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Innovation on Chemical Process Equipment for Advanced Yield &Positive Socio-Economic Impact (Distillation Column)

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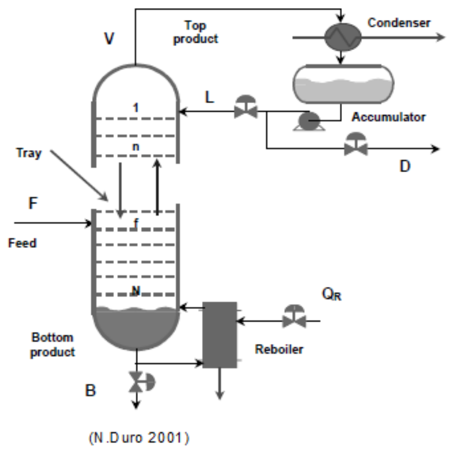
Abstract: In industries the most traditional separation technique and unit is distillation and distillation column respectively, which in the past have presented a lot of control difficulties, Mostly columns with high purity constraint exhibits "ill conditioning" problems and well-built interaction between the composition loops; it has been proven that perhaps improbability is the major reason we experience control problems. Devising robust control for distillation columns is a vital way to provide good composition control; this in fact corrects the "ill condition" problems. The proper handling of the strong interaction is achieved through a properly designed multivariable controller, making a distillation column a favourable unit to be applied a robust multi variable control technique.
MATLAB-Simulinks is one of the most used block diagram simulation software. The distillation column dynamics was modelled in simulinks environment, many simulations was conducted, observing the behaviour of the distillation column under various conditions. First condition is the observation of top composition without controller and with controller, the controller employed is a PID controller.
This brought us to controller tuning using the process reaction curve to read off the parameters for the estimation of the controller gain, integral time and derivative time.
The equation used is for the estimation of this parameter is the Cohen-Coon equation for open loop. Further simulations were carried out to test for stability and robustness. The same procedure mention above was applied to the bottom composition.
Afterwards the two loops were integrated together to give the [LV] configuration MIMO. Further observation was taken. With the perfect understandings of this control configuration and technique, a significant reduction in operation cost may be obtained.
Keywords: Innovation, Distillation Column, Advanced Yield

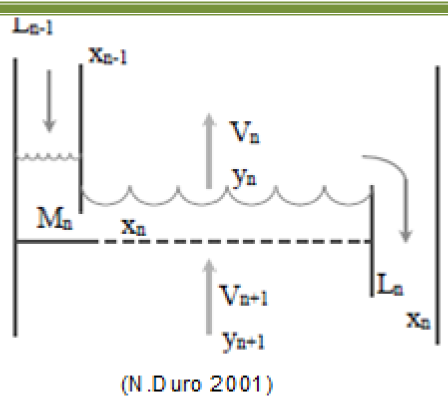
1. Introduction

In chemical process industry distillation column is the most paramount unit due to the fact that; it is the most effective separation technique/method with large throughput.

Distillation columns are complex, they have several components, which plays vital role, either in heat transfers, energy transfer or mass transfer. An ideal distillation column comprises of a vertical column packed with trays or plates, which ensures separation of components, re-boiler attached at the bottom which provides heat for vaporization, at the top of the a condenser, that cools and condenses the vapour exiting at the top of the column. The reflux drum is another important chamber; it contains the condensed vapour from where the liquid reflux is recycled back at the top of the column.



(Figure 1.1) Schematic diagram of a distillation column



(Figure1.2) Column tray

Modelling and control of a distillation column is a demanding task because of its intricate dynamics resulting from the integration of reaction and separation. Its actions are highly nonlinear owing to changing signs and path of the process gain. Control problems arise due to the complex interactions between vapour-liquid equilibrium (VLE), chemical kinetics, vapour-liquid mass transfer, and diffusion inside the particle catalyst and also time delays due to control flaws. (M.T. Vester*et al*1993).

Control of distillation column can be classified into two different aspectsSteady state control and dynamic control (J.A. Fabroa 2005)

The steady state control gives no room for disturbances in the control loop; this is commonly achieved by the use of industrial PID controllers. These PID controllers are designed to function within the specified operation limit obtained from a linear process model. Dynamic control becomes very necessary and important when there is alteration or changes in the operating conditions. In the case of such occurrence the best kind of controllers are predictive controllers. This is because they are used to generate set points for the basic PID controllers. Secondly when a column is operating far from its normal conditions the dynamic control strategies are applied. (J.A. Fabroa 2005)

1.1 Motivation (Objectives of Control System/Technology)

In an operation of a unit in chemical process targets are set to be met, these are imposed by the designers to operators and generally to all technical and non-technical staff of a given establishment. They are expected to meet with both social and economic conditions of operations in ever presence of changing disturbances. The requirement may include the following bellow;

1.2 Safety

Safe operation of any chemical process is always the first consideration basically because of the well-being of individual in the plant and secondly consistent income. Moreover the operational values of pressure temperature, concentration and many other related conditions must be within the stipulated or allowable limit. For example, if a distillation column is designed to operate at a pressure of about 100 psig, a control system should be provided to maintain the pressure below or on the designed value to avoid hazards which will lead to loss of life and properties during plant operation.

1.3 Product Specification

The desired quantity, quality should be produced. Some production facilities produce beyond their specified quantity and below specified quality, for instance a plant may require producing 5 million litters of ethylene per day and of 99% purity, it is the work of a control system installed to guarantee that the production level stays at the desired specification which 5 million litters and at 99% purity.

1.4 Environmental Regulation

This varies from region to region, but it remains a factor to be considered. Depending on the region, laws both federal and state have specific level stipulated for operations; this may include chemical concentrations, temperature, and flow rate from plants etc. For example the amount of SO_2 ejected into the atmosphere from a plant, the amount of emulsion in produced water returned to the sea/ocean.

1.5 Operational Constraints

Differentequipment used in chemical operation/process have constraints innate to their operation. These constraints are to be satisfied throughout the plant operation. For instance pumps must keep up to a certain net positive suction head, overflowing or drying of tanks should be avoided, preventing the flooding of distillation columns, the upper limit of temperature in a catalytic reactor should not be attained, for the catalyst not to be damaged. The only way to ensure that the above mentioned constraints in chemical process are maintained is by providing a control system.

1.6 Economics

Chemical process plants must be conventional with the market circumstances; this may include the accessibility and availability of raw materials and demand of the finished products. It should also conform as economical as probable in its exploitation of capital, human labour, raw materials and energy. Moreover it is essential that operating conditions of a given plant are controlled at given most favourable levels of least operating cost, utmost profit and so on.

The above mentioned requirements explained the necessity for a nonstop observation of the operation of a chemical plant and external intervention (control) to assure the contentment of operational goals. This is achieved through realistic arrangement of equipment which includes measuring devices, valves, controllers, computers etc. and human factors like designs (designers) and operators which makes up a control system setup. Once a control system meets up with; suppressing the effects of external disturbances, guarantee the stability of the chemical process, optimizing the performance of a chemical process, the control system is said to be satisfactory. (G.Stephanopoulous, 2003)

1.7 Research Objectives

Designing a control system involves recognition of control purpose; choice of appropriate measurement and manipulation as well as the determination of loops linking these; also identification of suitable control laws.

The main aim of this research work is to monitor various techniques and methods in controlling a distillation column; this process consists of various/multiple variables, some of which are controllable where as others are not. Analysis done will help to improve process performance over time by studying. Various techniques will be reviewed and discussed to provide basic knowledge of the numerous techniques applied in the chemical process industries to control distillation column. A practical example using MATLAB- Simulinks will be used to simulate and control a distillation column. Conclusion will be made from the observation recorded.

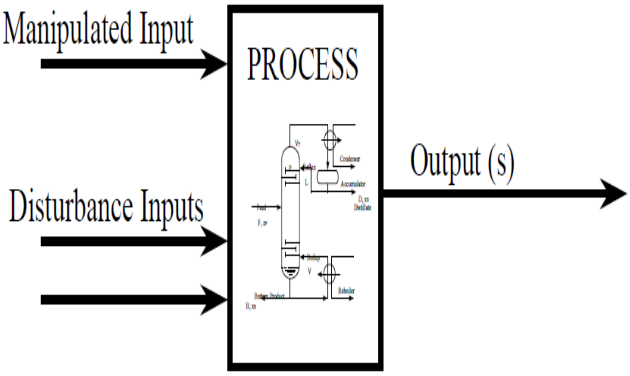
2 Review

2.1Short History on Control And Simulation

The pioneer principles of control system have long been in existence for centuries assisting humans in manipulating of tools and equipments efficiently. Evidently in the last century more work has been put to place to develop better control systems which are robust, complex and complicated. Massive development of control system over time has propelled the growth of process simulation from low to and irrelevant area to a very powerful and important tool and area in our daily life (W.Y.Svrcek et al, 2006).

2.1.2 Background*:*Basically a chemical process is defined by the diagram above, with many inputs and outputs.Controlled variable y in most systems is composition, temperature, level (liquid), pressure, flow. This could also be any other variable depending on the process to be controlled to give a certain fixed yield (value) stipulated as the set point (y_{sp}). The manipulated variables will be changed /adjusted to bring the process back to the set point.

2.1.3 Control system: A process comprises of many variables (units), these includes manipulated input (*u*) in most cases they are manipulated. Disturbance (*D*) also known as load input, output (*y*) which is the controlled variable



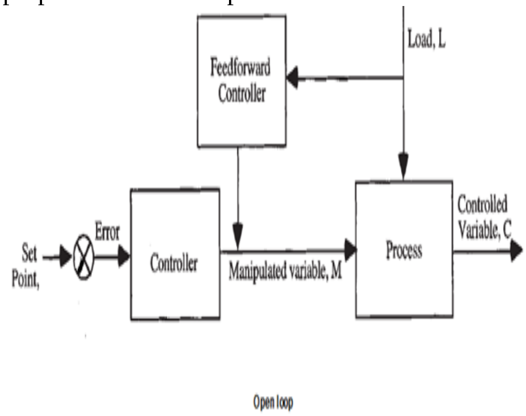
(Figure 2.1) schematics of a Process system

2.1.4 Disturbances: This are variables that change deviating the process output from required operating value (setpoint), this could come in form of changes in ambient temperature, environmental conditions, flow, pressure etc. This have can be sub-classed into measured and unmeasured signals.

2.1.5 Manipulated variables (mv): in a control system this variables are selected and adjusted to affect the process output variables. The outputs are referred to as **controlled variables** due to the fact input variables are adjusted to make a difference (control).

2.2 Open Loop System

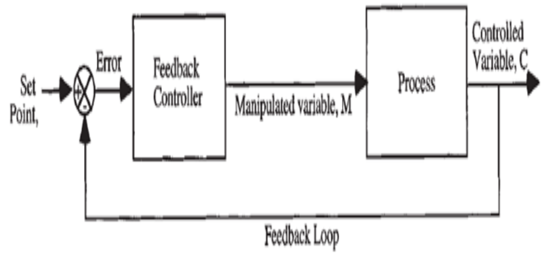
In this system the manipulated variables are placed either manually or programmed without the aid of measurement obtained from the process, this is always attainable from processes without disturbances. Here automatic control switches are made available if need be for manual alteration of the manipulated variable if the control system's performance is not acceptable. These systems make use of the input and produce an output through the action of the plant. The feature of the internal workings of the plant is vague and to some extent unimportant. The important factor here is the relationship between the input and the output. Good open loop systems have a value of output proportional to the input.



(Figure2.2)

2.3 Closed Loop

This is more advanced than the open loop because this configuration makes use of measured variables to adjust the manipulated variable to attain control objectives, meaning that the system utilizes feedback which is gotten from the output. Feedback and Feed-forward loop systems are both closed loop



Closed loop

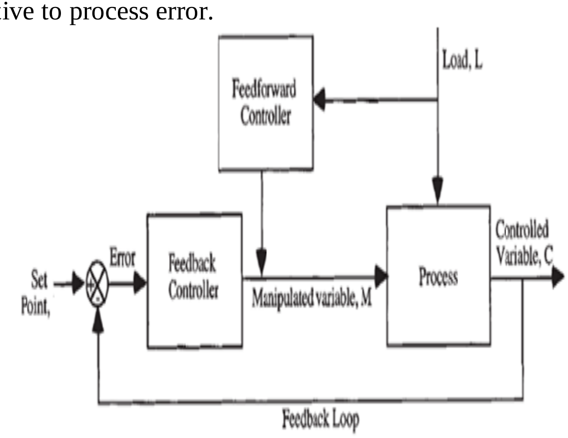
(Figure2.3)

2.4 Feed-Forward Control Loop

The resulting deviation in this process is minimized by the use of measured disturbances variables by proper positioning of the setpoint. Feed-forward works better and efficiently with the integration/combination of feedback controllers, it reduces the offset to the nearest minimum, these offsets occurs due to inaccurate calculation and measurement of the load component. (J.P. Shunta, 1985).

Advantages of a Feed-forward control loop include;
1) Takes action before the effect of a disturbance been observed by the system
2) Works well for a system with big dead-time.
3) The response does not introduce instability in a closed-loop system.

The disadvantage includes;
1) Needs measurement of all feasible disturbances directly.
2) Finds it hard to handle unmeasured disturbances.
3) They are very sensitive to process error.



Feedforward with feedback

(D.W. Green, R.H. Perry 2008)

(Figure 2.4)

2.5 Feedback Control Loop

Here the set point (*Ysp*) is in direct comparative with the controlled variable (u), which is moved in such a way to minimise error (*E*), as a result of the action of the controller on the error. The feedback control loop invocations can be traced back to more than 2000 tears back, but the first documented industrial application would be James Watt's application of the loop to control the speed of steam engines about 200 years back. (C.A. Smith, 2006). Feedback controllers have so many advantages which includes;
1) Corrective actions taken do not need identifying of the disturbance in the process and do not need taking measurements too.
2) The process model must not be explicit.
3) Basically they are robust and control the process with less error.

The disadvantages include;

1) In most closed loops, they cause instability in their response.
2) Process with large time constant and recurrent disturbance are not properly handled by the system
3) The system has to feel the effect of the process disturbance before a controller action will be taken.

For a feedback system consideration of two first-order process transfer function are taken. (B.W Bequette 1998);

$G_1(s)=\frac{K_1}{T_1s+1}$ (2.1)

$G(s)=\frac{K_2}{T_2s+1}$ (2.2)

$G_{c1}s=\frac{G_1s}{1-G_1(s)G_2(s)}=\frac{\frac{k_1}{T_1s+1}}{(\frac{k_1}{T_1s+1})\frac{K_2}{T_2s+1})}$(2.3)

$$=\frac{K_1(T_2s+1)}{(T_1s+1)(T_2s+1)-k_1k_2}$$

$G_{c1}(s)=\frac{K_1(T_2s+1)}{(T_1T_2s^2+(T_1+T_2)s+1-k_1k_2}$(2.4)

$G_{c1}(s)$ Is stable when the roots of $T_1T_2S^2(T_1+T_2)s+1-k_1$ are stable

$G_1(s)=\frac{2}{5s+1}$(2.5)

$G_2(s)=\frac{k_2}{10s+1}$(2.6)

Since K_1= *2*, then K_2 must be less than 0.5 for stability.

As numerical check, let k_2= -1. Then solve for the roots of:

$T_1T_2S^2+(T_1+T_2)S+1-k_1k_2=0$ (2.7)

We find

$50s^2+15s+3=0$ (2.8)

(Equations from B.W Bequette 1998) page 274.

Here are some applications of feedback control system in chemical process industries and annotation used to represent them; pressure control (PC), flow control (FC), composition control (CC), temperature control (TC), level control (LC).

2.6 Control Structures

In chemical process industries the widely used control structure is the **PID** controller which stands for Proportional (based on present error), Integral (based on passed error), and Derivative (based on the future error). This is a controller with three elements, each of the above mentioned element have its effect in a system. In the application of the PID controllers, a good knowledge of the processes is required so as to know the action of the controller to be implemented in the process to attain a desired control. (M. Araki and H. Taguchi 2003).

2.7 Control Techniques

In various literatures there are many control techniques applied in process control. In the study of process dynamics the process variables and control signals are function of time, the best way to represent it is the Laplace transform of a function of time, f (t). It is defined by the formula below

$F(s)\ =\ L\ [f\ (t)]\ =\ \int_0^\infty f\ (t)\ e-stdt$(2.9)

The domain function in the equation can be transformed into an algebraic problem as represented by the function G(s). (P.C. Chau 2002).

2.7.1 Proportional (P): This is based on the present error of the process, various literatures have a various equations/mathematical expressions used in representing this action**,** and Willis (1999) represents the action as

$\frac{mv(s)}{e(s)}=k_c(Laplacedomain)ofmv(t)=\ mv_{ss}+k_ce(t)(timedomain)$ (2.10)

The proportional control mode regulates the output signal in direct proportion to the controller input (error) e. The changeable variable/parameter will be the controller gain (k_c). Most times it is been confused with the process gain (k_p). The bigger the value of $k_c$an elevated yield of output error will be expected. Example a controller gain of 1 yielding 10% error, by range will adjust the controller output by 10% of scale. Most process instrument producers use Proportional Band (PB) instead of k_c. (M.J. Willis 1999). The indication of the time domain in the equation is that the control system needs alteration around the steady state operating point. In equation 2.10 an expression (mv_{ss}) represents the steady state signal for the manipulated variable; this ensures that at zero error the controlled variable is at the setpoint. The term vanishes at the Laplace domain due to the deviation variable representation. At the tail end of it all, a proportional controller only minimises error but do not completely remove it except the process has an integrating capabilities. There is always offset between the actual value and the desired value (setpoint). (J.Chandrasekaret al.2007).

2.7.2 Proportional Integral (PI): Various chemical processes do not favour being controlled with an offset. Elevated level of intelligence is added so that the process will be controlled. The added parameter in this case is the integral mode. The integral action automatically takes a corrective action by removing offsets (error) that arise between the set-point and process output within space of time. This time the integral time (Ti) is the specified modifiable parameter. (N.A Rahman, I.R.A Manap 2003). The PI is represented mathematically as;

$\frac{mv(s)}{e(s)} = kc[1 + \frac{1}{T1s}]\ or mv(t) = mv_{ss} + kc[e(t) + \frac{1}{T1}\int e(t)dt]$ ... (2.11)

The PI is also widely referred to as reset, this reset is used to illustrate the time it takes for the integral action to make same alteration in manipulated variable as the Proportional modes primary change. (Y.S. Ryu and R.W. Longman, 1994).

2.7.3 Proportional Integral Derivative (PID): Derivative action predicts where the process dimensions, with the aid of time rate change of the controlled variable that is the derivative. Derivative time (T_D) or rate time is the main characteristics of the derivative action. Derivative time (T_D) units are in minutes. Mathematically the PID algorithm is represented below;

$\frac{mv(s)}{e(s)} = kc\left[1 + \frac{1}{T1s} TDs\right]$ or $mv(t) = mv_{ss} + kc[e(t) + \frac{1}{Ti}\int e(t)dt + TD\frac{de(t)}{dt}\ ]$ (2.12)

Unlike P and I, Derivative action depends on gradient the error, but if the error is constant the derivate action will be of no effect. (M.H Moradi 2003).

2.8 Controller Tuning

For an effective control of a system the parameter of a controller parameter have to be adjusted to drive the process to a desired setpoint. Controller tuning is simply is selecting the best values of k_c,Ti,T_D. There are many methods used in controllers tuning in process industries. According to "Luyben" (1997) page 92, 80 percent of Loops are tuned experimentally by an instrument mechanic, with the result accuracy up to 75 percent. This is because the mechanism uses the past experience of similar loop in the system.

2.8.1 Controller tuning methods: There are many methods used in controller tuning, the value of the tuning parameter is a function of the required closed loop response and on the dynamic quality or behaviour of the other element of control loop, mainly the process. (C.A. Smith 2006).

2.8.2 Rules of thumb: The basic kind of control loops are pressure, level, flow and temperature. In most cases the category of controller and setting are same when applied to the above mentioned loops.

Flow loops: PI controllers are the most commonly used in flow loops. High value of proportional band and low gain is used to trim down the outcome of the noisy flow signal due to flow instability. A small value of integral time/reset time (Ti) is used to get quick setpoint tracking. The process dynamics are usually very rapid; the sensors observe the change in flow almost instantly, but the dynamics of the control valve happens to be the slowest the loop, so therefore a low value of reset time can be used.

Level loops: So much interest is not taken from where the liquid level is in most process, as long as it stays between minimum and maximum values.

This makes it most suitable for the proportion controllers to be used in controlling liquid level to give even changes in flow rate and sort out variations in flow rates to downstream.

Pressure loop: Pressure control loop varies from very tight to fast loop similar to flow control to slow averaging loop. (Youzhe Ji.et at 2009)

Temperature loops: Temperature control loops are frequently quite slow because of the sensor lag and the process heat transfer lags. The most suitable controller for these loops generally is the PID controllers. Proportional band settings are reasonably minute, which is as a reference to the temperature transmitter and control valve sizes. When the process is fast, the smaller value of Ti can be laid down. The noise in the transmitter determines how the derivative time is set. Most times the T_D is set to one-fourth the process time constant.

2.8.3 Online Trial and Error: This method of controller tuning takes a great deal of time to get the appropriate value for the parameters k_c, Ti, T_D of a controller. According to "Luyben" 1997, to tune a controller on-line, a good instrument mechanic follows the route or something similar to the following:

i) The controller on a manual mode, the integral and derivative action should be eliminated from the controller, that is; set Ti to the highest value (minutes) and T_D at the lowest minutes.
ii) Set of k_c at a small value, maybe 0.2.
iii) Set the controller to automatic mood.
iv) Stipulate low setpoint or load change and monitor the response of the controlled variable. The response is sluggish due to low gain.
v) k_c should be increased by a factor of 2, afterwards make a minute change in the setpoint or load.
vi) Further increase should be implemented on k_c, this fifth step should be repeated over till the loop becomes very under-damped and oscillatory. When this occurs the gain observed is called the ultimate gain.
vii) Back off k_c to half the ultimate gain value.
viii) At this point, integral action should be brought in by dropping of Ti by factor of 2 making minute disturbances at each value of Ti to see effect.
ix) Locate the value of that of Ti makes the loop very under-damped, and set Ti at double the value.
x) Commence the introduction of the derivative action by increasing the T_D. Load changes ought to be used to agitate the system, and the derivative should take action on the process variable signal. Locate the value of T_D that gives the tightest control without amplifying the noise in the process variable signal.
xi) Raise the value of k_c over again by pace of 10 percent until the desired specification on damping coefficient or overshoot is satisfied.

The above listed procedure does not work in all the loops. The process is unbalanced at high and low values of controller gain, but is so stable over some in-between range values of gains. The above mentioned process will be applied in this work in comparison of other methods of controller tuning.

2.8.4 **Ziegler-Nicholas Method (ZN):** This has been a very important method of tuning a PID controller, the rules however has severe shortcoming, they utilize deficient process information and the design norm gives closed loop systems with reduced robustness. In the year 1942, J.G Ziegler and N.B Nicholas in proposed two methods which includes; a step response method and a frequency response method. (K.J. Astrom, T. Hagglund 2004). They are mainly used as first assumption but they are likely to be too under-damped for most process control purposes. The ZN is a good way to start controller tuning, this method involves at the initial, find the ultimate gain (k_u). When the loop attains the limit of stability with proportional feedback controller only, the period of the consequential oscillation is called ultimate period (P_u). The unit is min/cycle. With the k_u and P_u the ZN settings are calculated by the equation below;

P (k_c) $= \frac{ku}{2}$ (2.13)

PI (Ti) $= \frac{Ku}{\mathbf{2.2}} \frac{Pu}{\mathbf{1.2}}$ (2.14)

PID (T_D) $= \frac{\boldsymbol{Ku}}{\mathbf{1.7}} \frac{\boldsymbol{Pu}}{\mathbf{2}} \frac{\boldsymbol{Pu}}{\mathbf{8}}$ (2.15)

Equation from, (W.L. Luyben. M.L. Luyben 1997). Page 97.

2.8.5 Tyreus-Luyben Method: This method is related to the ZN method but gives more conservative settings making it more suitable for chemical process control use. (W.L. Luyben. M.L. Luyben 1997). The method makes use of ultimate gain Ku and ultimate frequency ώu

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PI (Ti) $= \frac{Ku}{\mathbf{3.2}}$ $2.2\ Pu$ (2.16)

PID (T_D) $= \frac{\boldsymbol{Ku}}{\mathbf{2.2}} 2.2\ Pu \frac{Pu}{6.3}$ (2.17)

2.9 Advanced control Systems

Over the years more research and development of more efficient control system that can give greater through put in processes. The basic feedback control system is the foundation of the more complex systems applied in industries today. These advanced control systems includes;

Advanced model predictive controllers(MPC), Robust multivariable predictive control, Cascade control, Time delay compensation, selective and override control, Non-linear control, inferential control, Adaptive control, generalized predictive controllers (GPC), Pneumatic controllers, electronic digital controllers, Distributed control system (DCS), Programmable logic controller, Field bus Controller. Some of the above mention control techniques are applied to distillation column while some are not. For more information see (D.W. Green, R.H. Perry 2008).

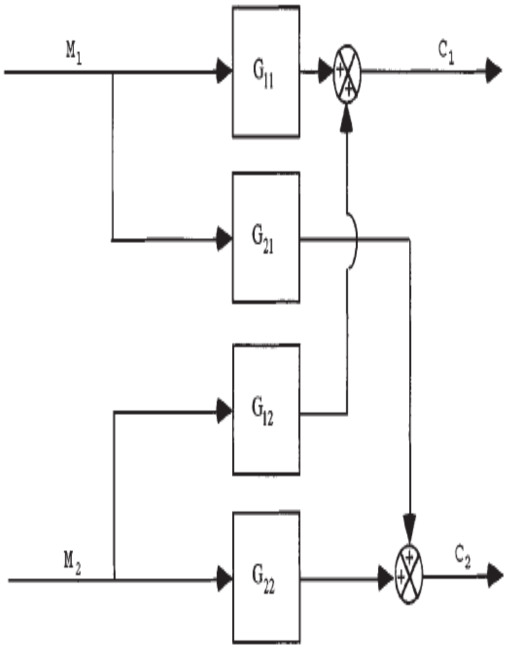
3 Methodology

3.1 Simulation/Control (Distillation Column)

3.1.1 Degree of freedom:before simulating a distillation column a set of specifications needs to be provided. To do this, one needs to understand the concept of degree of freedom. This is defined as; degree of Freedom= Number of unknowns – Number of equations. The degree of freedom helps in the specifying of the control objectives of any chemical process. An ideal binary distillation column comprises of six degrees of freedom but due to the two main (external) disturbances Feed rate (F) and composition (zF). The values of these two variables are specified by external factors. (G.Stephanopolous. 2003).

3.1.2 Specification of control objectives: With the elimination of two degrees freedom, we are only left with four control objectives in the simulation of ideal binary distillation column. Top composition of distillate stream (xD), Bottom composition (xB), Condenser holdup (mD) and re-boiler holdup (mB), the above mentioned variables are required to kept at a required value.

3.1.3 Control configurations: There are many configurations used in the control of a distillation column which includes [LV], [DV], [LB], [DB] and [L/D,V/B]. The control property of various configurations may be considerably different; basically the study of the relative gain array (RGA) which have to do with the study of steady state two way interaction expressed it and gave a basic insight to the gain of system loops . The RGA describes how the gain of loops changes when a relative loop is closed. The configuration to be analysed is the [LV] configuration this is done with reflux L and boil-up V as independent variables for the control of top composition yD and bottom composition xB. They are adjusted to obtain good level control. (S.Skogestad1997). The two inputs affects the two outputs, so proper pairing of the manipulated and controlled variable is required to achieve a good multiple input multiple out control, this process is referred to as 2 x 2 system, signifying the number of inputs and outputs. (M. El-Fandi, et al 1998). The uses of decoupling to remove the unwanted interaction between loops have been in place in the earliest work in multivariable control. (Figure 3.1) below is a structure of decoupling.



Example of 2 × 2 transfer function.

(Figure 3.1) Decoupler for MIMO

3.2 Distillation Column Design And Operation

The model of distillation column used in this research multiple inputs/ multiple outputs (MIMO), this was designed based on the study by Skogestad and Morari

3.2.1 Assumptions: Binary mixture, constant pressure, constant relative volatility, constant molar flows, no vapour holdup, linear liquid dynamics, equilibrium on all stages, total condenser. The column is made up 40 theoretical stages,

Data

Feed rate F = 1.01 [kmol/min]

Feed composition z_F = 0.5 [mole fraction units]

Feed liquid fraction q_F = 1 [for saturated liquid]

Reflux flow L_T = 2.706 [kmol/min]

Boil-up V = 3.206 [kmol/min]

Parameters/ Notations

L_i-- Liquid flow from stage i (kmol/min).

V_i – Vapour flow from stage i (kmol/min)

X_i-- Liquid composition of light components on stage i (mol fraction)

y_i– Vapour composition of light components on stage i (mol fraction)

M_i -- Liquid holdup on stage i (kmol)

D/B – Distillate top/bottom product flow-rate (kmol/min).

L = LT – reflux flow (kmol/min).

V = VB – boil-up flow (kmol/min).

F = Feed rate (kmol/min).

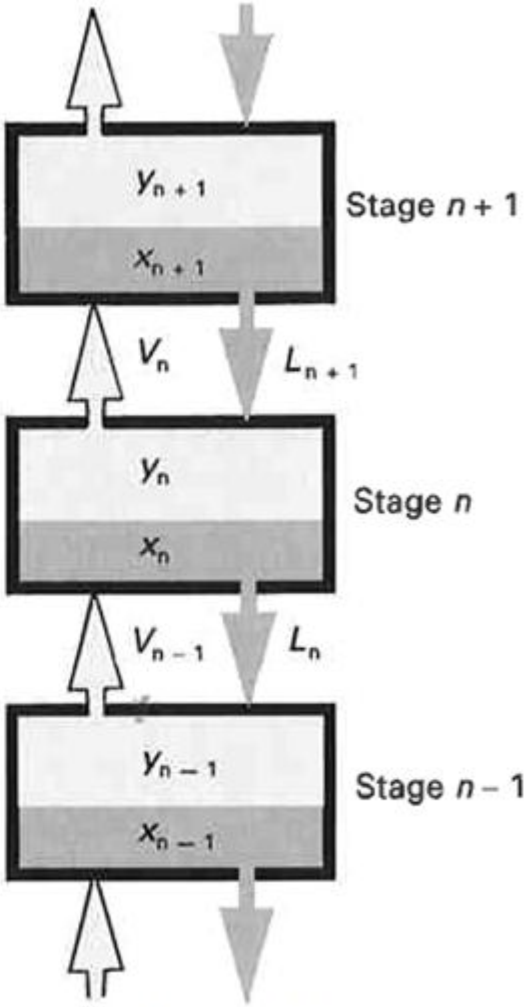
z_F – feed composition (mole fraction)

i – Stage no [1= bottom NF = feed stage, NT = Total condenser]

Alpha – relative volatility (light and heavy components).

Taul - time constant [min] for liquid flow dynamics on each stage

lambda - constant for effect of vapour flow on liquid flow ("K2-effect")



(I.J. Halvorsen, S.Skogestad 2000)

(Figure 3.2) internal schematic of a working column tray by tray

Material balance (stage i); $\frac{dM_i}{dt} = L_{i+1} - L_i + V_{i+1} - V_i =$

Material balance (light component) stage $\frac{dM_i x_i}{dt} = L_{i+1}x_{i+1} \ V_{i-1}y_{i-1}L_i x_i - V_i y_i$

The above equation yields; $\frac{dx_i}{dt} = \frac{dM_i x_i}{dt - x_i}\frac{dM_i}{dt}/M_i$

Derivative liquid mole fraction.

The vapour composition y_i is simultaneous to the composition x_i on the same stage through. The algebraic vapour liquid –equilibrium (VLE) becomes;

y_i = alpha x_i/(1+(alpha-1)x_i). Alpha is the relative volatility. The assumptions' constant molar flows and no vapour dynamics yields; $V_{NF-1} + (1 - qF)V_i = v_{i-1}$

Not for feed stage (except partly vaporized feed stage).

Overall, the flow of liquids depends on the hold-up on stage above, the vapour flow as follows. This is a lineralized relationship. Francis Weir's formula is an alternative used in this scenario.

$$L_i = L0_i + \frac{M_i - M0_i}{taul + (V - V0)_{i-1}} xlambda$$

.

$L0_i$ (kmol/min) is the nominal values for liquid flow

$M0_i$ (Kmol) is the hold-up on stage i.

Vapour flow into the stage also affects the hold-up, lambda may be positive owing to more vapour bringing rise to bubbles which may press on till the liquid leaves the stage. When lambda is at big value (> 0.5) this leads to the flattening out of the re-boiler hold out for a while in direct response to an increase in boil-up, on the other hand if lambda (>1) there will be an inverse response (Skogestad and Morari 1988).

It is generally difficult to estimate lambda for tray columns because sometimes lambda is negative due to pressure drop caused by larger boil-up resulting to hold up in down comers. Lambda is close to (0) in packed columns. The equation above can be applied to all the stages except in the condenser, feed stage and re-boiler.

Feed stage; i = NF it is assumed that the feed is mixed directly into the liquid at the

Feed stage; $\frac{dM_i}{dt} = L_{i+1} - L_i + V_{i-1} - V_i + F$

$$\frac{dM_i}{dt} = L_{i+1}x_{i+1} + V_{i-1}y_{i-1} - L_iX_iV_iy_i + z\_F$$

Total condenser; $= NT\ (M_{NT} = M_D L_{NT} = L_T\ )$

$$\frac{dM_i}{dt} = V_{i-1} - L_i - D$$

$$\frac{dM_ix_i}{dt} = V_{i-1}y_{i-1}L_ix_i - Dx_i$$

Re-boiler; $i = 1(M_i = M_BV_i = V_B = V)$

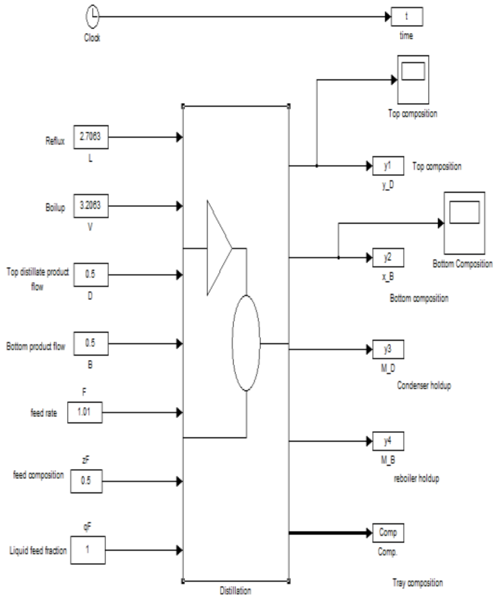
$$\frac{dM_i}{dt} = L_{i+1} - V_i - B$$

$$\frac{dM_ix_i}{dt} = L_{i+1} - V_iy_i - Bx_i$$

(I.J Halvorsen and S.Skogestad 2000)

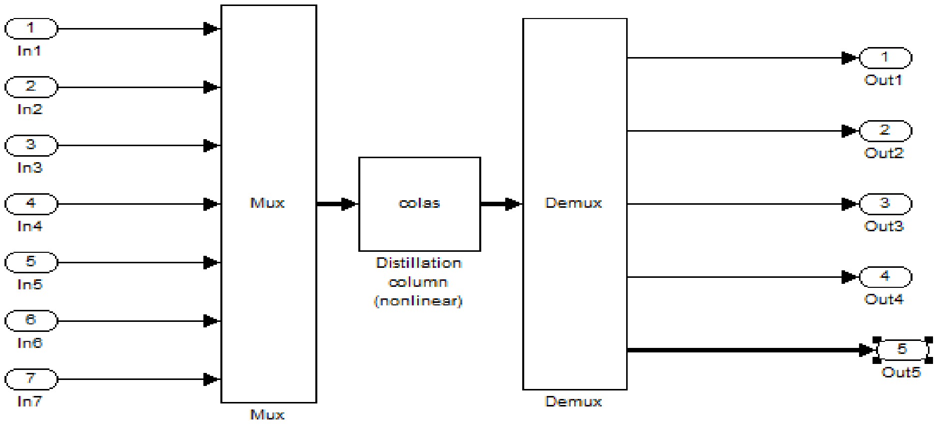
3.3 Building the Simulation

The simulation was built using various blocks in Simulinks (Figure 3.3). Blocks are dragged and dropped from Simulinks block library. On the left hand side are constant block that represents the inputs of the distillation column, the values assumed for the simulation are specified in them. On the left are the outputs they are specified with to workspace blocks in the block library. Attached to output one and two is the scope block (graph) that plots the output response. The main column is a subsystem that contains the blocks in (Figure



(Figure 3.3) Block diagram of distillation column simulation

Below is (Figure 3.4), it is made up of three blocks, mux, s-function and demux. The s-function block "colas" contains MATLAB files/function by (S.Skogestad 1997). The mux block takes in the input while demux leads to the output.

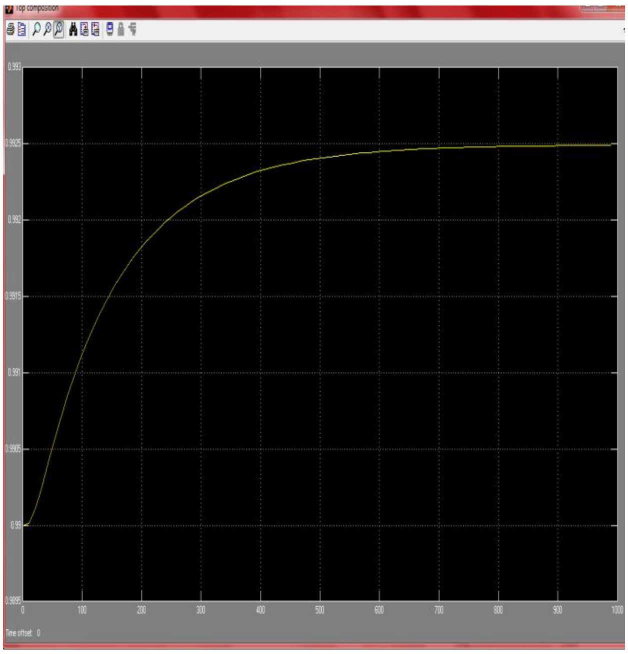


(Figure 3.4) Internal diagram of the simulation blocks

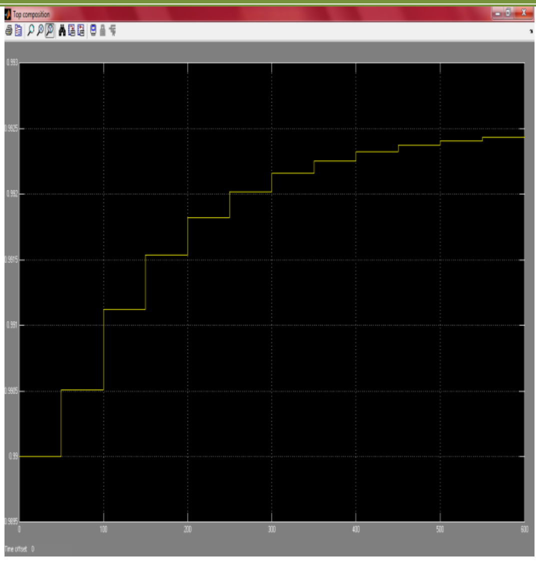
4 Results

4.1 Discussion

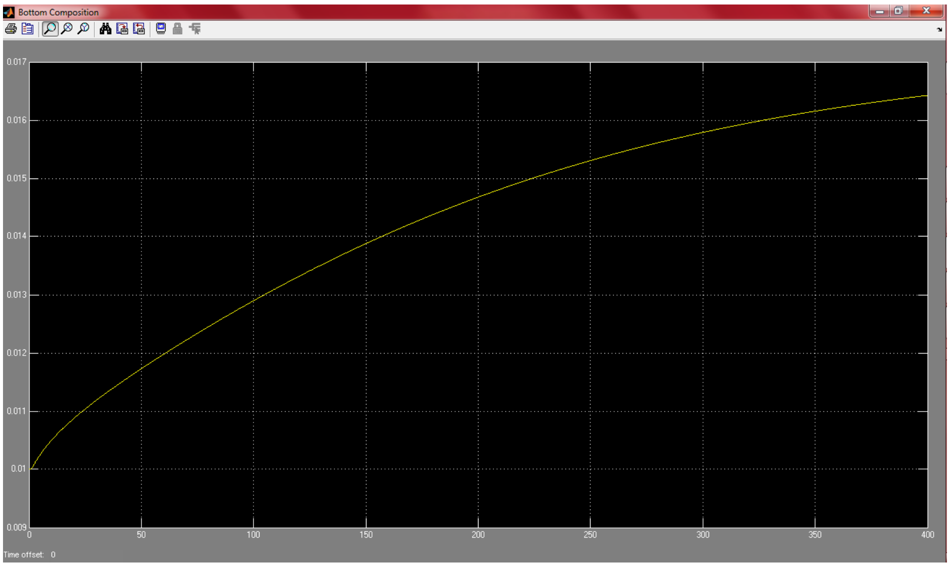
The graphs below are the representation of the simulation specifying the output.
On Y axis is composition (kmol/min).
On X axis is time in (seconds).
This stipulated above is applicable to all the graphs.
The column is simulated for one thousand seconds a linear progressive response is giving for top composition yD (Figure 4.1 and 4.2) and bottom composition xB (Figure 4.3). The distillate (yD) is at 0.99(99%) purity. A test was conducted with the increasing of the value of feed rate (F) from 1.01 to 1.4 [kmol/min] with block parameter step (figure 4.4); this signifies the effect of external disturbance as stipulated earlier. (Figure 4.5) illustrates effect of elevated disturbance to a process, in our case a distillation column.



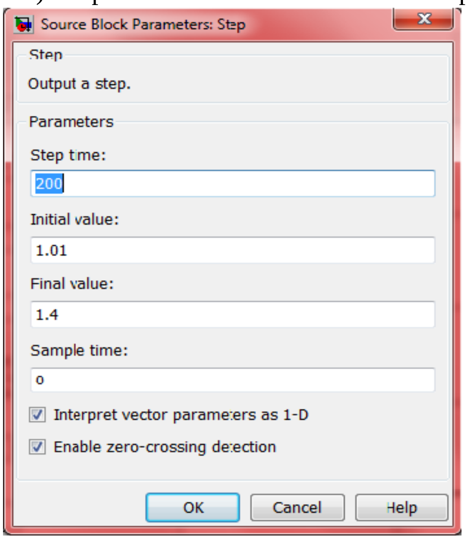
(Figure 4.1)Response of uncontrolled distillation column top composition



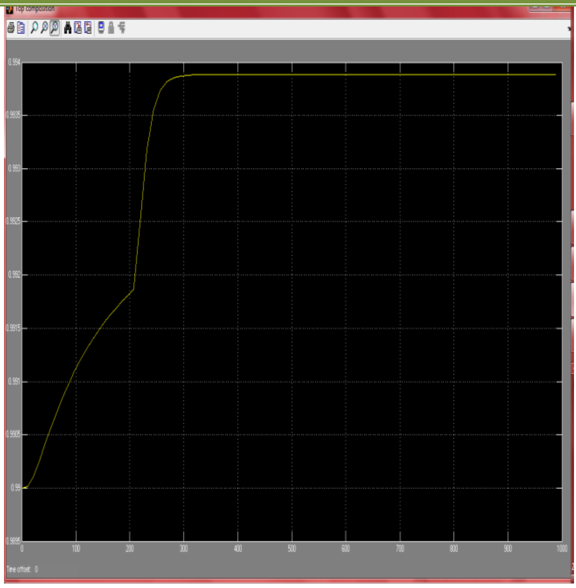
(Figure 4.2) Response of uncontrolled distillation showing step by step response



(Figure 4.3) Response of uncontrolled bottom composition

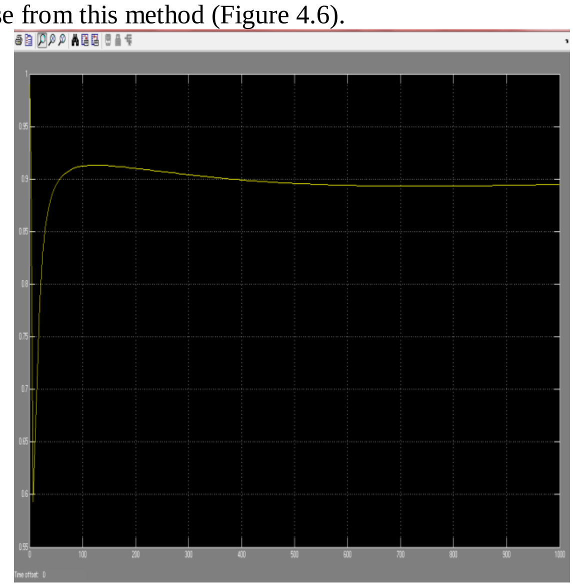


(Figure 4.4) Step block, source of disturbance to the column.



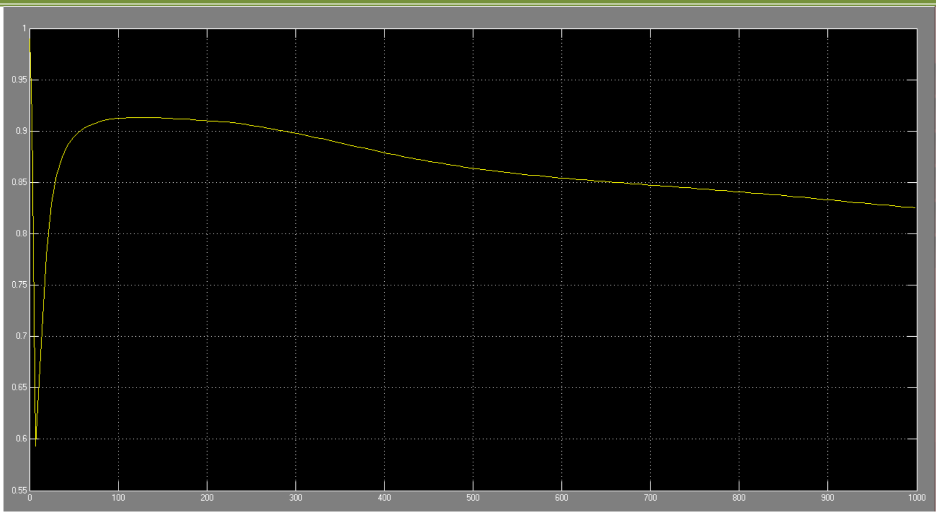
(Figure 4.5) Response of top comp (yD) with elevated level of disturbance.

The controller used (PID) is found in continuous blocks in simulinks block library.
The design/ implementation of a controller to this system; on-line trial and error method as described in chapter two was tried out to generate parameter for the controller, values *KcTi Td*was selected and simulated for 1000 seconds, below is the response from this method (Figure 4.6).



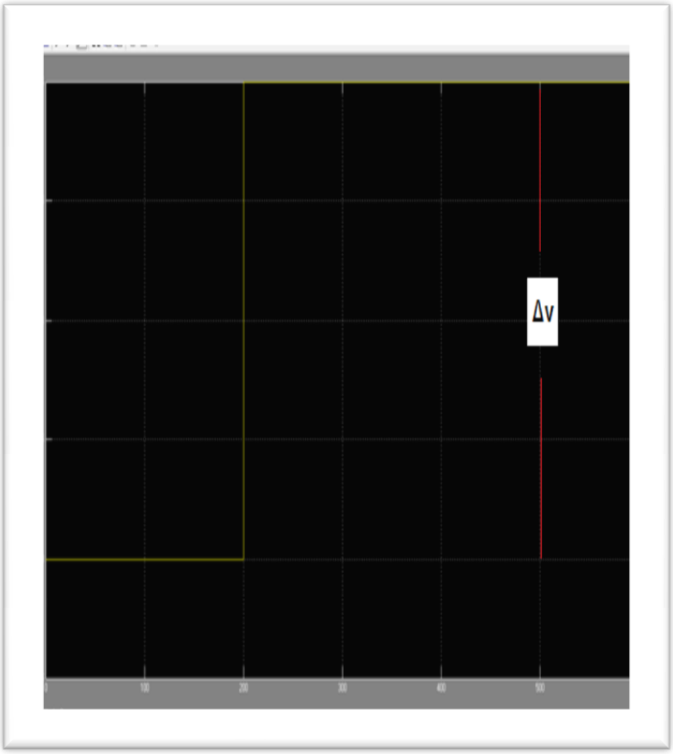
(Figure 4.6) Response of online trial error method tuning

The responsons seems alright but test for robustness and stability was conducted; this was done by altering the set-point from the initial value 0.9 to 0.85. The step change was implemented at time 400 secs. Below (Figure 4.7)

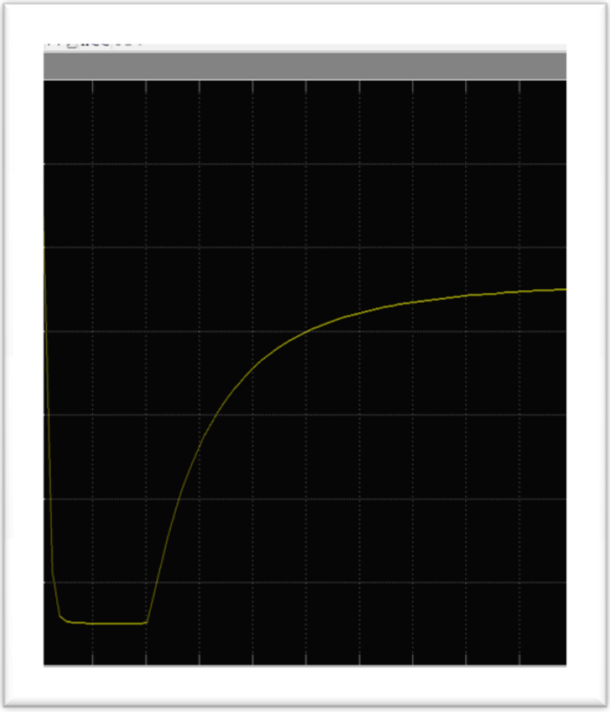


(Figure 4.7) Response of unstable control using online trial and error method

To get more robust control of throughput, a more advanced technique is applied “process reaction curve” process step testing for open loop. With the application of this, process gain (*Kp)* dead time (*T*o*)*, Time constant*(tθ)*are obtained from the reaction curve generated, these parameters are then used for the estimation of controller tuning parameters; controller gain (*Kc)* integral time *(Ti)* derivative time *(Td)*. This is done with the process set to manual mode (without a controller), a significant step change is impacted on the process to observe a change but the step should not be too large for it not to distort the process non-linearity’s, then after the changes can be measured as described in chapter two. (Figure 4.8) plot of step change (ΔV) initial value of reflux is 2.3kmol/min and final value of reflux is 2.5 kmol/min; which took effect by 200 seconds into the simulation. The response in (Figure 4.9) is a plot of the output (yD) “process reaction curve”. The plot covers the entire time of simulation from the introduction of the step until the system reaches its new steady state. This may last between few minutes and several hours. It took the top composition 600 seconds to attain steady state.

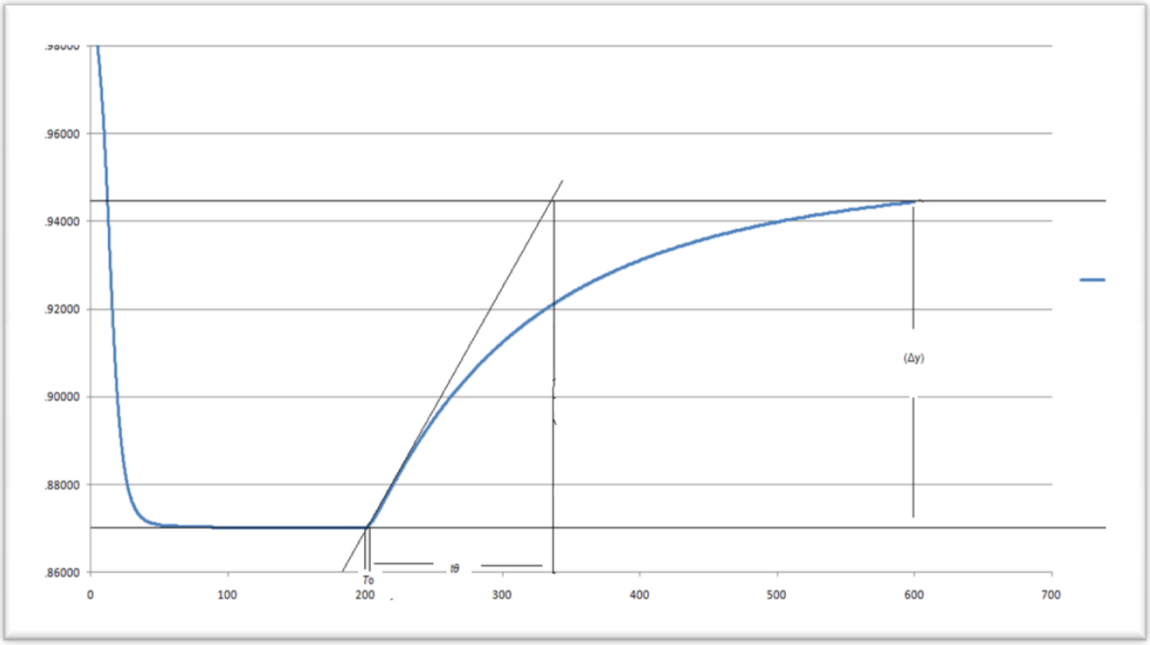


(Figure 4.8) Response to step change



(Figure 4.9) Process reaction curve for yD top composition.

Estimation of process gain, time constant and dead time for controller tuning.



(Figure 4.10) Process reaction curve (yD) for generating tuning parameters.

From figure 9 Δv is calculated and from Figure 4.10 Δy is taken of the graph to solve for the process gain. Time constant and dead time is then read off the plot.

$$Kp = \frac{\Delta y}{\Delta v} \frac{\Delta \text{ process value}}{\Delta \text{ manipulated value}} \quad (4.1)$$

$$= \frac{0.9442 - 0.87}{2.5 - 233} = 0.37$$

$T\theta$ = 330-200 = 130 secs.

To = 0.03secs.

The values above were substituted into Cohen-Coon equation for calculating PID controller (P.C.Chau 2002)

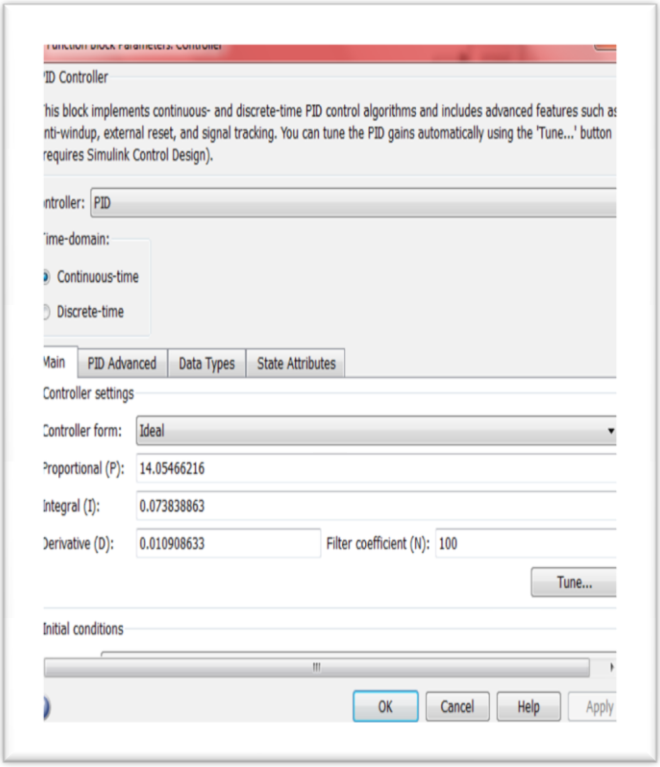
$$Kc = \frac{1}{Kp}\frac{T\theta}{To}\left(\frac{4}{3} + \frac{To}{4T\theta}\right) \qquad (4.2)$$

$$Ti = To\frac{32+6To/T\theta}{13+8To/T\theta} \qquad (4.3)$$

$$Td = To\frac{4}{11+2To/T\theta} \qquad (4.4)$$

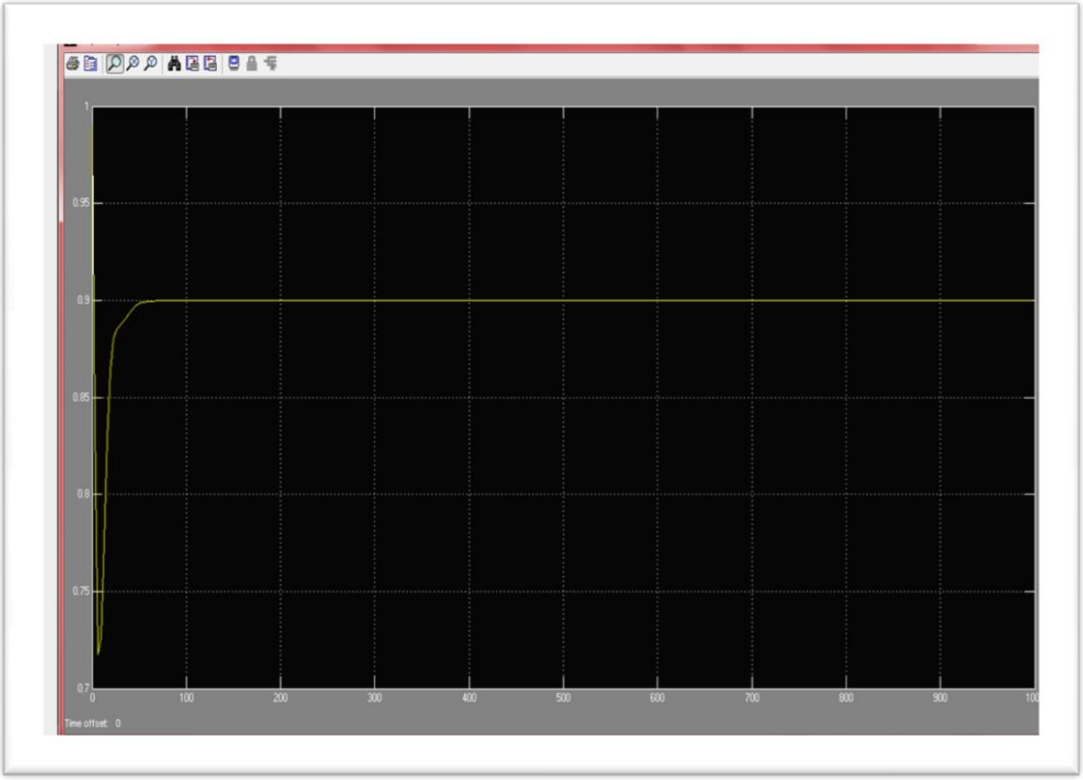
The following values were obtained for;
Kc = 14.05466216
Ti = 0.073838863
Td = 0.010908633

With the values for controller gain, integral time and derivative time for a PID controller above, further simulation was conducted. Below figure 4.10 is the display of PID block used for the simulation and control of (yD).



(Figure 4.11) PID controller block

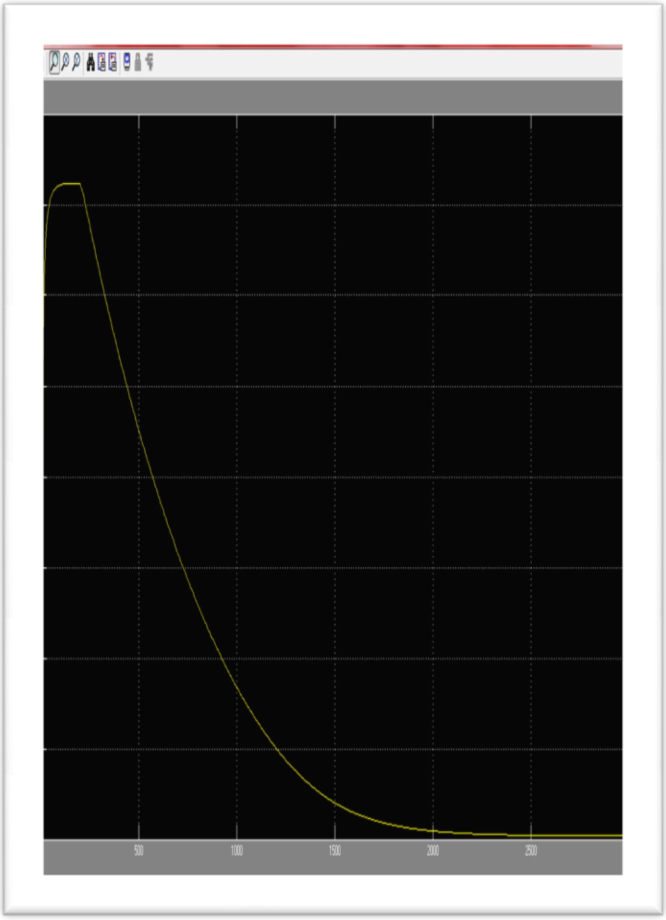
After simulation for 1000 seconds, Figure 4.12 is the plot of the response, the controller was found to be effective driving to process output (yD) to set-point. Further simulation was performed; it was for 1500 seconds, this time to check for robustness and stability. This was done by increasing the level of disturbance into the process and set-point alteration. Feed rate was changed from 1.01kmol/min to 1.05 kmol/min 200 seconds into the simulation while the set was altered from initial value 0.85 to 0.90, 500 seconds into the simulation. Figure 4.13 is the plot of response. It was observed that the controller exhibits total robustness.



(Figure 4.12) well tuned response of top composition

(Figure 4.13) Response with elevated disturbance and set-point alteration

To tune the controller for bottom composition (xB), similar procedure used for the estimation of the parameters used for tuning for (yD) process "reaction curve " is also used figure 4.14. The curve shows different behavior from the response shown on figure 4.9 for (yD). The response was generated using step change of input from the initial value of input 1.5 to final value 3.2063, 200 seconds into the simulation which lasted 3000 seconds. The process got to steady at 2500 second, afterwards the values of the dead time; time constant and process gain are read of the plot.



(Figure 4.14) Process reaction curve for bottom composition

Kp = $\frac{\Delta y}{\Delta v} \frac{\Delta \text{ process value}}{\Delta \text{ manipulated value}}$ (from equation 4.1)

Kp = $\frac{0.0039 - 0.7235}{3.2063 - 1.5} = = \frac{-0.7196}{1.7063}$ = -0.4217312313

$T\theta$ = 910-200 = 130 secs

To = 60secs

The following values were substituted into Ziegler-Nichols equation for estimating PID controller parameter; equation 4.2, 4.3, 4.4

The following values were obtained for;

Kc =-36.07578615

Ti =112

Td = 28

The tuning parameter was inserted into the PID controller to control the bottom composition.

The above values were not able to tune the controller.

5 Evaluation

5.1Evaluation Of Research Methods Used In The Project

The project from onset was full of research till the final minute. The research was completely essential for the comprehension of the ideas, method and to settle on the best advances in solving the various problems that came upon in the project.

The first research into the project was trying to understand the dynamic behaviour of the distillation column. This made it possible to gain reasonable knowledge on how distillations occur in various columns which were basically unfamiliar to me before the project commenced.

Secondly, research into process control in chemical processes was a vital aspect that I indulged in. Prior to the conception of this research I had poor knowledge of process control but now reasonable knowledge has been acquired. The use MATLAB-SIMULINKS presented the next uphill task which was tackled adequately.

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5.2Evaluation Of Personal Objective

Having researched and developed a controller for some part of a distillation column, the project aim was not fully achieved. The personal goal of the project was to device a complete control [LV] configuration to control the throughput of distillation column and implementing other advanced control techniques to analyse the throughput.

Moreover personally the level achieved have broadened my comprehension on process control and use of MATLAB-SIMULINKS.

6 Recomendation

Despite huge literature publications on distillation control there are so many constraints which remain to be fixed these includes the influence of pressure on composition response, the mode and manner vapour-liquid equilibrium affects composition response and also proper comprehension of the effect of energy balances on composition control. The most important aim of this research is to identify the most effective control structure applied to distillation.

Further work on this research is recommended. On the accomplish of the top composition control using a PID controller, application of other advanced type of controller which includes; Internal Model Control (IMC), Cascade Control, Model Predictive Control (MPC), Programmable Fuzzy Logic Control, is recommended to compare the structure which yields better throughput .

Also the 5 x 5 control of distillation column, presently the knowledge of a 5 x 5 control is unknown, this strategy could present the most effective way a distillation column could be controlled, may be under some circumstances this kind of control scheme might be reasonable.

7 Conclusion

In this project, we reviewed the control of distillation column, by prioritising on composition control. It was discovered that the PID controller which is the most popular controller used in industries are very effective and drives the process to set point with minimum offset and also with less time. With the appropriate process values gotten from the process reaction curve, the tuning parameters were calculated using Cohen-Coon equation for open loop step change. It was found to be a very robust control after so many alterations. Same procedure was applied to the bottom composition to estimate the tuning parameter. With values of time constant, process gain and dead time read of the process reaction curve, they were substituted into the Cohen-Coon equation to estimate the tuning parameter for the controller. The values of the estimated were not satisfying further attempt was made by using Ziegler-Nichols equation for the estimation of controller parameter; this too did not yield a desired result.

Due to time constrain, the development of the most appropriate procedure to estimate the right tuning parameter for the bottom composition control was not accomplished. This did not enable the application of [LV] configuration for the controlling distillation columns.

Various controllers tuning procedure was outlined in Appendix A, if more time was allocated all the procedure related to the process researched will be applied to the work and beyond, which will surely yield a positive result.

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