.33	46.66	37.32	91.04	91	78.86	87.86	93.03	91.36
4	63.45	37.04	30.32	87.22	87.89	57.98	67.37	67.22	81.64

In **Table 5-4** it is seen that the level loop has a much lower GMV and MVI value than the flow loop and thus it would be thought that a tightening of the tuning parameters of the level loop would be required to improve the performance of the plant, but this is clearly not the case as run 2 shows improved plant performance for a slacking off of the tuning parameters for the level loop and tightening the tuning of the flow loop.. The reason for this is that the level loop has a slower response to setpoint changes than the flow loop, thus resulting in a lower GMV and MVI value and the flow loop has an influence on the level loop so the more constant the flow loop is the less disturbances the level loop sees and the more stable the entire plant is, as seen in the form of the highest PWI achieved from all the experiments performed.

The GMV value takes the movement of the manipulated variable into account, but does not have enough of an effect to sort the loops in a corresponding order to their influence on the performance of the plant. The ISE and IAE both rank the loops more closely to the influence that the loop would have on the performance of the plant.

5.3 Temperature Rig

This unit has a single temperature controller but various deadtime routes, the performance of a PI controller and that of a PI controller with dead-time compensation are compared as well as the performance seen for the two different deadtime routes. Each of these experiments was performed with setpoint changes, at the same time and of the same magnitude. The PI controller with dead-time compensation uses the same controller setting as the PI controller. Ideally a Smith predictor would allow for increased tuning parameters but the direct influence of implementing a Smith predictor on the PI controller is examined here.

Table 5-5 gives the controller tuning parameters and the selection of deadtime used for each of the experiments performed on the Temperature rig. The only difference between the different runs is the controller type (PI and PI with a Smith predictor) and the amount of dead time (Loop 1, short dead time or Loop 2, long dead time) as indicated.

Table 5-5: Temperature rig settings

Run	Controller	Dead Time	Loop Gain (K_c)	Loop Tau (τ_i)
1	PI	Loop 1	0.25	0.000100
2	PI	Loop 2	0.08	0.000015
3	PI with Smith	Loop 1	0.25	0.000100
4	PI with Smith	Loop 2	0.08	0.000015

The PWI is a simplified version of the plant-wide performance index, which effectively results in an ISE this is shown in equation **(5.04)**. This is the case since the loop has only one objective and that is to maintain a specific setpoint, the plant also only has one manipulated variable.

$$PWI = 100 * \left[1 - \frac{(\text{Actual Temperature} - \text{Desired Temperature})^2}{(\text{Desired Temperature})^2} \right] \qquad \textbf{(5.04)}$$

The data and results for deadtime loop 1 with a PI controller are shown in **Figure 5-16**, generated using run 1 settings and with a PI controller with dead-time compensation are shown in **Figure 5-17**, generated using run 3 settings.

In these experiments no oscillations are expected, but the c) portion of the plot is still given for consistency and to confirm this. Since these experiments are done with different setpoints throughout the evaluation time a different linear region is expected for each of these different setpoints if no oscillations are present.
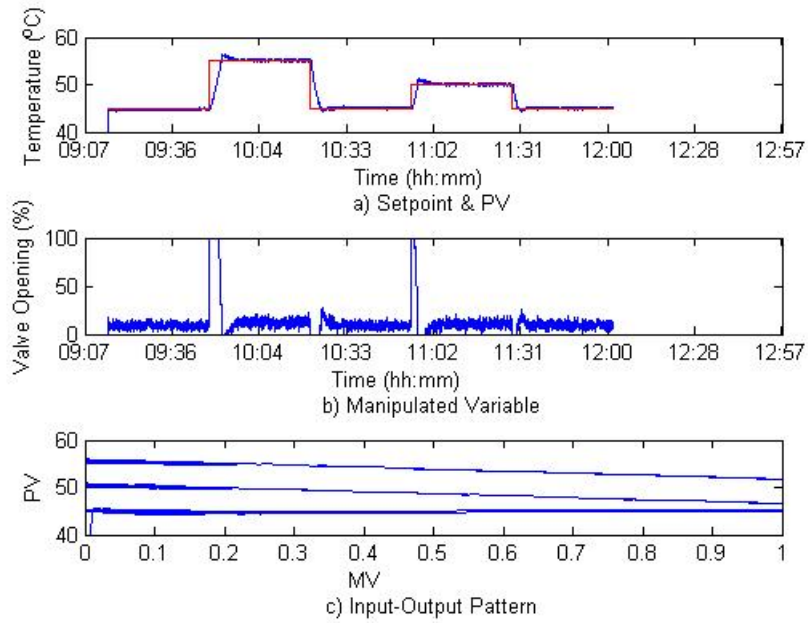
Figure 5-16: PI loop 1 data – run 1
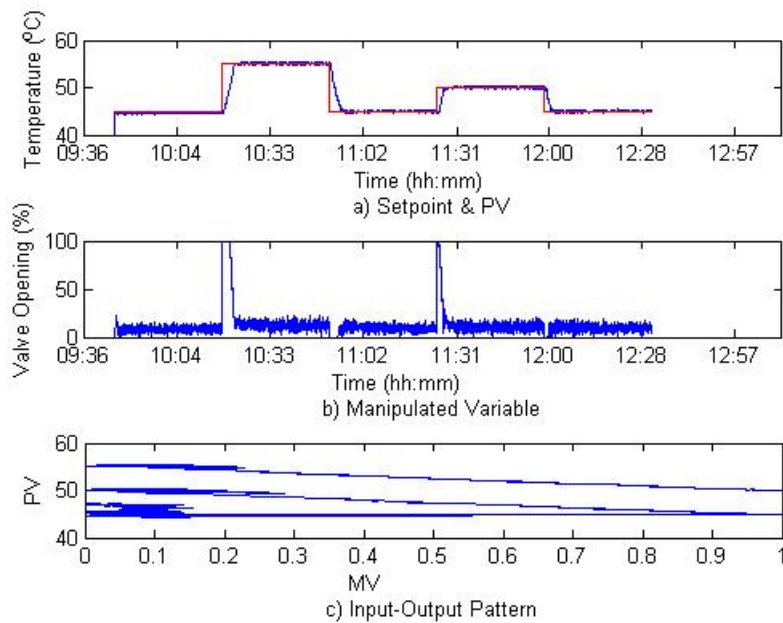
Figure 5-17: PI with Smith loop 1 data – run 3

The data and results for deadtime loop 2 with a PI controller are shown in **Figure 5-18**, generated using run 2 settings and with a PI controller with dead-time compensation are shown in **Figure 5-19**, generated using run 4 settings.

Figure 5-18: PI loop 2 data – run 2

Figure 5-19: PI with Smith loop 2 data – run 4

From plot (a) of **Figure 5-16 – 5-19** it can be seen that the PI controller with dead-time compensation results in less overshoot to a setpoint change than the PI controller. The PI controller with dead-time compensation does however take longer to reach setpoint.

The manipulated variable plot (b) shows that for the loop with increased deadtime (which is achieved by changing the flow path by opening valve HV003 and closing valve HV004 in **Figure A-3** in the Appendix) required smaller, slower changes to the system in order not to destabilise the system.

The PI controller with dead-time compensation makes even slower changes to the system than what the PI only controller does. This can clearly be seen in **Figure 5-18 and 5-19** after the setpoint changes, where the fraction opening due to the PI controller is much larger than that for the PI controller with dead-time compensation. The thyristor position is very aggressive with a fully open setting when heating is required. This type of action is required to increase the heating rate of the fluid, but needs to be limited to prevent large overshoots, especially due to the effect of deadtime.

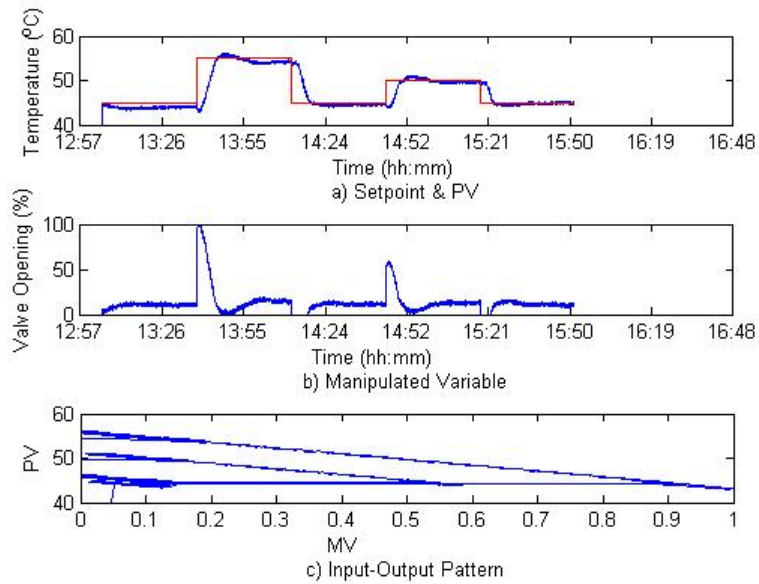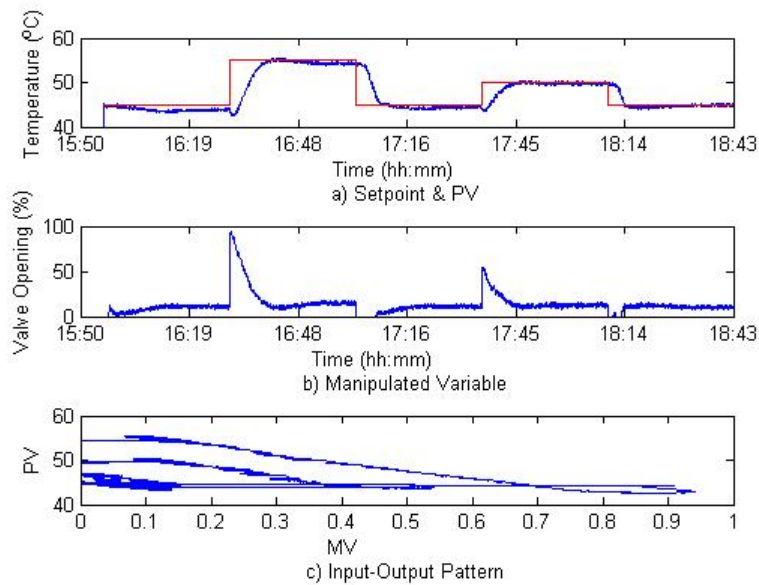The PI controller with dead-time compensation successfully achieves this and consequently has a more gradual increase in temperature which is clearly visible in **Figure 5-18 and 5-19**.

The plot (c) shows linear patterns for each of the setpoint regions in this case ($45\,^{\circ}$C, $50\,^{\circ}$C and 55°C), there are three different linear regions because three different setpoints were used in the generation of this data and in each of them the controller has no oscillations present and thus the linear pattern is observed. This is the expected result for these figures since the frequency of control error sign changes calculation for these loops also shows that no oscillations are present.

Table 5-6 gives the results obtained for evaluation of the single loop methods as well as the PWI calculated for each run.

Table 5-6: Temperature Rig Results

Run	Controller	Dead Time	GMV	MVI	ISE	IAE	PWI
1	PI	Loop 1	35.87	26.36	90.24	89.74	72.70
2	PI	Loop 2	30.67	16.44	73.68	73.77	68.81
3	Smith	Loop 1	37.40	25.69	89.77	89.86	72.58
4	Smith	Loop 2	32.70	15.82	70.78	72.41	68.17

When observing the PWI in **Table 5-6** for each of the different deadtime loops, the PI controller has marginally better performance than the PI controller with dead-time compensation. A value of 72.70 for the PI loop and that of 72.58 for the PI controller with dead-time compensation is a marginal difference but shows that the model used for the Smith predictor requires more attention if it is to be a more successful method for controlling the loop than a simple PI control algorithm. The same is seen for loop 2 with values of 68.81 for the PI controller compared to 68.17 for the PI controller with dead-time compensation.

This difference between the PI controller with dead-time compensation and the PI controller is likely due to the inaccuracy of the model used for the Smith predictor, the model used is developed from first principles which makes certain assumptions about unmeasured variables such as ambient temperature which results in minor model errors.

The only single loop performance measure that shows that the PI controller with dead-time compensation performs better than the PI control configuration is the GMV, this is the case since the PI controller with dead-time compensation has less movement on the manipulated variable, so if the movement of the manipulated variable is of importance then the PI controller with dead-time compensation would be a better choice in spite of the slight decrease in overall performance. The PI controller with dead-time compensation also has a smaller overshoot than that of the PI control algorithm.

Deadtime loop 2 has reduced performance and is evident due to an offset from setpoint. This is the case since the deadtime of the system greatly influences the performance of the controller. The controller does however maintain the value within an acceptable range of the setpoint. The PWI for the loop 2 deadtime is lower than that for the loop 1 deadtime as a result of this deviation, as seen in **Table 5-6** a value of 72.70 for loop 1 compared to a value of 68.81 for loop 2. The PI controller with dead-time compensation also shows this reduced performance for longer deadtime loops.

5.4 Small Distillation Column

For this unit the PWI is a desired outcome to successfully evaluate how well the PWI for the different distillation columns compares as well as the comparison with the different units tested.

The influence that a single poor performing controller has on the PWI is also observed. This will be achieved by changing the controller tuning parameters of each of the different single loop controller. Only one controller will be changed per experiment to successfully observe this.

The optimum conditions defined for this distillation column are the same as that defined for the large distillation column, except that the feed rate can no longer be fixed and therefore the setpoint cannot be used as the optimum value. The optimum value for the feed is then defined as the maximum throughput achievable for this column. The qualities of the top and bottom products are still achieved by comparing the actual temperatures with that of the boiling point temperatures for the different products. The utility usage is still determined for the amount of feed that enters the column, but for this unit no measure of the cooling water flow rate is available, so this element will be left out of the PWI calculation. The actual calculation for the PWI will then be given by equation **(5.05)**.

$$\text{PWI} = \frac{100}{5} * \left[\frac{\dfrac{\text{Actual Product}}{\text{Possible Product}} + \dfrac{\text{Ethanol Quality}}{\text{Optimal Ethanol Quality}} + \dfrac{\text{Water Quality}}{\text{Optimal Water Quality}}}{+ \dfrac{\text{Actual Feed}}{\text{Optimal Feed}} + \dfrac{\text{Optimal Electrical Usage}}{\text{Electrical Usage}}} \right] \quad \textbf{(5.05)}$$

Table 5-7 gives the controller settings for the control loops of the small distillation column as well as the single loop MVI values and the PWI for each of these changes.

The highlighted areas of **Table 5-7** are the changes that were made to each of the control loops, to observe the influence of single loop controllers on the plant-wide index. No specific disturbance was applied during these runs and the results are that of the steady state values which are achieved after a successful start up of the unit.

Table 5-7: Small distillation column tuning parameters and results

Run	Distillate Level			Boiler Temperature			Plate10 Temperature			Boiler Level			Plate1 Temperature			
	Gain (K_c)	Tau (τ_i)	MVI	Gain (K_c)	Tau (τ_i)	MVI	Gain (K_c)	Tau (τ_i)	MVI	Gain (K_c)	Tau (τ_i)	MVI	Gain (K_c)	Tau (τ_i)	MVI	PWI
1	0.45	0.0015	38.00	0.9	0.005	48.97	0.07	0.001	28.99	0.06	0.0002	39.66	0.1	0.0003	56.10	71.31
2	0.4	0.002	47.28	1.3	0.008	64.41	0.07	0.001	54.48	0.06	0.0002	29.24	0.1	0.0003	36.45	87.20
3	0.5	0.0025	46.48	1.3	0.008	60.81	0.07	0.001	47.10	0.06	0.0002	41.35	0.1	0.0003	39.61	86.03
4	0.5	0.0025	52.72	1.1	0.006	60.92	0.07	0.001	39.78	0.06	0.0002	28.61	0.1	0.0003	32.76	83.69
5	0.5	0.0025	46.48	1.3	0.008	57.32	0.07	0.001	27.24	0.06	0.0002	12.32	0.1	0.0003	16.76	77.80
6	0.5	0.0025	58.00	1.3	0.008	55.20	0.1	0.002	21.67	0.06	0.0002	8.64	0.1	0.0003	9.97	78.20
7	0.5	0.0025	57.46	1.3	0.008	66.01	0.07	0.001	22.13	0.06	0.0002	14.91	0.1	0.0003	7.88	79.86
8	0.5	0.0025	63.01	1.3	0.008	63.36	0.07	0.001	28.58	0.09	0.0003	26.37	0.1	0.0003	44.12	81.04
9	0.5	0.0025	63.69	1.3	0.008	54.73	0.07	0.001	28.68	0.06	0.0002	16.87	0.1	0.0003	52.11	79.54
10	0.5	0.0025	70.08	1.3	0.008	57.83	0.07	0.001	33.72	0.06	0.0002	30.17	0.2	0.0006	56.48	78.29
11	0.5	0.0025	56.46	1.3	0.008	58.49	0.07	0.001	28.68	0.06	0.0002	26.23	0.1	0.0003	50.10	78.03

An important element to note in **Table 5-7** is the effect the improving of Plate1 Temperature control has on the other controllers; this is seen in the run 10 and run 11 data. This increased performance results in a decreased performance in the boiler temperature, the PWI does however show a slight increase, but since Plate1 Temperature is directly related to the Quality term in the PWI it would be expected to be a larger increase but this is offset with the decreased performance observed in the boiler temperature which is the electrical term in the PWI calculation.

The weight assigned to such elements would therefore be important, in other words if better quality is desired the increased expense of utilities would be acceptable and the PWI could be increased more by assigning a larger weight to the quality than to the utilities. This type of decision is specific to each unit and that is why a weight option is available to customise the performance requirements for each unit.

Table 5-8 gives the order of the control loop influence on the plant performance. The values are calculated by taking the difference of the plant performance (PWI), found in **Table 5-7**, from before and after the changes to the controller settings of each loop, this is done for each combination of data, i.e. using run 2 and run 3 for the distillate level and run 4 and run 5 for the boiler temperature and so on.

Table 5-8: Order of influence

Loop	PWI Change
Boiler Temp	5.89
Plate10 Temp	1.66
Boiler Level	1.50
Distillate Level	1.17
Plate1 Temp	0.26

From **Table 5-8** it is indicated that the boiler temperature control loop has the largest influence on the performance of the plant. This has merit since the vapour flow in the column is crucial to the operation of the column.

The second most important control loop is Plate 10 temperature, which is controlled by adjusting the feed flow rate. The location of this loop is acceptable since the feed flow rate is important to achieve a product flow rate.

The third is the boiler level, also correctly situated since an increased level increases the boilup requirement and thus influencing the correct operation of the column.

The next is the distillate level, this is situated near the end which is correct since the only influence that this level has on the column is that it gives the supply head for the pump which pumps the reflux, and a change in head is unlikely to affect the flow rate of the reflux dramatically.

The last loop in the order of influence is Plate 1 temperature. This position appears a little strange at first glance, since the top plate temperature largely influences the product quality. Upon investigation of this it was found that a change of the reflux flow rate has very little effect on the top plate temperature and thus the ability of the controller to achieve the desired setpoint is hampered. This explains why the performance of this control loop would not influence the performance of the plant dramatically.

Table 5-9 gives the change in the single loop performance based on different methods as well as the plant performance chance seen for that specific single loop performance change, which is a result of a change in controller tuning parameters.

Table 5-9: Performance changes

Loop	PWI Change	GMV Change	MVI Change	ISE Change	IAE Change
Boiler Temp	5.89	2.94	3.60	-0.01	-0.06
Plate10 Temp	1.66	1.06	0.46	0.21	0.03
Boiler Level	1.50	11.75	9.70	13.71	13.88
Distillate Level	1.17	0.40	0.80	0.81	-0.53
Plate1 Temp	0.26	5.50	5.61	0.03	0.53

The generalised minimum variance (GMV) only has correlation to the plant in that increases in the loop performance are also seen as increases in the performance of the plant. This is also the case for the MVI.

If the difference between the change of the minimum variance index (MVI) and the change in the plant performance is observed, it can be seen that a large increase in the performance of a single loop does not dramatically increase the performance of the plant as can be seen in the boiler level and Plate1 Temperature from **Table 5-9**. The reverse is also seen in Plate10 Temperature.

Table 5-9 also looks at the integral of the absolute error (IAE) and the integral of the square error (ISE). These methods not only have no relation to the plant performance but also show a decrease in performance while the plant performance has actually shown an increase. This is the case since the loops with this specific phenomenon are affecting other more critical loops as well resulting in an improved plant performance despite the decreased single loop performance.

In summary it can be said that none of the single loop methods evaluated showed any relationship to that of the plant performance as this would be largely reliant on how the plant performance is defined and how interactive the loops are. The only valuable information obtained from this is that it is important to determine the plant-wide performance as it is a better indicator of improvements than just examining the effect of an improvement on a single loop.

The interaction between the loops also play a role and as seen in some case an increase in performance for one loop could result in a decreased performance for another loop, and if the loop with decreased performance is directly associated with controlling one of the main performance objectives for the plant a reduced PWI could be observed. These single loop improvements will however still result in a more stable plant but will not necessarily play a major role in driving the plant to its inherent optimum.

6 CONCLUSIONS AND RECOMMENDATIONS

The generic plant-wide performance monitoring tool successfully represents different installations. The best and worst PWI's for all the rigs tested are shown in **Table 6-1**. This PWI gives a good indication of how close the plant is to the specified optimum for each rig.

Table 6-1: Best and worst PWI's for all unit operations tested

	Large Distillation Column	Small Distillation Column	Level and Flow Rig	Temperature Rig
Worst PWI	64.36	71.31	67.05	68.17
Best PWI	74.06	87.20	75.59	72.70

In **Table 6-1** it can be seen that the best performing plant is the small distillation column. This can be the case since the most time was spent optimising the control for this plant.

The level and flow rig is the next best performing plant. This also makes sense since this plant operates satisfactorily and has also been optimised by previous operators of the rig.

The Temperature rig shows a lower performance than the large distillation column. This is debatable because the temperature rig has a very slow responding control loop while the large distillation column has one control loop that is still oscillatory. The order of these two is probably correct as the large distillation column has four other loops to improve the overall performance whereas the temperature rig relies solely on the efforts of one controller and if that performance is poor then the overall plant performance will follow suit.

This comparison of various plants based on the PWI is however limited and requires further comparisons to substantiate whether the PWI can accurately compare different plants.

The PWI is however still a useful tool to be able to assess the performance of a specific plant before and after specific changes like the addition of advanced process control (APC). The PWI could also serve to evaluate whether plant modifications that have been done actually achieve the desired improvement in the plant performance.

The combination of oscillation detection methods can successfully determine the cause of an oscillation - whether it is due to disturbances, stiction or incorrect tuning parameters. This requires the use of the frequency of the control error sign changes to determine if an oscillation is present. When detected, the cross correlation method is used to determine whether the oscillation is valve- or disturbance-related. If disturbance-related the necessary action can be taken to eliminate the effect of such disturbances and in so doing remove the oscillation.

If the source of the oscillation is determined as being valve-related, a further analysis is required in the form of an input-output pattern recognition method, which can then isolate whether the valve problem is stiction-based or based on incorrect tuning parameters and the necessary corrective actions can then be taken. The pattern for the incorrect tuning parameters has been confirmed as being a tightly wound circular pattern. This was illustrated on both the large distillation column and the level and flow rig.

From evaluation of the small distillation column data a relationship exists between plant performance and single loop controller performance in the form of an increase in the performance of a single loop does generally increase the performance of the plant. This is also verified in the other unit operations. It is however not a successful comparison as a large increase in loop performance does not necessarily result in a large increase in plant performance. This result is however largely dependent on the method that was used to determine both the single loop performance and the plant performance, as well as the interaction of the loops causes some more critical loops to perform worse while the loop being evaluated actually performed better. It is for this reason that it is important to look at overall plant performance and not just to sum the effects of the single loop controllers to estimate the performance of the plant.

This is also however an expected result as certain loops would have a larger influence on the plant performance if the variable which the loop is controlling is a variable which is directly used in the determination of the plant performance. Some other loops which are not directly used in the determination of the plant performance would only have a secondary effect on the plant performance and therefore improving that loop would result in large changes in the loops performance but only minor changes in the performance of the plant.

It is for this reason that the development of the plant-wide performance evaluation was required. Now it is possible to improve the plant performance by paying special attention to loops that directly affect the plant performance first, while the other non-critical loops take the back seat.

Once the main improvement in performance has been brought about the other control loops can also be improved to bring the overall performance of the plant up even further.

These loops that influence the plant performance most therefore have to be identified by changing the controller settings slightly and evaluating the change in plant performance. This then give the order of influence of the controllers and the controllers affecting the plant performance most can be tended to first and the not so important controllers can be evaluated once the major improvements to the plant performance have already been brought about.

The PWI is therefore capable of evaluation different unit operations successfully, the source of oscillation can successfully be detected and the various single loop assessment methods available have been compared and found that in all cases it is important to evaluate the PWI in conjunction with these single loop assessment methods to prevent a negative influence on the plant as a whole.

Possible future topics following from this investigation are:

- Make an allowance for a setpoint band and not be limited to a specific setpoint.
- The plant-wide performance monitoring method should now be tested on industrial data to see the applicability in industry; this should be done on various units, i.e. reactors, compressors, heat exchanges, cooling towers and distillation columns.
- The plant-wide performance monitoring method can further be adapted to become an on-line monitoring tool.
- The plant-wide index can be improved by making use of a geometric average instead of the current arithmetic mean.
- The oscillation detection procedure should also be tested on industrial data.

- This method can also be tested on advanced process control (APC) units, or serve as justification for APC on a specific unit.
- Investigate the possible change of weighting for each of the elements of the PWI. This change would result in an index that is more influenced by the changes of certain elements and less influenced by others. This could therefore make that changes in the single loop performance would be more directly visible in the PWI.

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APPENDIX A – PIPING AND INSTRUMENTATION DIAGRAMS
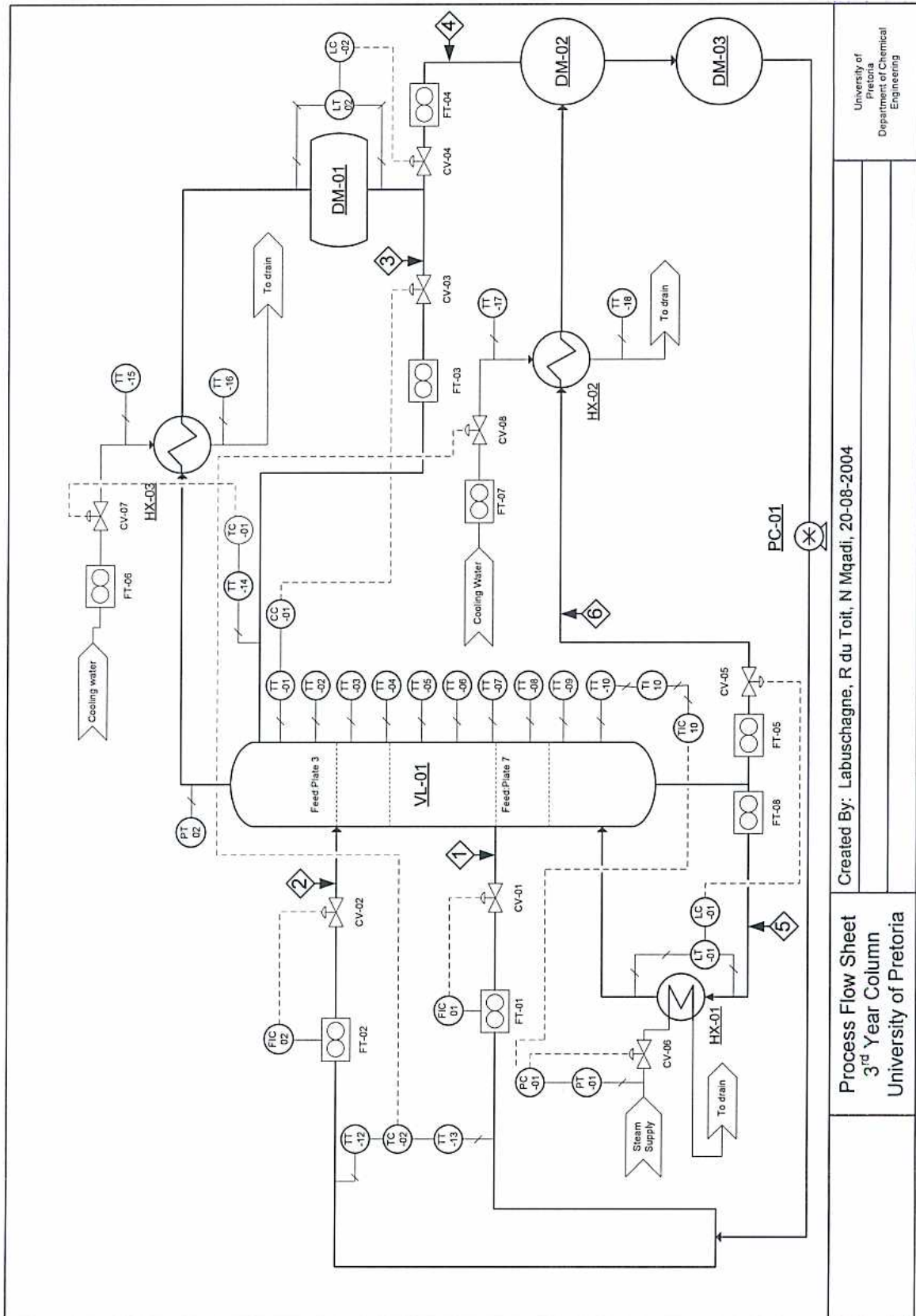
Figure A-1: P&ID - Large distillation column

Figure A-2: P&ID - Level and Flow
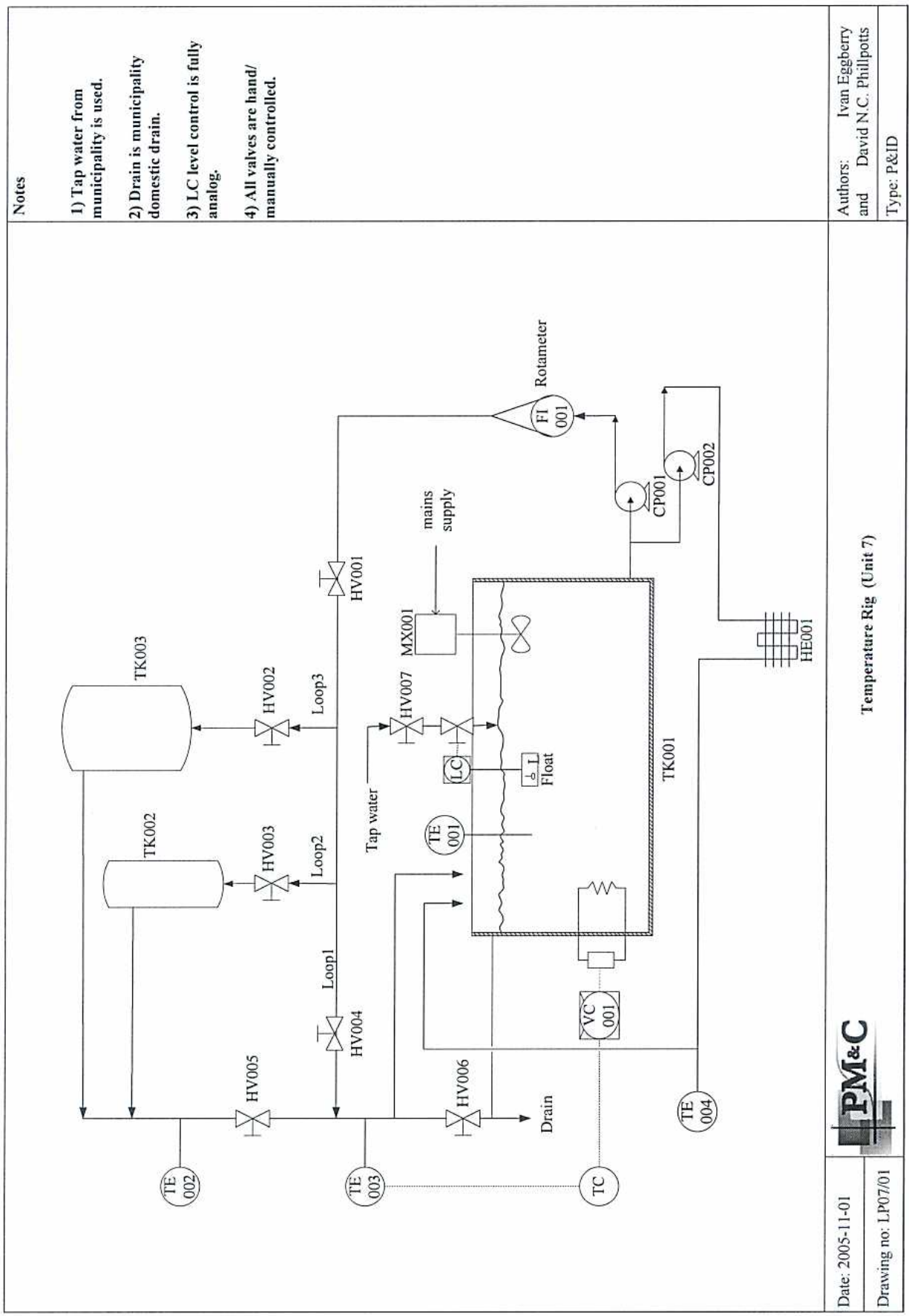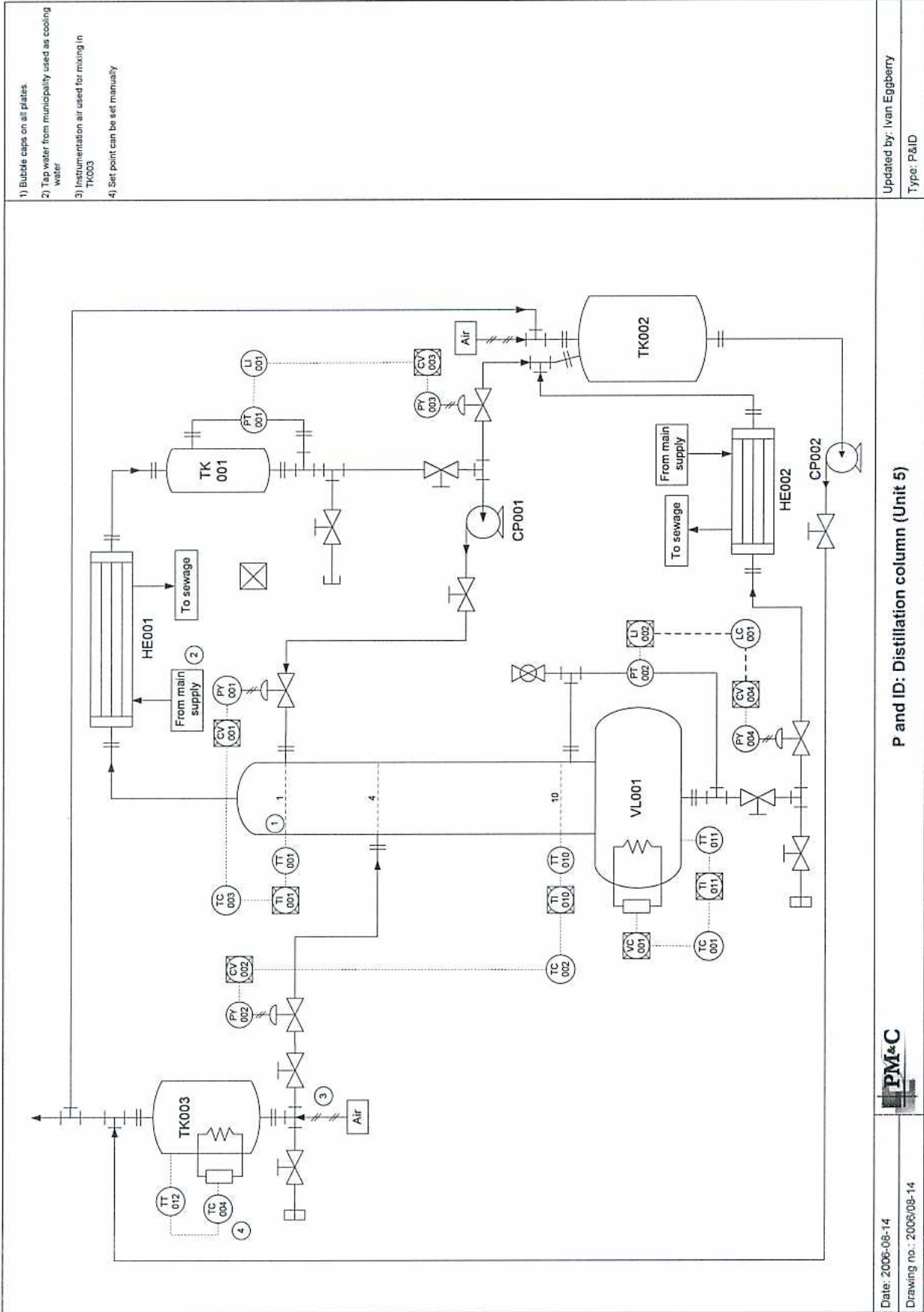
Figure A-3: P&ID - Temperature rig

Figure A-4: P&ID - Small distillation column

APPENDIX B – ACCOMPANYING FILES

*.olf -	Files with the .olf extension is operating data logged with the OPC Toolbox.
*.mat -	Files with the .mat extension is operating data logged for the various installations.
*.rpt -	Files required for the generation of the report.
*_Performance.m -	The specific performance calculation for each installation.
corr_coeff.m -	Function that computes the cross correlation coefficient for various lags.
Data_Dimen.m -	Function that reshapes any two vectors to be of equal dimensions.
Data_retrieve.m -	Function that imports the OPC log file (*.olf) into the Matlab current directory.
element_removal.m -	Function that removes specified elements from signal vectors.
Generalised_harris.m -	Function that computes the generalised minimum variance index.
harris.m -	Function that computes the minimum variance index according to Fellner's formula.
harrisLoc.m -	Function that computes the minimum variance over short intervals for plotting the localised minimum variance.
matloader.m -	The function used to load the appropriate data file of the various installations.
opcread.m -	Function that imports and creates the tag structure.
oscil.m -	Function that executes the frequency of control error sign changes algorithm for oscillation detection.

Plant_wide_GUI.m -	The M-file that launches the plant-wide performance interface.
plantwide_Generalised_harris.m -	Function that computes the generalised minimum variance index for all the loops on the plant.
plantwide_harris.m -	Function that computes the minimum variance index for all the loops on the plant.
Plots_Data.m -	Function that plots the relevant graphs for report generation.
PowerSpec.m -	Function that calculates the PSD.
PW_report.m -	The report generation file used to generate the plantwide report in html format.
refine_time.m -	Function that sets the evaluation period.
sigqual.m -	Function that sets the quality strings to a format that can be plotted.
Single_Loop_Assessment_GUI.m -	The M-file that launches the single loop performance interface.
Single_loop_report.m -	The report generation file used to generate the single loop performance report in html format.