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M'HAMED BOUGARA-BOUMERDES UNIVERSITY

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Department of Automation and Electrification of Industrial Processes

Master's Thesis

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Field : Hydrocarbones

Specialty : Industrial Instrimentation in Petrochemical

Comparative Interactions Analysis of APC and PID Controllers in Distillation Column at Skikda GL1K Complex

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In front of the jury :

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Thanks

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Dedication

For the courage given to me by God's grace, inspiring me to complete this project.

To my dear parents, « M. Lahcen and B. Zohra », your endless love, sacrifices, and support have been the strong base for my academic journey and career. Thank you for always being by my side. I am very grateful to you for teaching me the values of hard work and ambition.

To my entire family, Especially my wife your presence and encouragement have illuminated every step of my path. You have been my companions, my sources of inspiration, and my greatest allies.

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To each of the individuals mentioned here, as well as to all those who have contributed to my academic and personal journey, I extend my sincerest thanks. This thesis is dedicated to your unwavering support, friendship, and constant inspiration.

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Dedication

For the courage bestowed upon me by the grace of God, inspiring me to accomplish this humble endeavor.

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To my loyal friends, you are more than just classmates; you are my second family. Your encouragement, unwavering support, and shared memories will remain etched in my heart forever.

To the team at D. Gitallia, your collaboration, expertise, and dedication have been integral to the completion of this project. I am grateful to have had the opportunity to work with such talented and committed professionals.

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KREMIS Khalid

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Abreviations list

LNG : Liquefied Natural Gas

LPG : Liquefied Petroleum Gas

APC : Advanced Process Control

DCS : Distributed Control System

PID : Correcteur Proportional Integral Derivative

RMPCT : Robust Multivariable Predictive Control Technology

CV : Controlled Variable

MV : Manipulated Variable

DV : Disturbance Variable

PL : Linear Programming for linear optimization

EU : Enginnering Units

FF : Feed Forwads

PSOS : Profit Suite Operator Statio

ST : Set Point

PV : Process Value

OP : Operation Point

RGA : Relative Gains Array

U: Command vector (inputs, setpoints) of dimension (p×1).

Y: Output vector (measurement) of dimension (q×1).

p: Number of system commands.

q: Number of system outputs

Z : Disturbance

$\boldsymbol{\lambda}$: Gain Relatif

Λ : matrix of relative gains

.x : is the product of Hadamard (the element-wise product)

$\boldsymbol{K_s}$: is the matrix of static gains

$\boldsymbol{K_s^{-1}}$:inverse of the matrix

$\boldsymbol{K_{ij}}$: is the static gain between u_i and y_j.

$\boldsymbol{K_p}$: Proportional gain

K_i : Integral gain

K_d : Derivative gain

BLT : Method for Multivariable Systems

W_{cr} : critical frequency

K_{cr} :critical gain

V_D :distillate flow rate from the condenser

V_R :residue flow rate from the reboiler

V_L : reflux flow rate at the top of the column

V_B :vapor flow rate at the reboiler

V_C :vapor flow rate at the top of the column

M_B :reboiler hold-up

M_C :condenser hold-up

P_C :column hold-up or pressure

X_C :top column composition

X_B :bottom column composition

Abstract

The objective of this thesis is to evaluate the performance of Honeywell's Advanced Process Control (APC) system in multivariable processes and to compare it with traditional PID controllers. This study was conducted at the GL1K complex in Skikda, which specializes in the liquefaction of natural gas. We examined the effectiveness of APC in managing the complex interactions within the process, focusing on key performance indicators such as stability, efficiency, and product quality. The results demonstrated that APC significantly improves process performance and reduces variability compared to conventional PID controllers. This research highlights the advantages of advanced control strategies in optimizing industrial operations and ensuring superior control over multivariable systems.

Resumé

L'objectif de cette thèse est d'évaluer la performance du système de contrôle avancé des procédés (APC) de Honeywell dans des processus multivariables et de la comparer aux régulateurs PID traditionnels. Cette étude a été menée au complexe GL1K de Skikda, spécialisé dans la liquéfaction du gaz naturel. Nous avons examiné l'efficacité de l'APC dans la gestion des interactions complexes au sein du processus, en nous concentrant sur des indicateurs clés de performance tels que la stabilité, l'efficacité et la qualité du produit. Les résultats ont montré que l'APC améliore considérablement les performances du processus et réduit la variabilité par rapport aux régulateurs PID conventionnels. Cette recherche met en évidence les avantages des stratégies de contrôle avancées dans l'optimisation des opérations industrielles et l'assurance d'un contrôle supérieur des systèmes multivariables.

تلخيص

من شركة هانيويل في العمليات متعددة المتغيرات (APC) تهدف هذه الأطروحة إلى تقييم أداء نظام التحكم المتقدم في العمليات
في سكيكدة، والذي يتخصص في تسييل الغاز GL1K التقليدية. تم إجراء هذه الدراسة في مجمع PID ومقارنته مع أدوات التحكم
في إدارة التفاعلات المعقدة داخل العملية، مع التركيز على مؤشرات الأداء الرئيسية مثل APC الطبيعي. قمنا بفحص فعالية
يحسن بشكل كبير من أداء العملية ويقلل من التباين مقارنةً بأدوات التحكم APC الاستقرار، الكفاءة، وجودة المنتج. أظهرت النتائج أن
التقليدية. تسلط هذه الدراسة الضوء على مزايا استراتيجيات التحكم المتقدمة في تحسين العمليات الصناعية وضمان التحكم PID
الفائق في الأنظمة متعددة المتغيرات.

General Introduction

General Introduction :

Liquefied Natural Gas (LNG) has emerged as a pivotal global energy source, renowned for its cleanliness and widespread availability. As a non-renewable primary energy resource, LNG plays a critical role in meeting the world's energy demands. Its production involves sophisticated processes, particularly within fractionation units equipped with distillation columns. These columns are integral to separating natural gas into its constituent components, each crucial for various industrial and commercial applications.

Controlling distillation columns poses a formidable challenge due to the intricate interactions between input variables (such as temperature and pressure) and output variables (such as product purity and yield). Achieving optimal operational efficiency requires advanced control strategies capable of managing these complex dynamics effectively.

In this study, we delve into several critical aspects by describing Fractionation Units and the Depropanizer Column: We explore the intricacies of fractionation units, focusing on the depropanizer column's role in recovering propane efficiently. Understanding its operational parameters is essential for optimizing production outputs and ensuring product quality.

we moved to Advanced Process Control (APC) and We provided a comprehensive overview of APC, detailing its components and functionality. APC systems are vital in industrial settings for maintaining stringent production standards and maximizing throughput.

then we moved to the Practical Application of Multivariable Control, in This section we delved into the implementation of multivariable control systems. It covers the relative gain matrix and interaction analysis techniques used to design robust control strategies. Practical simulations on distillation columns illustrate how PID controllers, employing the BLT method, manage varying inputs and disturbances to optimize process performance.

we moved to the Validation of Control Systems which we analyzed the relative gain matrices of transfer functions between manipulated variables (MVs) and controlled variables (CVs). The goal is to ensure that each MV effectively regulates its corresponding CV without adversely impacting other variables, thus enhancing overall system stability and efficiency.

This structured approach aims to elucidate the complexities of LNG production, fractionation processes, and the critical role of advanced control systems in optimizing industrial operations. By integrating theoretical insights with practical simulations, we contribute to advancing the understanding and application of efficient energy management strategies in complex industrial environments.

CHAPTER.I. Presentation of the GL1K Complex

I.1. Introduction

As part of the strategic objective to enhance the value of natural gas resources, primarily from the Hassi R'Mel field, LNG plants have been constructed in the north of the country. Their main purpose is the export of LNG to Europe and the USA via LNG carriers.

Among these plants, we mention the GL1/K complex in Skikda, whose construction began in March 1969 and production started in November 1972. This complex covers an area of 90 hectares and receives natural gas from the Hassi R'Mel field through a 580 km long, 40-inch diameter pipeline. Its annual production capacity is 6.7 million m³ of LNG, and it has a storage capacity of 196,000 m³ of LNG. It employs 1,200 permanent workers.

I.2. History of the Complex (GL1/K)

I.2.1. SONATRACH

SONATRACH was created on December 31, 1963, by decree 63-491 to take responsibility for the transportation and marketing of hydrocarbons. It quickly expanded by decree No. 66-296 on September 22, 1966, to become a company involved in the exploration, production, transportation, processing, and marketing of hydrocarbons. In 1996, SONATRACH's powers extended to include exploration, produc

tion, refining, and the manufacture of chemical products. It held only 20% of the production from foreign companies operating in Algeria, such as S.N.REPAL CAMEL.

Figure I.1 Panoramic view of the GL1K complex

I.3. Geographical Location

The plant is located 3 km east of the city of Skikda and currently covers an area of approximately 92 hectares.

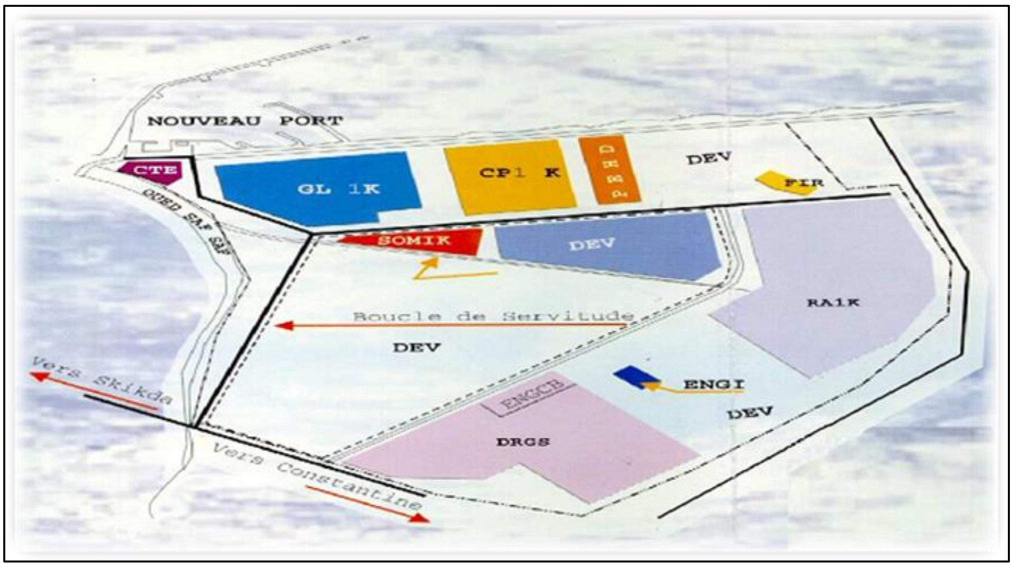


Figure I.2 Situation of the GL1/K complex in the industrial zone

I.4. Presentation of the GL1/K Complex in Skikda:

The GL1/K natural gas liquefaction complex in Skikda is located approximately 3.8 km east of Skikda, a coastal city in northeastern Algeria. It occupies a total area of 93 hectares. The plant is bordered to the north by the Mediterranean Sea, to the east by CP1/K, to the south by the refinery and terminal, and to the west by SONELGAZ. **[1]**

The project for constructing the GL1/K complex was carried out in three phases according to a planned schedule:

In 1967, a call for tenders was issued, with the bidders being foreign companies.

The project studies spanned three years, from 1965 to 1967.

Construction work began in 1967.

Briefly, some of the companies that participated in the establishment of the various units of the complex include Techip, Pritchard Rhodes, Pullman Kellog, and I.H.T.

The GL1/K complex is supplied with natural gas (NG) from the HASSI RAMEL field, transported to the complex via a 580 km long, 40-inch diameter pipeline (1 inch = 2.54 cm). After treatment, liquefaction, and storage, LNG is loaded onto tankers. The annual production capacity of GL1/K is 13.2 million m³ of LNG, with a storage capacity of 308,000 m³ of LNG.

In addition to LNG, the complex produces:

-1029 tons/day of ethane (C2).

-978 tons/day of propane (C3).

-680 tons/day of butane (C4).

-373 tons/day of light naphtha (C5).

Six (06) units of finished product liquefaction: (U10, U20, U40, U5p, U6p)

The first three units (U10, U20, U30) were built by the French company TECHNIP according to the TEAL process and started production in November 1972. The U40 unit was constructed 85% by Pritchard Rhodes and taken over by Pullman Kellog in March 1979, with a capacity of 6000 m³/day of LNG. The 5p and 6p units were also built by Pritchard Rhodes (57%) and taken over by Pullman Kellog in April 1979, with a total capacity of:

An auxiliary plant designed to supply the liquefaction units with:

I.4.1. Demineralized Water

Used to feed the various boilers of the complex. The circuit includes:

Raw Water Treatment: Consisting of:

Four (04) parallel production lines of fresh water by ion exchange resin from seawater.

Two (02) stations producing fresh water by vacuum distillation, with a production capacity of 500 tons/hour.

I.4.2. Electricity and Steam

The unit has three (03) turbo-alternator groups with a nominal power of 7.5 megawatts (MW). Their turbines are driven by steam from three (03) boilers, each with a capacity of 45 tons/hour of steam. Part of the steam produced is used to start the boilers of the units, with the rest used for other equipment such as seawater desalination.

I.4.3. Instrument Air

Dust-free, dried, and compressed air at seven (07) bars. It is produced by a series of filters, dryers, and compressors installed in the central plant.

I.4.4. Nitrogen

Produced by air distillation, essential and particularly used during measurements, with a production capacity of 400 m³/day.

A storage and operation unit for finished products and two LNG loading centers:

Responsible for storing LNG:

Five (05) LNG storage tanks.

Two (02) LNG loading centers.

Storage capacity of 315,000 m³ of liquid and shipping to loading docks for tankers.

A LPG processing unit:

Separates butane and propane from the butane-propane mix from the liquefaction units.

I.5. Presentation of Existing Units

The GL1/K LNG complex currently includes:

- (03) liquefaction units in production.
- storage and dispatch unit.
- LPG unit.
- Auxiliary plant.

The complex mainly includes:

- Three (03) LNG liquefaction trains (U10, U5p, and U6p).
- An LPG unit for processing and storing propane and butane.

A storage park and removal facilities, including:

- Three (03) LNG storage tanks with a total capacity of 196,000 m^3.
- Two (02) LNG loading stations.

A central unit for utility production:

- Demineralized and distilled water.
- Steam.
- Instrument and service air.
- Nitrogen.
- Fuel gas.

In addition to LNG, the complex produces:

- 1915 tons/day of ethane.
- 1818 tons/day of propane.
- 1554 tons/day of butane.
- 917 tons/day of light naphtha.

Due to an incident on January 19, 2004, the complex's production capacity was reduced by 40%.

I.5.1. Auxiliary Plant

The auxiliary plant is an autonomous system designed to supply units 10, 5p, and 6p with electricity, air, cooling water, and nitrogen.

I.5.2. Storage and Dispatch

The LNG storage unit includes 05 tanks, with 03 of 56,000 m^3 capacity each and 02 of 70,000 m^3. This unit has two LNG dispatch pumps, with a loading rate of 6000 m^3/hour and 12000 m^3/hour.

I.5.2.1. LPG Unit

Constructed by the Japanese company I.H.I, it initially started in September 1973. Its purpose is to process the product from units 10, 20, 30, and 40 to separate it into commercial propane and butane. It also ensures the storage of these products and those from units 5 and 6. It handles the cooling and storage of propane and butane from RA1K, with a storage capacity of:

- 02 propane tanks: 12,500 tons each, mainly for export.
- 01 butane tank: 20,000 tons, primarily for national consumption.

I.6. Mega Train

The New Skikda LNG Train project (Algeria) consists of an independent 4.5 MTPA LNG train, utilities, and offsites (storage), intended to replace the capacities of units 20, 30, and 40, which were destroyed by fire in 2004. The train includes the following sections: Natural gas metering and compression/feed system, AMDEA treatment using the BASF process, natural gas treatment (decarbonation, dehydration, mercury removal), APCI liquefaction process, and fractionation of ethane, propane, butane, and natural gasoline. **[1]**

Table I.1 table represents the different products of the mega train

Product	**Quantity**
LNG	611 999 kg/h
Enriched helium gas	20 205 kg/h
Ethane	22 404 kg/h
Propane	28 235 kg/h
Butane	23 312 kg/h
Natural gasoline	14 786 kg/h

I.6.1. Presentation of the Mega Train Units:

I.6.1.1. Utilities

I.6.1.1.1. Unit 51: Electricity Production

There are five generators, 51-MJ01-A/B/C/D/E, driven by gas turbines powered by fuel gas from the fuel gas tank. Each generator can produce 21.74 MW at 11 KV and 50 Hz, totaling 110 MW when all generators are in operation. There are also five Diesel Generator Sets (DEGs), 51-MJ02-A/B/C/D/E, for backup and startup, each producing 1.719 MW.

The Skikda plant requires about 80 MW when operating at its design capacity. **[2]**

I.6.1.1.2. Unit 56: Instrument Air and Service Air Systems

The Instrument Air and Service Air Systems (Unit 56) produce compressed air and instrument air for all plant users.

There are three (03) air compressor sets, 56-MJ01-A/B/C, operating at 100%: one is in service, and the other two are on standby. Each is designed for 3345 Nm^3/h at a nominal discharge pressure of 10 bar eff. The atmospheric air compressed by the compressors passes through the air-wet separator 56-ML02 and then through two (02) air dryer sets, 56-ML01-A/B, operating at 100%, with one in service and the other on standby.

I.6.1.1.3. Water Systems

The objective of the cooling water system, Unit 09, is to:

- Provide cooling water or make-up water for the machines in the mega train unit.

The objective of the water system in Unit 59 is to:

- Store and deliver industrial water (service water) to be used in the new LNG train for the fire water system, treat it to make it potable, and store and deliver potable water.
- Store and deliver demineralized water to be used in the new LNG train.

I.6.1.1.4. Hot Oil Systems

The hot oil used in Units 08, 18, and 58 is Shell Thermia Oil B, which primarily consists of hydrocarbons with carbon numbers between C20 and C50.

- Unit 08 Storage Tank:

The hot oil storage tank 08-MF01 is an atmospheric type designed to hold the entire inventory of hot oil for the LNG train plus a 10% safety margin. This tank is nitrogen-blanketed.

- Unit 18 Hot Oil System:

The hot oil system of Unit 18 is intended to meet the heat demand of the train unit reboilers.

- Unit 58 Hot Oil System for Fractionation:

The hot oil system of Unit 58 is intended to meet the heat demand of the reboilers in the fractionation unit (Unit 07), the reboiler of the purification column 15-MC15, and the GN reheater 01-MC02.

I.6.1.1.5. Unit 65 Flare System

The purpose of the flare system, Unit 65, is to ensure the safe treatment/disposal of vapor and liquid hydrocarbon streams generated during plant commissioning, shutdowns, malfunctions/failures, and emergencies.

I.6.2. Train and Fractionation

I.6.2.1. Unit 01 Feed Gas Compression/Measurement

The feed gas GN comes from the Hassi R'mel fields. It passes through a measurement system 01-IC03 and then to the GN compression system to compress the gas to the required pressure for downstream processes.

I.6.2.2. Unit 12 Decarbonation (AGRU)

Its purpose is to reduce the CO2 content in the GN to less than 50 ppmv using aMDEA under BASF license, to prevent blockages (CO2 solidification) in downstream units' equipment (Liquefaction, Unit 15).

I.6.2.3. Unit 13 Dehydration:

Its goal is to dry the gas to achieve a residual water content of less than 01 ppmv, to prevent hydrate formation in the lines and equipment operating at low temperatures. The unit operates using three molecular sieve bed dryers, with one in adsorption mode, one on standby, and one in regeneration mode. Regeneration is done using heated and depressurized gas from the liquefaction process.

I.6.2.4. Unit 14 Mercury Removal:

Its purpose is to reduce mercury levels in the GN by adsorption using a sulfur-impregnated activated carbon bed. If mercury is not removed, there is a risk of mechanical damage to downstream units and equipment made of aluminum.

I.6.2.5. Liquefaction and Refrigeration:

I.6.2.5.1. Unit 15 Liquefaction

The objective of liquefaction is to:

Cool the GN entering the purification column, separating heavy hydrocarbons from GN, and producing a liquid feed stream to the fractionation unit (Unit 07).

Cool and partially liquefy the head of the purification column in the main exchanger.

Cool the LPG and iso-pentane from fractionation with MR refrigerant and mix them with GN to produce LNG in the main exchanger.

Cool and liquefy the GNT (devoid of heavy hydrocarbons) to produce LNG in the main exchanger.

Produce a helium-rich gas for export.

Remove nitrogen from the LNG and produce HP fuel gas.

Mix BOG from the new and existing LNG storage tanks with some GN and regeneration gas from the dehydration system to produce plant fuel gas.

Evacuate LNG to the new LNG storage tank.

Produce methane for refrigerant makeup.

I.6.2.5.2. Unit 16 Refrigeration

The objective of refrigeration, Unit 16, is to:

Produce and circulate MR and propane refrigerant using turbo-compressors and condensers.

Cool and partially liquefy MR using propane refrigerant.

Cool and liquefy GN using MR refrigerant in 15-MC05.

I.6.2.5.3. Unit 07 Fractionation

The objective of the fractionation unit, Unit 07, is to:

Produce quantities of ethane, propane, butane, and gasoline at least equal to those produced by the existing units 20, 30, and 40.

Produce the ethane and propane needed for refrigerant makeup.

The fractionation unit consists of the following distillation columns and associated equipment:

- Demethanizer - 07-MD01
- Demethanizer - 07-MD02
- Depropanizer - 07-MD04
- Debutanizer - 07-MD06
- Deisopentanizer - 07-MD11
- GPL reinjection 07MD08.

I.6.2.5.4. Storage and Shipping

Unit 71 LNG:

The purpose of Unit 71, LNG Storage and Shipping, is to ensure the storage of LNG produced by the new train and the existing Skikda units, using the LNG storage tank 71-MF01.

Unit 76 Propane:

The purpose of this unit is to ensure the storage of propane produced by the new LNG train in the new propane storage tank 76-MF01.

Unit 76 Butane:

The purpose of this unit is to ensure the storage and transfer of butane produced by the new LNG train at Skikda, using the butane storage tank 76-MF02.

Unit 76 Gasoline:

The purpose of this unit is to ensure the storage and transfer of gasoline produced by the new LNG train at Skikda in sphere 76MD03.

Laboratory Presentation:

The laboratory operates 24 hours a day to meet all analysis needs required by the various units, including analyses requested by department engineers, such as oil analyses and imported product analyses. The analyses are conducted according to a program described by the laboratory managers and the different departments.

The laboratory is responsible for controlling the products and materials entering production. It serves as an indicator of the complex's proper functioning. Two types of analysis are performed in the GL1/K laboratory:

- Gas analysis by chromatography.
- Specific water analysis.

I.7. Conclusion

The GL1/K process is one of Algeria's important new hydrocarbon hubs, strategically located in Skikda. Nearly all units have been renovated since 2008, making it a modern and vital complex that significantly advances Algeria's industrial sector and strengthens its economic relations globally. In this chapter, we presented the various units responsible for natural gas treatment, liquefaction, and the extraction of several by-products. The next chapter will focus on the description of the fractionation unit, demethanizer, deethanizer, depropanizer, debutanizer, and deisopentanizer, providing a detailed overview of these critical components.

CHAPTER.II. Description of the Fractionation Unit

II.1. Introduction

To separate the so-called "heavy" constituents from natural gas in order to meet the commercial specifications of LNG, the constituents to be separated are as follows:

- Ethane
- Propane
- Butane
- Gasoline (C5+)

These components must be separated from natural gas to produce Liquefied Natural Gas (LNG) while meeting commercial standards. The new LNG Train is designed to work with the total reinjection of these products, thus ensuring production in line with market requirements.

II.2. Fractionation columns and product usage

In the fractionation unit described, there are five distillation columns, each installed to separate the lightest constituent and recover it at the head from the load to which it has been fed. This process allows the different components to be recovered according to their volatility, thus ensuring an efficient separation of the constituents from the initial mixture. **[3]**

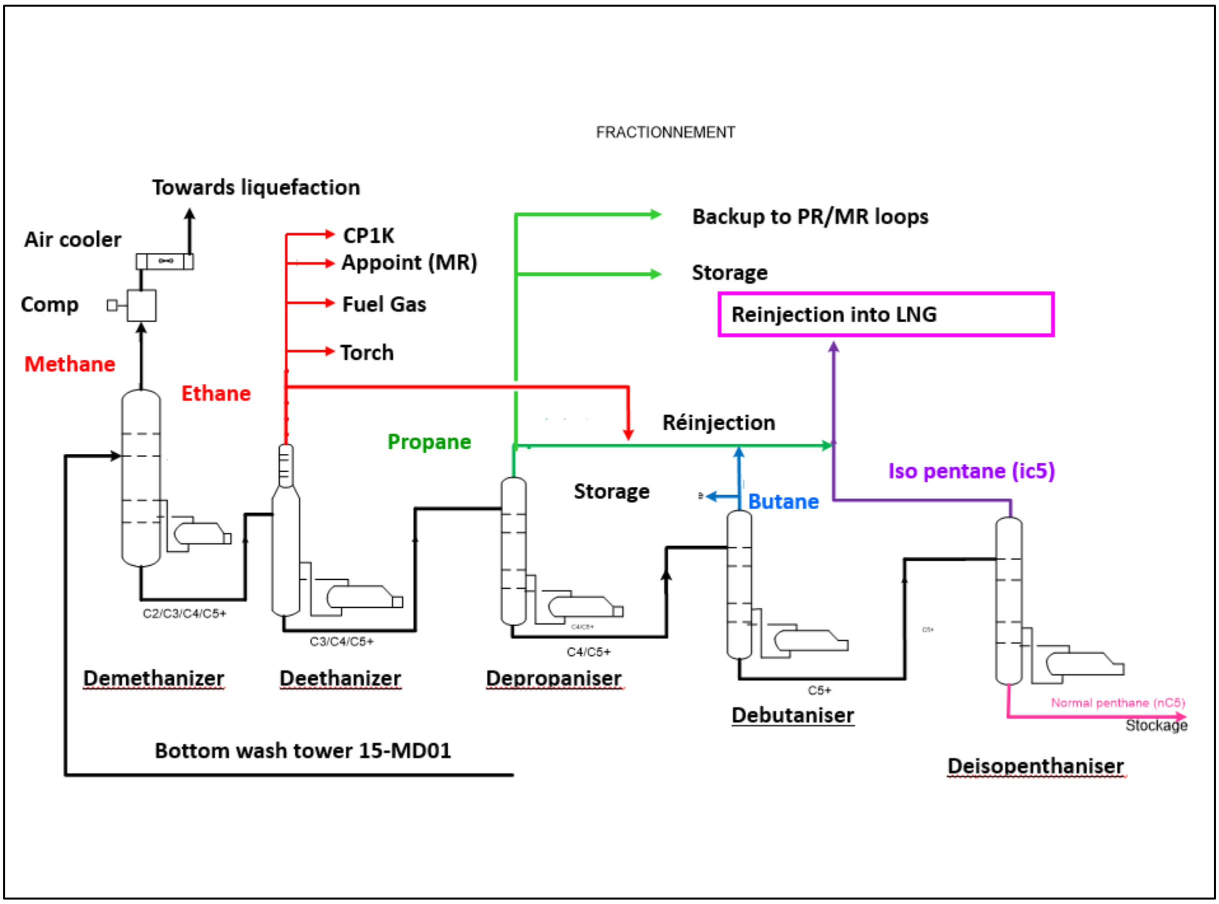


Figure II.1 Descriptive diagram of the Fractionation Unit

Table II.1 Use of products

Ethane	- Supply to CP-1/K - Re-injection into LNG	- Makeup in the MR loop - Re-injection into Fuel Gas
Propane	- Storage and shipping - Re-injection into LNG - Makeup in the PR loop	
Butane	- Storage and shipping - Re-injection into LNG	
Gazoline (iC5)	- Re-injection into LNG	
Gazoline (nC5)	- Storage and shipping to RA-1/K	

II.2.1.Demethanizer 07-MD01

The demethanizer is a type of distillation column used to separate methane from a mixture of gases. It is equipped with 17 trayed plates with flapets and calottes, and it receives a charge containing methane from the bottom of the wash tower. The charge is fed into the column at the sixth tray. The purpose of the demethanizer is to separate all the methane from the charge, as its presence in the residues that feed the deethanizer would disrupt its operation.[4]

The overhead product of the demethanizer is partially condensed in the propane condenser (07MC16), and the liquid portion is collected in the balloon (07MD10) to be used as reflux. The remaining part is sent to the suction of the recycle compressor (b.o.g recycle) to increase its pressure and reintroduce it into the line that feeds the wash tower. The demethanizer is equipped with two reflux pumps (07-MJ08 and 07-MJ08/A). The reboiler (07MC01) is of the kettle type and uses hot oil from unit 58 for heating.

The molar ratio of methane to ethane in the residues must be limited to a maximum of 0.015 to meet the quality specifications of the ethane produced in the deethanizer.

The schematization of these equipment is presented in the figure below :

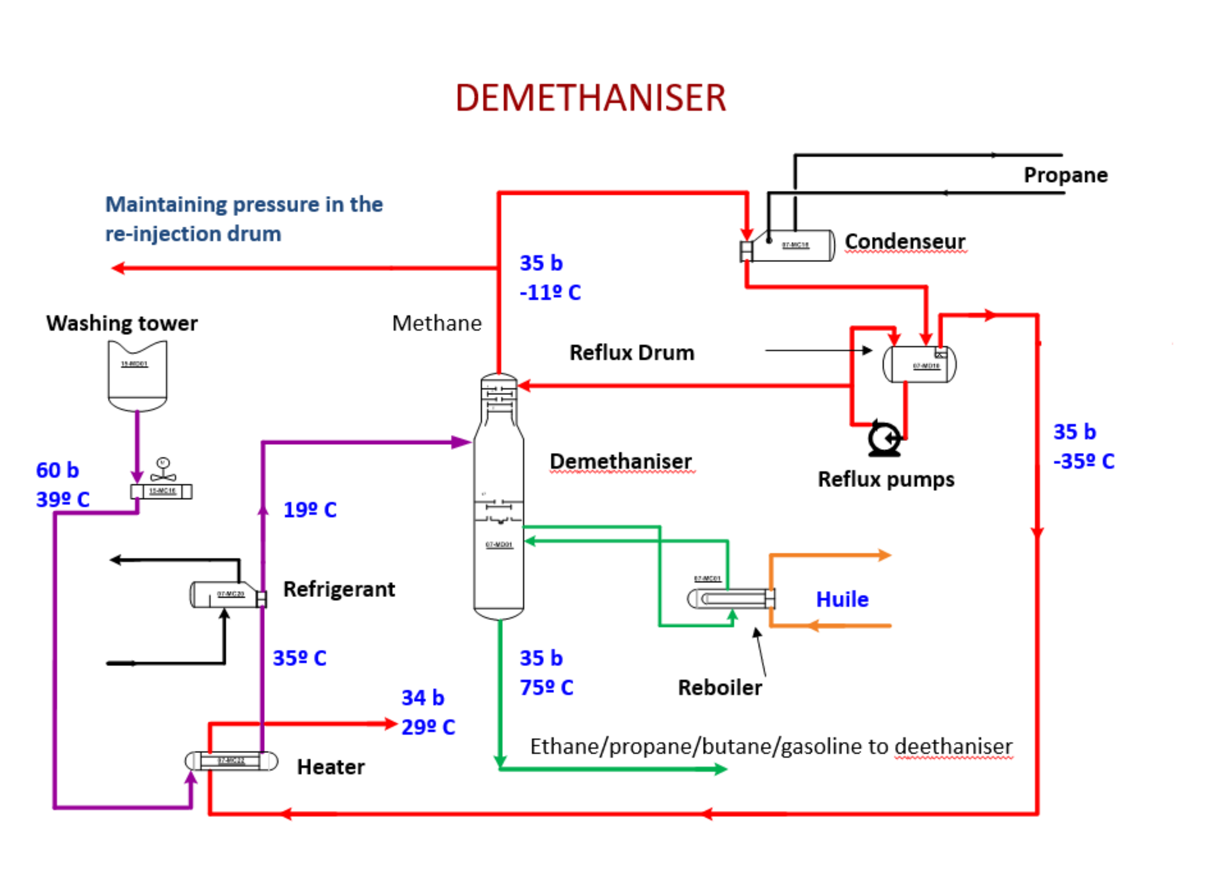


Figure II.2 Descriptive diagram of the Demethanizer

II.2.2.Deethanizer 07-MD02

the deethanizer is a partial condensation distillation column with 35 trayed plates with flapets and calottes. It receives a feed (from ethane to gasoline) from the bottom of the demethanizer to separate and recover ethane at the top. The feed is introduced above the 15th tray.

The reboiler 07-MC02, of kettle type, uses hot oil from unit 58 to heat the products at the bottom of the deethanizer to limit the ethane content in the residues that feed the next column (depropanizer).

The overhead product (98% ethane) is partially condensed in the propane condenser 07-MC03 and then recovered in the reflux drum 07-MD03 to be

- Used as reflux for the deethanizer.
- For reinjection into the LNG (after refrigeration).

The other part remains in gaseous form and is used (after reheating) For various purposes like:

- Feeding CP1-K.
- Supplementing the MCR loop.
- In the fuel gas network.

The deethanizer is equipped with two reflux pumps, 07-MJ01 and 07-MJ01/A.

The schematization of these equipment is represented in the figure below :

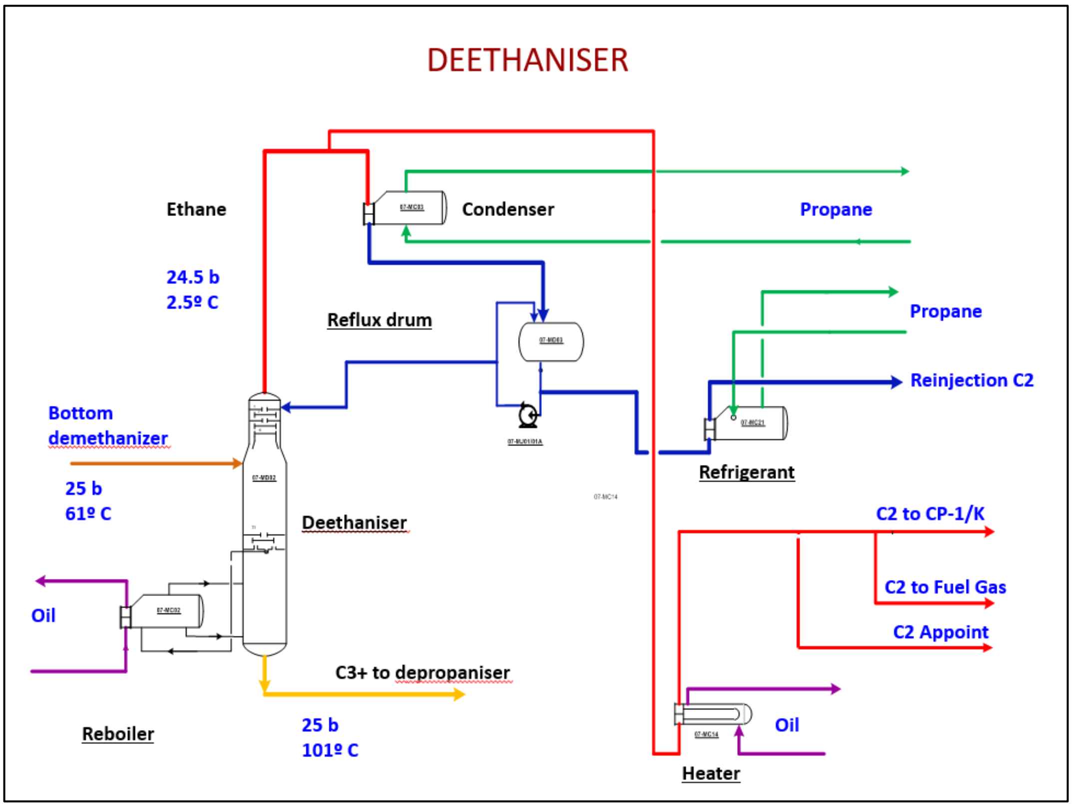


Figure II.3 Descriptive diagram of the Deethanizer

II.2.3.Debutanizer 07-MD06

The debutanizer is a type of distillation column used in natural gas processing to separate and recover butane from a mixture of hydrocarbons. It is a total condensation distillation column equipped with 32 cap valve trays and fed by the bottom liquid from the depropanizer above the 12th plate. The role of the debutanizer is to separate and recover the butane contained in the load coming from the bottom of the Depropanizer.

The reboiling is carried out by the 07-MC06 kettle type reboiler to limit the content of butane (butane and isobutane) in the bottom products which feed the column next (deisopentanizer).

The overhead vapors (butane and isobutane) are condensed by the air cooler 07-MC07 and then collected in the reflux flask 07-MD07. Part of the liquid butane is used as reflux to the debutanizer using reflux pumps 07-MJ03/A to the debutanizer, while the rest is cooled and sent to :

- Storage
- Reinjection into the LNG

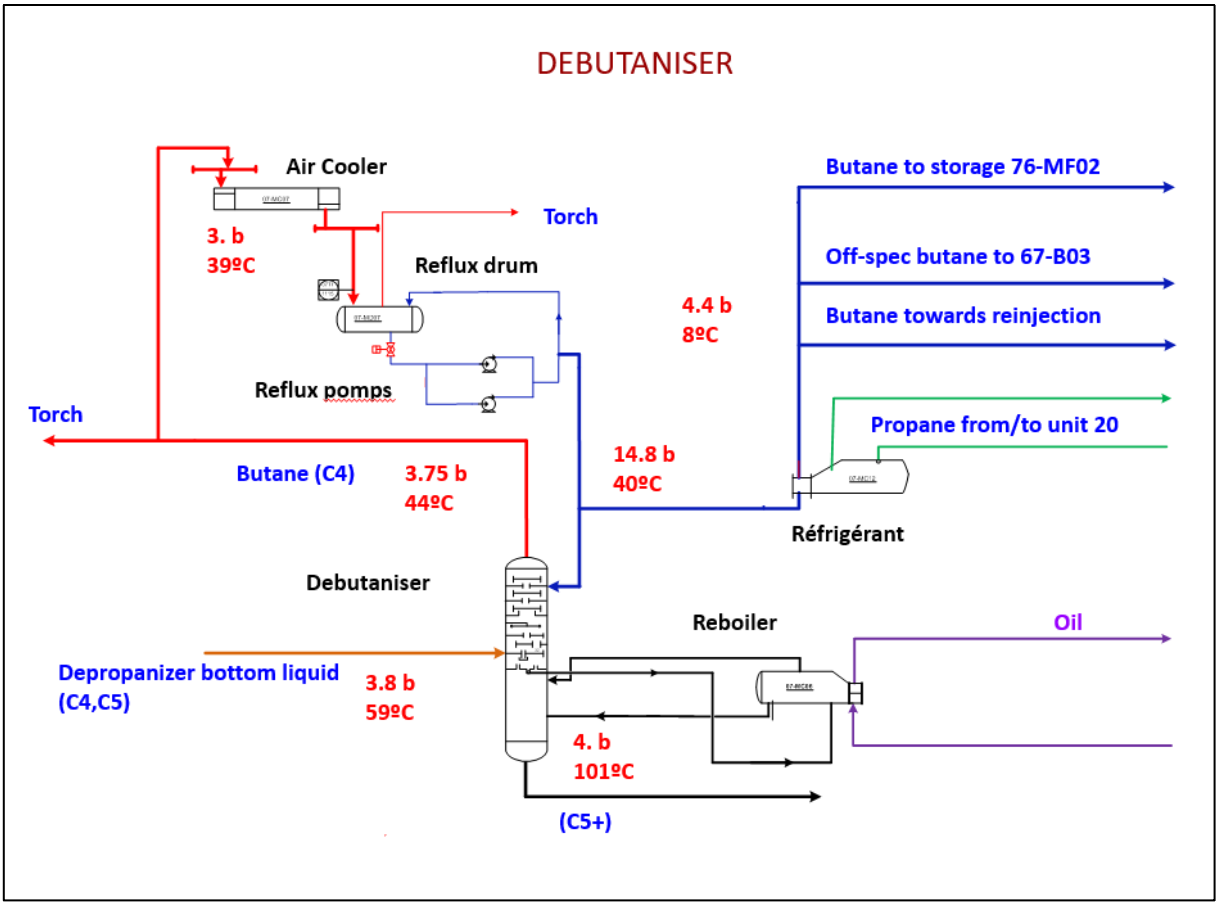


Figure II.4 Descriptive diagram of the Debutanizer

II.2.4. Deisopentanizer 07-MD11

The deisopentanizer is the last distillation column in the fractionation unit. It is equipped with 48 cap valve trays and receives the bottom products from the debutanizer above the 23rd tray. This column is installed to separate and recover isopentane in the overhead, resulting in a gasoline with a vapor pressure (TVR) of less than 0.77 bar effect at the bottom.

The reboiler of the deisopentanizer 07-MC24 is of the kettle type and uses hot oil from unit 58.

The overhead vapors are condensed by the air cooler 07-MC23 and then collected in the reflux flask 07-MD12. The reflux flow is provided by the two pumps 07-MJ09/A.

The isopentane produced is cooled in the propane exchangers (07-MC25, 07-MC26) and reinjected using pumps 07-MJ11/A directly into the LNG without passing through the reinjection unit. The gasoline produced at the bottom of the deisopentanizer is sent to RA1-K by pumps 07-MJ12/A and passes through the air cooler 07-MC13 for cooling.

The schematic representation of these equipment components is depicted in the following figure :

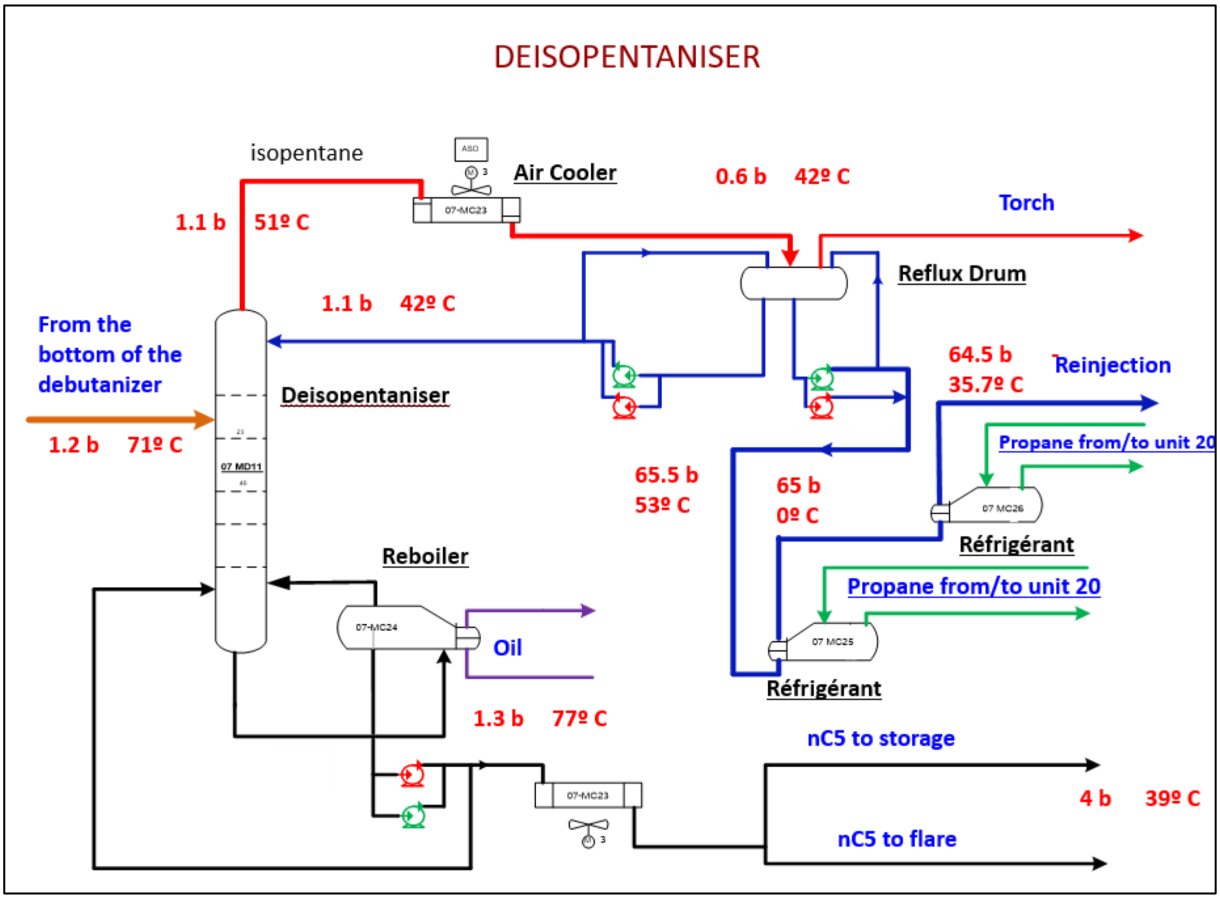


Figure II.5 Descriptive diagram of the Deisopentanizer

II.3. The description of the Depropanizer column

The depropanizer 07-MD04 is the third distillation column in the fractionation unit. Its primary objective is to produce high-quality refrigeration propane and to limit the propane content in the bottom products to meet the quality specifications of the butane at the top of the debutanizer.

II.3.1.Product Specification

Propane and butane specifications are as follows:

Propane :

- Density: 0.502
- Vapor pressure: 11.5 – 19.3 bar at 50°C
- Boiling point: -44°C

Butane :

- Density: 0.559 Min
- Vapor pressure: 6.9 bar at 50°C
- Boiling point: -1°C

II.3.2.Role and operation of each equipment in the section of depropanizer :

II.3.2.1.Depropanizer 07-MD04

Role : The depropanizer 07-MD04 is the third distillation column in the fractionation unit. Its function is to separate and recover propane at the top from the feed originating from the bottom of the deethanizer.

Reboiling occurs at the column's bottom to control the propane concentration in the bottom product of the depropanizer and ensure compliance with specifications.

Operation : the depropanizer is equipped with 37 trays and receives the feed (from propane to gasoline) from the bottom of the deethanizer 07-MD01. The reboiling of the bottom of the column is ensured by a reboiler using hot oil (07-MC02). The bottom products (propane, butane, and gasoline) are sent as feed to the depropanizer. This setup allows the depropanizer to function effectively in separating and recovering propane while maintaining the desired quality specifications for the subsequent processes in the fractionation unit.

II.3.2.2.Reboiler 07-MC04

The depropanizer is equipped with a kettle-type reboiler (07-MC04) with a tube-and-shell heat exchanger. The reboiling is ensured by hot oil that circulates on the tube side.

The depropanizer has a draw-off tray at the bottom that allows for direct flow to the reboiler. A baffle inside the reboiler allows the liquid to flood the tubes while leaving enough vapor space for the vapors to circulate back to the column.

The vapors return below the draw-off tray and rise through the valves, while the liquids flow over the baffle inside the shell of the reboiler and return to the bottom of the depropanizer column. This setup allows for efficient reboiling and separation of propane from the feed stream.

II.3.2.3.Air Cooler 07-MC05

Role : Condensing the vapors from the top of the depropanizer 07-MD04

Operation : The condenser is an induced draft dry cooler composed of 3 boxes and 9 motor fans. Each chamber is equipped with a variable-speed motor (3 in total) that receives a speed set point from the pressure controller of the depropanizer column. When there is an increase in pressure in the column, the speed of the fans is increased, and vice versa. This system effectively regulates the condensation of vapours in the condenser according to the pressure variations in the depropanizer column.

II.3.2.4.Reflux Drum 07-MD05

The liquids condensed in the depropanizer condenser 07-MC05 are collected in the reflux drum 07-MD05. A portion of the liquids are returned to the top of the depropanizer column as reflux, while the remaining liquids are pumped to the propane refrigeration system (07-

MC15).The uncondensed gas in the head condenser of the depropanizer (07-MC05) will be sent to the flare.The reflux drum is an essential component in the distillation process, as it allows for the recirculation of a portion of the distillate back to the column, which enhances the separation efficiency.

II.3.2.5.Reflux Pumps 07-MJ02/A

Role : Ensuring reflux for depropanizer 07-MD04

Operation : These two centrifugal pumps operate alternately (one in operation, one on standby). They are equipped with recirculation lines for protectionagainst minimum flow.

II.3.2.6.Refrigerant 07-MC15

The produced propane refrigerant 07-MC15 is a kettle-type heat exchanger with tube and shell configuration, using liquid propane as the refrigerant. The exchanger receives refrigerant from the propane refrigeration unit (20-MC06) and cools the propane to -36ºC before being sent to:

- Propane storage tank 76-MF01.
- LNG reinjection (if needed).
- Both propane refrigeration loops (Unit 16 and Unit 20) as a supplement.

II.3.3.System Description

The depropanizer 07-MD04 receives its feed from the deethanizer 07-MD02. The liquid hydrocarbons in the deethanizer, initially at a pressure of 24 bar and a temperature of approximately 101ºC, are depressurized to 13 bar and around 74ºC before being introduced above the 16th tray of the depropanizer. Equipped with 37 tight trays, the depropanizer enhances the separation of light and heavy components. Reboiling is necessary to restrict the propane content at the bottom of the depropanizer, ensuring compliance with the quality specifications of the butane leaving the debutanizer's top.

The reboiler, 07-MC04, of the depropanizer is a kettle-type heat exchanger with tube and shell configuration, using hot oil for heating. It includes a draw-off tray at the bottom to direct liquid to the depropanizer reboiler. Inside the reboiler, a baffle facilitates liquid flooding of the tubes while allowing ample space for vapor circulation to the column. Vapors enter the column below the draw-off tray and rises through the tight trays. Liquid flows by gravity over the baffle at the bottom of the reboiler shell and returns to the bottom of the depropanizer.

To optimize component separation, the temperature at the depropanizer's bottom is controlled at 98ºC by adjusting the hot oil flow rate to the reboiler 07-MC04.

The overhead vapors from the depropanizer are condensed by the air-cooled condenser 07-MC05 and collected in the reflux drum of the depropanizer 07-MD05. Some of the liquids from this reflux drum are sent to the column by pumps 07-MJ02/A to serve as reflux, while the rest are directed to the propane cooler 07-MC15. The reflux flow rate ensures high-quality overhead product, meeting specified requirements.

The propane cooler 07-MC15, also a kettle-type, utilizes liquid propane from the external propane refrigeration system as a refrigerant. Propane liquids are cooled to -36ºC in this exchanger before being routed either to the LPG reinjection drum 07-MD08 or to the propane storage tank 76-MF01. Additional propane supply for Unit 16 refrigeration and the external Unit 20 propane refrigeration is sourced from downstream of the propane cooler.

The system includes provisions for sending off-spec propane from the upstream of the propane refrigeration unit to the LNG start-up/Bupro pressure relief ball (65-MD05) for disposal, and for sending off-spec propane from the downstream of the propane refrigeration unit to the off-spec propane storage tank (67-B-03) of the existing units. **[3]**

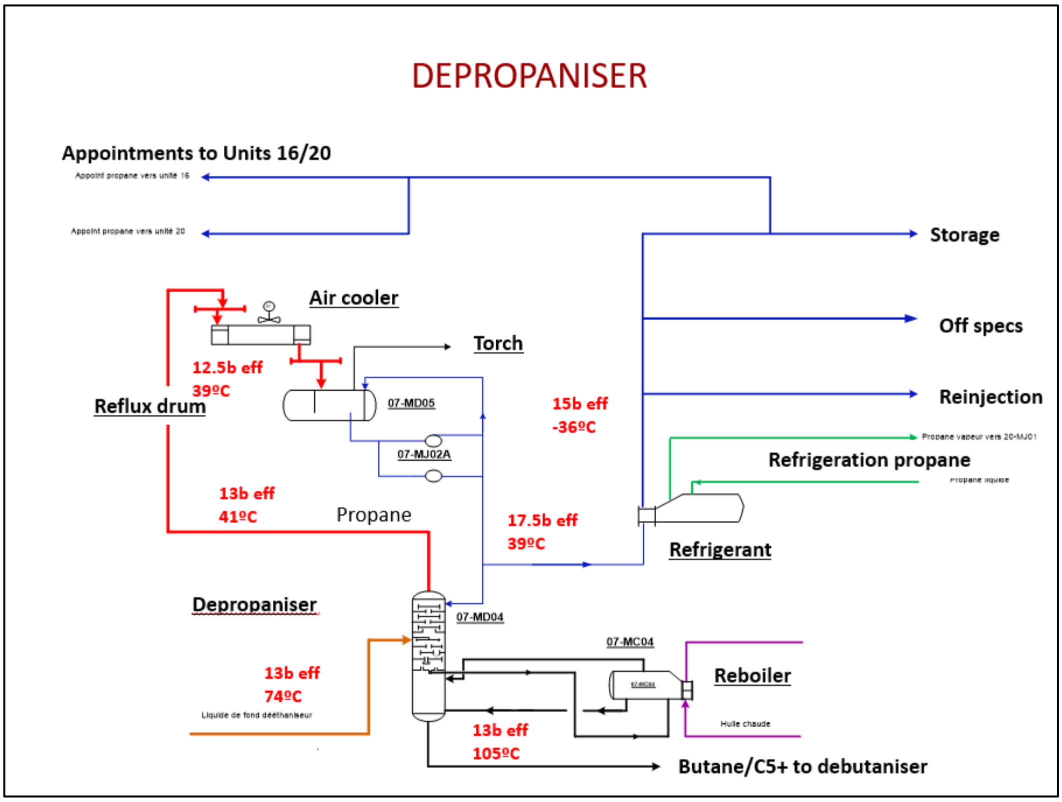


Figure II.6 Descriptive diagram of the Depropanizer

II.4. Conclusion

In this section, we have provided a comprehensive overview of the fractionation unit, highlighting its crucial role in separating natural gas into valuable components. Each product derived from this process has specific uses: some feed into the MR and PR cooling loops, others are directed towards the fuel gas unit, and the rest are prepared for storage and shipping. We also detailed the depropanizer section, describing the function of each piece of equipment involved. Understanding the fractionation unit's operations and outputs underscores its importance in the LNG industry's overall efficiency and product quality

CHAPTER.III. Advanced Procces Control System in GL1K Complex

III.1. Introduction

Honeywell Advanced Solutions has implemented an innovative solution for advanced control and online optimization applications, such as the one implemented in the GL1K complex in Skikda. The implementation of Advanced Process Control (APC) technologies plays a crucial role in optimizing various units within the facility. The APC controllers are strategically implemented across key units such as Acid Gas Removal, Fractionation Units, Liquefaction and Refrigeration, Scrubber, and Mixed Refrigerant Composition Controller. These controllers are designed to enhance operational efficiency, maximize production throughput, manage product specifications, and ensure the optimal performance of critical processes within the LNG complex.

APC is software installed at the Distributed Control System (DCS) to better manage the complex using advanced control, it adopts a layered approach to process optimization, facilitating real-time performance monitoring, life cycle maintenance, and advanced analytics, and to improve the operation of PID control loops. **[8] [6]**

III.2. Main objectives of this application

The objectives of implementing "Advanced Process Control" (APC) are:

Minimize self-consumption: By optimizing the plant's operations, APC can reduce the energy consumption of the plant, leading to cost savings and a more efficient use of resources.

Increase Production: APC can help increase production by optimizing the process conditions, improving the yield of high-value products, and reducing downtime.

Ensure and manage product quality: APC can ensure that the products meet the required specifications by monitoring and controlling the process conditions in real-time.

Improve the stability of the plant's operation: APC can help stabilize the plant's operation by identifying and addressing potential issues before they become major problems.

Reduce the energy consumption of the plant: APC can help reduce the energy consumption of the plant by optimizing the process conditions and identifying opportunities for energy savings.

Maximize profits: APC can help maximize profits by optimizing the production process, reducing costs, and improving the yield of high-value products.

Increase the commercial propane yield (ambient and refrigerated): APC can help increase the commercial propane yield by optimizing the separation and purification processes, leading to higher yields of high-value products.

By achieving these objectives, APC can help improve the overall efficiency and profitability of the plant, while also reducing its environmental impact. **[8]**

III.3. APC General structure

9 APC controller are implemented:

- UNIT 12, Acid Gas Removal: (12AGRU)
- UNIT 07, Fractionation Units: (07DEC1,07DEC2,07DEC3,07DEC4,07DEC5)
- UNIT 15-16, Liquefaction and Refrigeration: (1516LR 16MR)

A simplified overview of the GL1K complex is presented below. The APC applications cover in the units enlightened in yellow

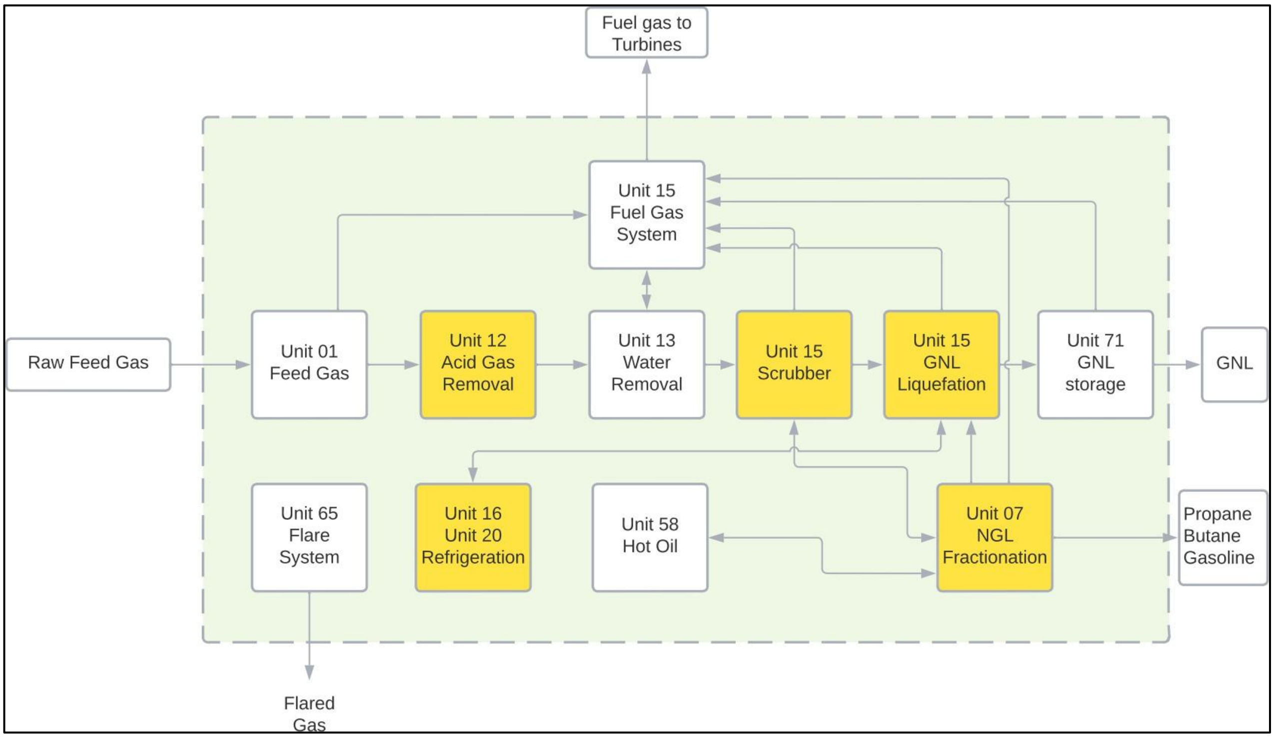


Figure III.1 APC Applications In GL1K Complex

The GL1K complex utilizes eight APC controllers to optimize operations and enhance production efficiency. These controllers manage critical process parameters in real-time using CVs, MVs, and DVs. Inferential models aid in accurate process prediction for precise control. The APC structure, integrated with advanced strategies and real-time monitoring, improves operational efficiency, reduces energy consumption, and ensures effective production. Honeywell's software maximizes profitability by optimizing production, managing specifications, and meeting constraints. Profit Controllers are designed for the LNG train to ensure safe and reliable operations while meeting optimization goals.

III.4. Profit controller

The applications of Advanced Process Control (APC) will be implemented using Honeywell's Profit Suite software, which is based on the "Multivariable Predictive Robust Control Technology". This application of multivariable control allows for the resolution of all control problems and the optimization of highly interactive industrial processes. To control a process, the Profit Controller (RMPCT) uses controlled variables (CVs), manipulated variables

(MVs), and disturbance variables (DVs) as inputs/outputs for control. These may include variables such as temperature, pressure, flow rates, and composition. **[7]**

III.4.1. Robust Multivariable Predictive Control Technology

RMPCT is an acronym standing for Robust Multivariable Predictive Control Technology. This control technique uses a model matrix to represent the dynamic relations between process variables and estimate the future plant behavior. It considers the effect of all involved variables and enables the optimization of process operations while respecting all constraints.Unlike traditional PID controllers, Which rely entirely on feedback from the process.

RMPCT controllers use models to predict the future behavior of the unit, allowing for more effective and efficient control. This technology operates at the supervisory control level, above the basic regulatory control scheme, and is particularly suited for complex processes where multiple variables are involved.

RMPCT is a model-based, multivariable, constrained, predictive control technology that differs significantly from standard PID controllers. It uses a model matrix to represent the dynamic relations between process variables and solve an optimization problem at each cycle to drive the plant in the most profitable region while respecting all constraints.

RMPCT's predictive capabilities enable it to reject disturbances faster than a PID loop, as shown in Figure 1. The controller updates its prediction of the plant's future behavior every 30 seconds for the GL1K APC applications, ensuring consistency between the Profit Controller projection and the actual process behavior.

III.4.2. Control variables (CVs)

A controlled variable is a variable that changes in response to a change in a manipulated or disturbance variable. Product qualities are examples of controlled variables. These variables are sometimes called dependent variables because their values depend on other set points and/or outputs. It is not possible to directly change the value of a dependent variable. To change a dependent variable value, moving the manipulated variables that affect it is necessary.

III.4.3. Manupilated variables (MVs)

A manipulated variable is a “handle” used to control the process. The manipulated variables are set points or outputs of existing PID controllers. For example, feed, product draw rate, or temperature controllers can be manipulated variables. The RMPCT controller will move these set points (or outputs) to meet the control targets that have been set. These variables are sometimes referred to as independent variables because they can be set regardless of other set points and outputs in the process.

Examples of MVs:

- Reflux flow rate.
- Reboil flow rate.
- Column head pressure.

- Feed flow rate

III.4.4. Disturbance variables (DVs)

A disturbance (or feed-forward) variable has an impact on the process but cannot be adjusted by the RMPCT controller. A typical disturbance variable is the ambient temperature, which affects many process variables but cannot be changed by Profit Controller. The controller can only “see” the change in temperature and make adjustments in other manipulated variables to correct for the expected disturbance. These variables, as the MVs, are also independent because they cannot be influenced by anything.

Examples of DVs:

- Ambient temperature.
- Number of air coolers.
- Feed composition (Analyzer).

To choose the variables in the model and to define their type (CV, MV or DV), you need to:

- Know on-site installations and changing operating variables.
- Have experience in the field and rely on the competence of the engineer.
- Work as a team to reduce or avoid error.

III.4.5. Controller execution cycle

The PID controllers that perform the basic regulatory control execute their control action typically every second or even every 0.5 seconds. The RMPCT controller has instead an execution cycle that is 30 seconds for GL1K Profit Controllers. Between two executions the controllers are inactive and every variation or disturbance on the plant, or a modification of the control strategy, is not considered until the next execution.

III.4.6. Time to steady state

A controller’s time to steady state represents the time necessary for the plant to reach a new steady condition after a variation has occurred in an independent (manipulated or disturbance) variable. This value represents the time window in the past or the future that can affect the strategy of the controller. For instance, a controller with a time to steady state of 3 hours will use the variations of the independent variables that occurred in the last 3 hours to calculate its predictions: the predictions will extend for 3 hours into the future.

III.4.7. steady-state value

The steady-state value is calculated at every controller execution for every manipulated and controlled variable. This value is chosen to drive the plant to a more profitable operating point in a way that respects all the quality specifications and the other process constraints; the controller will try to move all the manipulated variables to ensure that each controlled variable will reach the calculated target.

III.4.8. RMPCT versus Standard Regulatory Control

An RMPCT controller differs significantly from standard PID controllers in several ways. Unlike standard PID controllers, which rely solely on feedback from the process, RMPCT controllers use models to predict the future behavior of the unit. This prediction capability allows the controller to plan a series of future moves for each manipulated variable, which will keep all controlled variables on target. A standard PID controller has one controlled variable (CV) and one manipulated variable (MV). For example, a flow controller has a measured flow rate as the CV and the output to the valve as the MV. A multivariable controller, on the other hand, can have two or more controlled and/or manipulated variables. The controlled variables can include items like draw temperature, tray temperature, column differential pressure, controller valve outputs, or product qualities. The manipulated variables can be set points of existing PID controllers or valve positions sent to existing field controllers. RMPCT controllers use a matrix of models to predict the future behavior of the unit.

These models describe the dynamic relationships between controlled and manipulated variables, including the magnitude of the change in the CV when a unit change has been imposed in the considered MV (gain) and the time it takes for the CV to reach this value (settle time). For example, if the 07TIC1027. SP is increased by one degree, the bottom pressure compensated temperature will increase about 1 degree as well, and it will take about 15 minutes. RMPCT controllers are also constraint controllers, ensuring that controlled variables remain within a specified range. This is particularly important in processes where maintaining specific conditions or equipment limits is crucial. **[6]**

III.4.9. How does RMPCT work

Each RMPCT controller executes at fixed intervals (every 30 seconds for GL1K application). As part of this execution, the controller performs the following:

- Read the current process values for all variables within the controller.
- Make a prediction of future CV behavior.
- Calculate Steady State Target: the MV settings that satisfy all variable limits, minimizing the costs.
- Determine a path of future moves to drive each MVs to the steady state target, minimizing CV error.

Based on changes in MVs and DVs occurred over the past time to steady state, if MVs and DVs will remain constant in the future, and using the model response curves, the controller will predict the behavior of the CVs over the next time to steady state, as shown in the following figure.

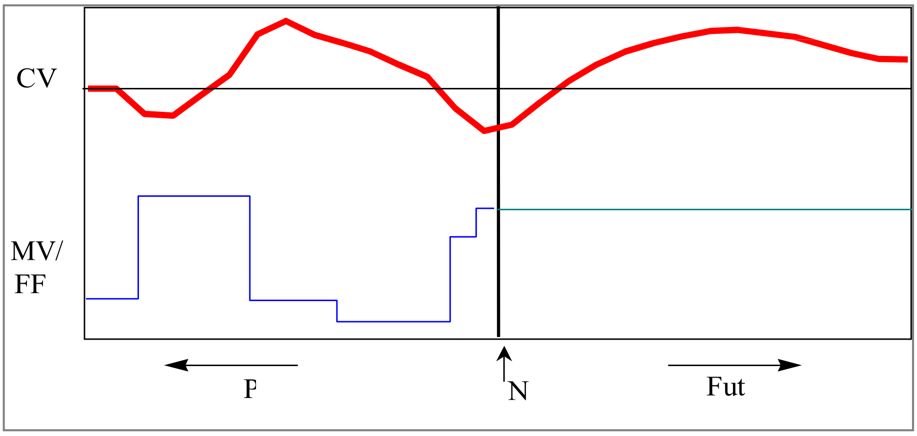


Figure III.2 RMPCT controller prediction calculations.

III.4.9.1. Linear programming

RMPCT uses a linear programming (LP) algorithm to determine the “best” combination of MV values that must be implemented to control the CVs. The best combination of MVs values is the combination that minimizes the overall cost, i.e. maximum profit.The LP’s priority is to find a combination of MVs moves that meet all the CVs and MVs limits. In general, there is several possible combinations that will meet all the limits, then the controller will choose the most profitable combination. If, however, the LP cannot keep all the CVs within their limits, it will allow the least important CV(s) to violate their limit(s). The controller will never allow an MV to violate its limits. The LP targets for all the MVs are displayed on the Operator View display in the Steady State column.

III.4.9.2. Move calculation

RMPCT predicts the future behavior of all the CVs within the controller. These predictions contain the effects of all the actions on the manipulated variables and the disturbances thatoccurred in the past Time to Steady State. These predictions are called “open loop predictions” because they represent what would happen if the controller wouldn’t make any further control action in the future (and if no new disturbances would impact the plant). Moving manipulated variables aims to achieve the desired LP solution and minimize the error on the controlled variables. The error on the controlled variables is defined as the difference between the calculated LP target and the predicted value.

RMPCT will select the future moves to create the curve that best fits the mirror image of the open-loop prediction about the set point so that the resulting effect of past and future actions will drive the controlled variable within limits.

RMPCT calculates 10 future moves for each manipulated variable at every execution time. Once the move plan has been calculated, only the first calculated control action is implemented for each MV; then, at the next cycle, the values of the currently controlled variables are compared with the prediction to consider any unmeasured disturbances; then the prediction is updated, feed-back action, and the move plan is recalculated.

III.4.9.3. Controller variables

The APC MVs, CVs, and DVs are reported below for each one of the controllers. For each MV, the strategy adopted by the controller is also reported with the following meaning:

- MAX: the MV will be maximized until a corresponding constraint is activated.
- MIN: the MV will be minimized until a corresponding constraint is activated.
- CONTROL: the MV will move when the constraint is activated.

MVs costs are the parameters used to define the controller strategy: if a manipulated variable has a negative cost, it is maximized whenever possible; on the other hand, if the cost is positive, it is minimized. In the special case of a manipulated variable with a cost of 0, it is defined as “minimum movement” MV: it is moved only when necessary to respect the CVs limits, with no optimization purpose.

The MVs follow the strategy when no other constraint on CVs is present: when the maximum or minimum limit on one of the MVs is reached, the optimization strategy for that variable stop, and the limit on the MV is never violated.

When one of the CV is violating its minimum or maximum limit, the optimization strategy on the related MVs is stopped if necessary, and these variables are moved to bring back that CV inside the limits. **[6]**

III.4.9.3.1. Acid Gas Removal Unit (U12AGRU) Controller Structure

The lists of the MVs, CVs, DVs of the Controller are reported in the following Tables:

Table III.1 List of CVs U12AGRU

No.	Tag	Description	Comments
CV1	13AI1008.PV	CO2 content sweet gas	APC will maintain this CV below its upper limit
CV2	LAM_FEED_R.PV	Lean amine to feed ratio	Calculated CV
CV3	12PDI1009.PV	Absorber delta pressure	
CV4	12PDI1240.PV	Flash column delta pressure	
CV5	LAM_FD_DT.PV	DT lean amine vs feed gas	Calculated CV
CV6	12TIC1043.PV	Regeneration temperature	For amine regeneration
CV7	12PDI1050.PV	Regenerator delta pressure	To avoid flooding
CV8	12TI1213.PV	Regenerator bottom temperature	

Table III.2 List of MVs U12AGRU

No.	Tag	Description	Direction	Comments
MV1	12FIC1117.SP	Lean amine to 12MD21	Control	To keep CV1 and CV2 within the limits
MV2	12TIC1102.SP	Lean amine temperature	Control	To control LAM_FD_DT
MV3	12FIC1070.OP	Regenerator hot oil flow	Control	To control the regenerator top temperature

Table III.3 List of DVs U12AGRU

No.	Tag	Description	Comments
DV1	01AI3005C.PV	Feed CO2 content	To anticipate effects of CO2 changes
DV2	12FI1023.PV	Feed to the unit	To anticipate effects of feed rate changes
DV3	00TI2004.PV	Ambient temperature	To adapt the unit to ambient temperature changes

III.4.9.3.2. Main Cryogenic Heat Exchanger Section (U1516LR) Controller structure

The lists of the MVs, CVs, DVs of the Controller are reported in the following Tables:

Table III.4 List of CVs U1516LR

No.	Tag	Description	Comments
CV1	15FIC1057_TARGET	Target GNL target production	
CV2	15FIC1057.OP	LGN Flow valve	
CV3	16JI1371.PV	16-MJ04 (propane compressor) VSDS power	
CV4	15TIC1064.OP	Scrubber temperature OP	
CV5	15PDI1016.PV	Scrubber pressure drop	
CV6	15TI1124.PV	MCHE 1st pass temperature	
CV7	C2SCRUB_INF	Ethane in 2nd pass/15AI1018E.PV	Inferential
CV8	C3SCRUB_INF	Propane in 2nd pass/15AI1018F.PV	Inferential
CV9	15TIC1056.PV	GNL MCHE outlet Temperature	
CV10	16CRIC1035.PV	MR Compression Ratio	
CV11	15TI1000.PV	LMR before mixing cold box	
CV12	15TI1441.PV	cold box outlet to 15MJ05 compressor	
CV13	16TI1018.PV	Light MR from cold band MCHE	
CV14	16FFIC1037.PV	MR ratio lean heavy	
CV15	16PI1075.PV	Suction Press St1 MR compressor	
CV16	16XI3912.PV	MR compressor estimated inlet density	

Table III.5 List of MVs U1516LR

No.	Tag	Description	Direction	Comments
MV1	15TIC1064.SP	Scrubber temperature	Maximize	To maximize product yield
MV2	15FIC1057.SP	LGN Flow	Control	LGN flow at target
MV3	16HIC1038.OP	Warm MR flow	Minimize	to decrease energy consumption
MV4	16UY1035A.OP	Cold MR flow	Minimize	to decrease energy consumption
MV5	16HIC1358.OP	IGV MR Compressor	Minimize	to decrease energy consumption
MV6	15FIC1001.OP	End Flash Gas cold box outlet	Maximize	to increase energy recovery

Table III.6 List of DVs U1516LR

No.	Tag	Description	Comments

DV1	00TI2004.PV	Ambient Temperature	To reject disturbances
DV2	16MI4778.PV	Humidity	To reject disturbances

III.4.9.3.3. Unit 07 Demethanizer (07DEC1) controller structure

The lists of the MVs, CVs, DVs of the Controller are reported in the following Tables:

Table III.7 List of CVs 07DEC1

No.	Tag	Description	Comments
CV1	07PIC1337.OP	Top pressure valve	To avoid saturation
CV2	C2TOP_INF.PV	C2 in top inferential	Updated with 07AI1216C.PV
CV3	07TI1411PCT.PV	Demetha tray7 pressure comp. Temp.	Calculated CV
CV4	C1BOT_INF.PV	C1 in top deetha inferential	Updated with lab 07SC002
CV5	07TI1008PCT.PV	DC1 bottom pressure comp. temp.	Calculated CV

Table III.8 List of MVs 07DEC1

No.	Tag	Description	Direction	Comments
MV1	07PIC1337.SP	top demetha pressure	Minimize	APC will minimize the pressure based on the C2 content in product.
MV2	07TIC1007.SP	demetha bottom temp	Minimize	APC will maximize to increase C1 and C2 recovery

Table III.9 List of DVs 07DEC1

No.	Tag	Description	Comments
DV1	07FIC1471.PV	Feed to the column	To anticipate effect of changes
DV2	00TI2004.PV	Ambient Temperature	Compensate the dynamic effect
DV3	58TIC1004.SP	Hot Oil Temperature	Compensate the dynamic effect

III.4.9.3.4. Unit 07 Deethanizer (07DEC2) controller structure

The lists of the MVs, CVs, DVs of the Controller are reported in the following Tables:

Table III.10 List of CVs 07DEC2

No.	Tag	Description	Comments
CV1	07PIC1022A.OP	Top pressure valve	
CV2	DC2_REF_R.PV	Reflux to feed ratio deethaniser	Calculated CV
CV3	07TI1402PCT.PV	Tray 8 PCT	Calculated CV
CV4	C2BOT_INF.PV	Ethane in top DC3	Inferential
CV5	07TI1031PCT.PV	Bottom press. compensated temp.	Calculated CV

Table III.11 List of MVs 07DEC2

No.	Tag	Description	Direction	Comments
MV1	07PIC1022A.SP	Top deetha pressure	Control	APC will manipulate the pressure to control the pressure valve.
MV2	07TIC1027.SP	Deetha bottom temp	Minimize	APC will minimize based on the C2 concentration in top depropa.

Table III.12 List of DVs 07DEC2

No.	Tag	Description	Comments
DV1	07FC1013.SP	Feed to the column	To anticipate effect of feed changes
DV2	00TI2004.PV	Ambient Temperature	Compensate the dynamic effect
DV3	58TIC1017.SP	Hot Oil Temperature	Compensate the dynamic effect

III.4.9.3.5. Unit 07 Depropanizer (07DEC3) controller structure

The lists of the MVs, CVs, DVs of the Controller are reported in the following Tables:

Table III.13 List of CVs 07DEC3

No.	Tag	Description	Comments
CV1	07PIC1061A.OP	Top pressure valve	To avoid saturation
CV2	C4TOP_INF.PV	Butane in propane inferential	Updated with 07AI3062C.PV
CV3	07TI1069PCT.PV	Tray 14 pressure compensated temperature	Calculated CV
CV4	07PDI1319.PV	Pressure drop rectifying section	To avoid flooding
CV5	C3BOT_INF.PV	Propane in top debuta inferential	Updated with 07AI1119A.PV
CV6	07TI1060PCT.PV	Bottom pressure compensated temperature	Calculated CV

Table III.14 List of MVs 07DEC3

No.	Tag	Description	Direction	Comments
MV1	07PIC1061A.SP	Top depropa pressure	Control	APC will manipulate the pressure to avoid saturation of the valve.
MV2	07FIC1086.SP	Reflux flow	Minimize	APC will minimize the reflux flow based on the C4 content in top product.
MV3	07TIC1375.SP	Depropanizer bottom temp	Minimize	APC will minimize temperature to reduce energy consumption

Table III.15 List of DVs 07DEC3

No.	Tag	Description	Comments
DV1	07FIC1035.SP	Feed to the column	To anticipate effect of feed changes
DV2	00TI2004.PV	Ambient Temperature	To adapt the unit
DV3	58TIC1017.SP	Hot Oil Temperature	Compensate the dynamic effect

III.4.9.3.6. Unit 07 Debutanizer (07DEC4) controller structure

The lists of the MVs, CVs, DVs of the Controller are reported in the following Tables:

Table III.16 List of CVs 07DEC4

No.	Tag	Description	Comments
CV1	07PIC1098A.OP	Top pressure valve	To avoid saturation
CV2	07PDI1278.PV	Pressure drop rectifying section	To avoid flooding
CV3	07AI1119C.PV	Pentane in butane	To avoid C5 loss in butane
CV4	DC4_REF_R.PV(1)	Reflux to feed ratio debutanizer	Calculated CV
CV5	DC4_DUTY_R.PV(2)	Hot oil to feed ratio debutanizer	Calculated CV
CV6	C4BOT_INF.PV	Butane in top deisopentanizer	Inferential updated with 07SC015
CV7	07TI1108PCT.PV	Bottom PCT	Calculated CV

Table III.17 List of MVs 07DEC4

No.	Tag	Description	Direction	Comments
MV1	07PIC1098A.SP	Top pressure	Control	To control the pressure valve.
MV2	07FIC1120.SP	Reflux flow	Minimize	APC will minimize the reflux flow
MV3	07TIC1100.SP	Tray 31 temperature	Minimize	APC will minimize the temperature

Table III.18 List of DVs 07DEC4

No.	Tag	Description	Comments
DV1	07FIC1073.SP	Feed to the column	To anticipate effect of feed
DV2	00TI2004.PV	Ambient Temperature	To adapt the unit to ambient temperature
DV3	58TIC1017.SP	Hot Oil Temperature	Compensate the dynamic effect

III.4.9.3.7. Unit 07 Deispentanizer (07DEC5) controller structure

The lists of the MVs, CVs, DVs of the Controller are reported in the following Tables:

Table III.19 : List of CVs 07DEC5

No.	Tag	Description	Comments
CV1	DC5_REF_R.PV(1)	Reflux to feed ratio deisopentanizer	Calculated CV
CV2	TVRBOT_INF.PV	TVR bottom product	Inferential
CV3	07TI1171PCT.PV	Bottom pressure compensated temp.	Calculated CV
CV4	DC5_DT_REB.PV(2)	Reboiler delta temperature	Calculated CV

Table III.20 : List of MVs 07DEC5

No.	Tag	Description	Direction	Comments
MV1	07FIC1192.SP	Reflux flow	Control	Minimized based on the reflux to feed ratio.
MV2	07TIC1174.SP	Tray 46 temperature	Minimize	Minimized based on the bottom specification.

Table III.21 : List of DVs 07DEC5

No.	Tag	Description	Comments
DV1	07FIC1168.SP	Feed to the column	To anticipate effect of feed changes
DV2	00TI2004.PV	Ambient Temperature	To adapt the unit to ambient temperature
DV3	58TIC1017.SP	Hot Oil Temperature	Compensate the dynamic effect

III.4.9.3.8. Unit 16 Mixed refrigerant composition (U16MR) controller structure

The lists of the MVs, CVs, DVs of the Controller are reported in the following Tables:

Table III.22 : List of CVs U16MR

No.	Tag	Description	Comments
CV1	16AI2465A.PV	MR nitrogen content	MR quality control
CV2	16AI2465B.PV	MR methane content	MR quality control

CV3	16AI2465C.PV	MR ethane content	MR quality control
CV4	16AI2465D.PV	MR propane content	MR quality control

Table III.23 : List of MVs U16MR

No.	Tag	Description	Direction	Comments
MV1	16HIC1306.OP	Methane to MR	Control	MR quality control
MV2	16HIC1310.OP	Ethane to MR	Control	MR quality control
MV3	16HIC1304.OP	Nitrogen to MR	Control	MR quality control
MV4	16HIC1308.OP	Propane to MR	Control	MR quality control

III.4.9.4. Custom calculations

In this section, the calculated variables used by the APC are described. **[6]**

III.4.9.4.1. U12AGRU Controller Calculated Variables

Calculated variables for the U12AGRU controller are reported in the following table.

Table III.24 : List of Calculated Variables U12AGRU

No.	Tag	Description	Comments
CV2	LAM_FEED_R.PV	Lean amine to feed ratio	Calculated CV
CV5	LAM_FD_DT.PV	DT lean amine vs feed gas	Calculated CV

LAM_FEED_R.PV : The lean amine-to-feed ratio is calculated as follows:

$$12\mathrm{LAM\_FEED\_R}.\mathbf{PV} = \frac{12\mathrm{FIC1117}.\mathbf{SP}}{12\mathrm{FI1023}.\mathbf{PV}}$$

LAM_FD_DT.PV : The difference in temperature between lean amine and feed gas is calculated as follows:

$$\mathrm{LAM\_FD\_DT}.\mathbf{PV} = 12\mathrm{TIC1102.SP} - 12\mathrm{TI1025.PV}$$

III.4.9.4.2. 07DEC1 Controller Calculated Variables

Calculated variables for the 07DEC1 controller are reported in the following table.

Table III.25 : List of Calculated Variables 07DEC1

No.	Tag	Description	Comments
CV3	07TI1411PCT.PV	Demethanizer tray 7 pressure compensated temperature	Calculated CV
CV5	07TI1008PCT.PV	Demethanizer bottom pressure compensated temperature	Calculated CV

07TI1411PCT.PV : The tray 7 pressure compensated temperature is calculated as follow:

$$07\mathrm{TI1411PCT}.\mathbf{PV} = 07\mathrm{TI1411}.\mathbf{PV} - \mathbf{23.36} \times \mathbf{ln}\left(\frac{07\mathrm{PIC1337}.\mathbf{PV}}{34}\right)$$

07TI1008PCT.PV : The bottom pressure compensated temperature is calculated as follow:

$$07\mathrm{TI008PCT}.\mathbf{PV} = \ 07\mathrm{TI0008}.\mathbf{PV} - \mathbf{72.7} \times \mathbf{ln}\left(\frac{07\mathrm{PIC1337}.\mathbf{PV}}{34}\right)$$

III.4.9.4.3. 07DEC2 Controller Calculated Variables

Calculated variables for the 07DEC2 controller are reported in the following table

Table III.26 : List of Calculated Variables 07DEC2

No.	Tag	Description	Comments
CV2	DC2_REF_R.PV	Reflux to feed ratio deethaniser	Calculated CV
CV3	07TI1402PCT.PV	Tray 8 PCT	Calculated CV
CV5	07TI1031PCT.PV	bottom PCT	Calculated CV

DC2_REF_R.PV : The Reflux to feed ratio of deethaniser is calculated as follows:

$$\mathrm{DC2\_REF\_R}.\mathbf{PV} = \frac{07\mathrm{FIC1047}.\mathbf{SP}}{07\mathrm{FIC1013}.\mathbf{SP}}$$

07TI1402PCT.PV : The tray #8 pressure compensated temperature is calculated as follows:

$$07\mathrm{TI1402PCT}.\mathbf{PV} = 07\mathrm{TI1402}.\mathbf{PV} - \mathbf{41.2} \times \mathbf{ln}\left(\frac{07\mathrm{PIC1022A}.\mathbf{PV} + 1}{24.5}\right)$$

07TI1031PCT.PV : The bottom pressure compensated temperature is calculated as follow:

$$07\mathrm{TI1031PCT}.\mathbf{PV} = 07\mathrm{TI1031}.\mathbf{PV} - \mathbf{27.7} \times \mathbf{ln}\left(\frac{07\mathrm{PIC1022A}.\mathbf{PV} + 1}{24.5}\right)$$

III.4.9.4.4. 07DEC3 Controller Calculated Variables

Calculated variables for the 07DEC3 controller are reported in the following table.

Table III.27 : List of Calculated Variables 07DEC3

No.	Tag	Description	Comments
CV3	07TI1069PCT.PV	Tray 14 pressure compensated temperature	Calculated CV
CV6	07TI1060PCT.PV	Bottom pressure compensated temperature	Calculated CV

07TI1069PCT.PV : The tray 14 pressure compensated temperature is calculated as follows:

$$07\mathrm{TI1069PCT}.\mathbf{PV} \ = 07\mathrm{TI1069}.\mathbf{PV} - 42.1 \times \ln\left(\frac{07\mathrm{PIC1061A}.\mathbf{PV}}{12.8}\right)$$

07TI1060PCT.PV : The bottom pressure compensated temperature is calculated as follows:

$$07\mathrm{TI1060PCT}.\mathbf{PV} = 07\mathrm{TI1060}.\mathbf{PV} - 35.0 \ \times \ln\left(\frac{07\mathrm{PIC1061A}.\mathbf{PV} + 1}{14}\right)$$

III.4.9.4.5. 07DEC4 Controller Calculated Variables

Calculated variables for the 07DEC4 controller are reported in the following table.

Table III.28 : List of Calculated Variables 07DEC3

No.	Tag	Description	Comments

CV4	DC4_REF_R.PV	Reflux-to-feed ratio	Calculated CV
CV5	DC4_DUTY_R.PV	Reboiler flow to feed ratio	Calculated CV
CV7	07TI1108PCT.PV	Bottom PCT	Calculated CV

DC4_REF_R.PV : The reflux to feed ratio is calculated as follow:

$$\text{REF\_FEED\_R}\,.\,\mathbf{PV} = \frac{\text{07FIC1120.}\mathbf{SP}}{\text{07FIC1073.}\mathbf{SP}}$$

DUTY_FEED_R : The hot oil flow to feed ratio is calculated as follow:

$$\text{DUTY\_FEED\_R}\,.\,\mathbf{PV} = \frac{\text{07FIC1103.}\mathbf{SP}}{\text{07FIC1073.}\mathbf{SP}}$$

07TI1108PCT.PV : The bottom pressure compensated temperature is calculated as follow:

$$\text{07TI1108PCT.}\mathbf{PV} = \text{07TI1108.}\mathbf{PV} - 35.22 \times \ln\left(\frac{\text{07PIC1098A. PIDA.}\mathbf{PV} + 1}{4.5}\right)$$

III.4.9.4.6. 07DEC5 Controller Calculated Variables

Calculated variables for the 07DEC5 controller are reported in the following table.

Table III.29 : List of Calculated Variables 07DEC5

No.	Tag	Description	Comments
CV1	DC5_REF_R.PV	Reflux to feed ratio deisopentanizer	Calculated CV
CV3	07TI1171PCT.PV	Bottom pressure compensated temp.	Calculated CV
CV4	DC5_DT_REB.PV	Reboiler delta temperature	Calculated CV

DC5_REF_R.PV : The reflux-to-feed ratio is calculated as follows:

$$\text{DC5\_REF\_R}\,.\,\mathbf{PV} = \frac{\text{07FIC1192.}\mathbf{SP}}{\text{07FIC1168.}\mathbf{SP}}$$

07TI1171PCT.PV : The bottom pressure compensated temperature is calculated as follows:

$$\text{07TI1171PCT.}\mathbf{PV} = \text{07TI1171.}\mathbf{PV} - 60.8 \times \ln\left(\frac{\text{07PIC1177A.}\mathbf{PV} + 1}{2.2}\right)$$

DC5_DT_REB.PV : The bottom delta temperature is calculated as follows:

$$\text{DC5\_DT\_REB}\,.\,\mathbf{PV} = \text{07TI1171.}\mathbf{PV} - \text{07TIC1174.}\mathbf{SP}$$

III.5. Operative and Control Objectives

III.5.1. Overview

The main parameters that determine the general operating strategy of the controller are the “costs” assigned to the manipulated variables and the “EU give-ups” specified for the upper and lower limits of the controlled variables. An overview of these will be given in the next section before the strategy of the controller is discussed in detail.

The costs assigned to the manipulated variables set the controller's strategy. For some variables, the sign of the cost parameter actively determines whether the controller will try to maximize (negative cost) or minimize (positive cost) the variable. The controller will stop pursuing thismaximization/minimization strategy specified by the MV costs in two circumstances:

- When the manipulated variable reaches one of its operating limits.
- When one of the controlled variables dependent on the MV reaches one of the operating limits.

III.5.2. Main control strategy

III.5.2.1. Acid Gas Removal Unit (U12AGRU)

APC will manipulate the lean amine to control CO2 in sweet gas.

APC will manipulate the regenerator hot oil valve to control the regenerator top temperature.

APC with its continuous action and monitoring will improve the overall stability of the unit.

III.5.2.2. Main Cryogenic Heat Exchanger Section (U1516LR)

APC will minimize the MR refrigerant flows against the GNL outlet temperature. This strategy is expected to decrease the fuel gas consumption for mixed refrigerant compression together with the fuel gas consumption of the propane refrigeration cycle used to cool down the mixed refrigerant. APC will keep the MCHE outlet GNL flow at target.

III.5.2.3. Unit 07 Demethanizer (07DEC1)

APC will minimize the demethanizer pressure and temperature against the methane specification in top deethanizer product and the heavier specification in top demethanizer product. The top demethanizer product is sent back to the BOG recycle compressor and fed to the scrubber column.

III.5.2.4. Unit 07 Deethanizer (07DEC2)

APC aims to reduce column energy consumption by minimizing the temperature against product specifications. This is achieved by optimizing the process conditions to reduce the energy required to maintain the desired temperature. By minimizing the temperature, APC helps to reduce the energy consumption of the column, which in turn reduces the overall energy costs and environmental impact of the process.

III.5.2.5. Unit 07 Depropanizer (07DEC3)

APC will optimize the process conditions to minimize the column reflux against product specifications and lower energy consumption by minimizing the temperature. This is achieved by controlling the reflux rate and temperature to minimize the energy required to maintain the desired conditions. By minimizing the reflux and temperature, APC achieves a more efficient energy usage, resulting in lower energy costs and a reduced environmental impact of the process.

III.5.2.6. Unit 07 Debutanizer (07DEC4)

The APC will optimize the unit minimizing column reflux against product specifications and it will minimize temperature to reduce energy consumption.

III.5.2.7. Unit 07 Deispentanizer (07DEC5)

APC will minimize the column ruflux against the reflux to feed ratio and it will minimize the bottom temperature against the bottom product specification.

III.5.2.8. Unit 16 Mixed refrigerant composition (U16MR)

APC will control the mixed refrigerant composition within the limits specified for the LNG production process. This is achieved by optimizing the process conditions to maintain the desired composition of the mixed refrigerant, which is critical for the production of high-quality LNG. By controlling the composition within the specified limits, APC ensures that the LNG meets the required specifications and standards, thereby ensuring the quality and safety of the final product.

III.5.3. Controller Model

The controller model is the key to the functioning of the RMPCT controller. It provides the predictive capabilities of the controller that allows it to avoid constraints violation. The controller model is a matrix of independent/dependent variables models. Not all independent variables (MVs or DVs) have a model with all dependent variables (CVs).In addition, the presence (or absence) of a model curve, combined with the CV limit priorities, determines which MV (or MVs) RMPCT will tend to move to control a CV.

The following tables show the “gain map” for the GL1K controllers and provide a simple overview of the online model. The gain map indicates that if RMPCT needs to control a specific CV, it will only move the MVs that have a model with the said CV (value of “+” or “-”). The gain map also shows the presence of process interaction: a change in any of the indipendents influences more than one controlled variable.

III.5.3.1. Acid Gas Removal Unit (U12AGRU)

The model matrix for the U12AGRU controller is shown below. The models used in Prediction Only are indicated in the table below with a blue-shaded cell. When a model is in prediction only the controller will not move the manipulated variable for the specific controlled variable.

Table III.30 : U12AGRU Gain Matrix

U12AGRU			MV1	MV2	MV3	DV1	DV2	DV3
			12FIC1117.SP	12TIC1102.SP	12FIC1070.OP	01AI3005C.PV	12FI1023.PV	00TI2004.PV
			Lean amine	Lean amine temp.	Reboiler flow	CO2 in Feed	Feed to the unit	Ambient temp.
CV1	13AI1008.PV	CO2 content sweet gas	-0.3		-3	20	0.0001	0.05
CV2	LAM_FEED_R.PV	Lean amine to feed ratio	6.95E-05				-1.70E-05	
CV3	12PDI1009.PV	Absorber delta pressure	0.15					
CV4	12PDI1240.PV	flash column delta pressure	0.10					
CV5	LAM_FD_DT.PV	DT lean amine vs feed gas		1				
CV6	12TIC1043.PV	Regeneration temperature	-0.002		1.54		-0.0003	
CV7	12PDI1050.PV	Regenerator delta pressure			3			
CV8	12TI1213.PV	Regenerator bottom temp.			0.3			0.03

III.5.3.2. U1516LR Gain Matrix

The model matrix for the U1516LR controller is shown below. The models used in Prediction Only are indicated in the table below with a blue-shaded cell. When a model is in prediction only, the controller will not move the manipulated variable for the specific controlled variable. The models in Lowest Use are indicated in the table below with a green shaded cell, meaning that the manipulated variables associated with that model will be used as the last option.

Table III.31 : U1516LR Gain Matrix MV1-MV5

U1516LR			MV1	MV2	MV3	MV4	MV5
			01PIC1017.SP	15TIC1064.SP	15FIC1057.SP	16HIC1038.OP	16UY1035A.OP
			discharge feed compressor	Scrubber temperatur	LGN Flow	Heavy MR flow	Lean MR flow
1	15FIC1057_TARGET	GNL flow from MCHE target			1		
2	15FIC1057.OP	LGN Flow valve	-0.3		0.001355		

3	16JI1371.PV	16-MJ04 VSDS POWER			0.02		
4	15TIC1064.OP	Scrubber temperature valve		1	0.001		
5	15PDI1016.PV	Scrubber pressure drop		5	0.0013		
6	15TI1124.PV	MCHE 1st pass temperature			0.008	-1.8	
7	C2SCRUB_INF	Ethane in 2nd pass/15AI1018E.PV			0.00061	-0.136	
8	C3SCRUB_INF	propane in 2nd pass/15AI1018F.PV			0.0002	-0.07	
9	15TIC1056.PV	LNG MCHE outlet Temperature			0.0014		-0.3
10	16CRIC1035.PV	MR Compression Ratio				-0.15	-0.05
11	15TI1000.PV	LMR before mixing cold box			-0.0080		-0.3
12	15TI1441.PV	Cold box outlet gas to 15MJ05 compressor			-0.0016		-0.06
13	16TI1018.PV	Light MR from cold band MCHE			-0.0010		-0.1
14	16FFIC1037.PV	MR ratio lean heavy				-0.001	0.003
15	16PI1075.PV	Suction Press St1 MR compressor				0.045	0.03
16	16XI3912.PV	MR compressor estimated inlet density				0.045	0.03

Table III.32 : U1516LR Gain Matrix MV6-DV2

U1516LR			MV6	DV1	DV2
			16HIC1358.OP	00TI2004.PV	16MI4778.PV
			IGV MR Compressor BP/MP	Ambient Temperature	Humidity
CV1	15FIC1057_TARGET	GNL flow from MCHE target			
CV2	15FIC1057.OP	LGN Flow valve			
CV3	16JI1371.PV	16-MJ04 VSDS POWER			
CV4	15TIC1064.OP	Scrubber temperature valve			
CV5	15PDI1016.PV	Scrubber pressure drop		0.28	-0.0056
CV6	15TI1124.PV	MCHE 1st pass temperature		0.2	-0.004
CV7	C2SCRUB_INF	Ethane in 2nd pass/15AI1018E.PV		0.015	-0.0003028
CV8	C3SCRUB_INF	propane in 2nd pass/15AI1018F.PV		0.008	-0.00016
CV9	15TIC1056.PV	LNG MCHE outlet Temperature		0.1	-0.002
CV10	16CRIC1035.PV	MR Compression Ratio	0.95		
CV11	15TI1000.PV	LMR before mixing cold box			
CV12	15TI1441.PV	Cold box outlet gas to 15MJ05 compressor		0.09	-0.0018
CV13	16TI1018.PV	Light MR from cold band MCHE			
CV14	16FFIC1037.PV	MR ratio lean heavy			
CV15	16PI1075.PV	Suction Press St1 MR compressor	-0.03		
CV16	16XI3912.PV	MR compressor estimated inlet density	-0.03		

III.5.3.3. Unit 07 Demethanizer (07DEC1)

The model matrix for the 07DEC1 controller is shown below. The models used in Prediction Only are indicated in the table below with a blue-shaded cell. When a model is in prediction only, the controller will not move the manipulated variable for the specific controlled variable.

Table III.33 : 07DEC1 Gain Matrix

07DEC1			MV1	MV2	DV1	DV2	DV3
			07PIC1337.SP	07TIC1007.SP	07FIC1471.PV	00TI2004.PV	58TIC1004.SP
			top demetha pressure	demetha bottom temp	Feed to the column	Ambient Temperature	Hot Oil Temperature
CV1	07PIC1337.OP	Top pressure valve	-7	2	0.025	0	
CV2	C2TOP_INF	C2 in top inferential 07AI1216C	-0.8	0.15	0.0009		
CV3	07TI1411PCT.PV	Demetha top pressure comp. Temp.	-11.43	2.14	0.013		
CV4	C1BOT_INF	C1 in top deetha inferential		-0.3			0
CV5	07TI1008PCT.PV	Demetha bot pressure comp. Temp.		1.22			0

III.5.3.4. Unit 07 Deethanizer (07DEC2)

The model matrix for the 07DEC2 controller is shown below.

Table III.34 :07DEC2 Gain Matrix

U07DC2			MV1	MV2	DV1	DV2	DV3
			07PIC1022A.SP	07TIC1027.SP	07FIC1013.SP	00TI2004.PV	58TIC1017.SP
			top deetha pressure	deetha bottom temp	Feed to the column	Ambient Temperature	Hot Oil Temperature
CV1	07PIC1022A.OP	Top pressure valve	-20	1.35	0.01	0	
CV2	DC2_REF_R.PV	Reflux to feed ratio deethaniser	8		-0.0009		
CV3	07TI1402PCT.PV	Tray 8 PCT	15				
CV4	C2BOT_INF.PV	Ethane in top depropanizer inferential		-0.3			0
CV5	07TI1031PCT.PV	bottom press. compensated temp.		1			0

Table 33 : Unit 07 Depropanizer (07DEC3)

The model matrix for the 07DEC3 controller is shown below. The models used in Prediction Only are indicated in the table below with a blue-shaded cell. When a model is in prediction only, the controller will not move the manipulated variable for the specific controlled variable.

Table III.35 : 07DEC3 Gain Matrix

U07DEC3			MV 1	MV 2	MV 3	DV 2	DV 1	DV
			07PIC1061A.SP	07FIC1086.SP	07TIC1375.SP	07FIC1035.SP	00TI2004.PV	58TIC1017.S
			top depropa	reflux flow	depropanizer bottom temp	Feed to the column	Ambient	Hot Oil
CV1	07PIC1061A.OP	top pressure valve	-35				0.5	
CV2	C4TOP_INF	butane in propane inf 07AI3062C+D		-0.000171	0.0013	3.04E-05		
CV3	07TI1069PCT.PV	tray 14 pressure compensated temp.		-0.018000	0.14	0.0032		
CV4	07PDI1319.PV	pressure drop rectifying section		0.01				
CV5	C3BOT_INF	propane in top debuta inf 07AI1119A			-0.5			0
CV6	07TI1060PCT.PV	bottom pressure compensated temp.			1.2			0

III.5.3.5. Unit 07 Debutanizer (07DEC4)

The model matrix for the 07DEC4 controller is shown below. The models used in Prediction Only are indicated in the table below with a blue-shaded cell. When a model is in prediction only, the controller will not move the manipulated variable for the specific controlled variable.

Table III.36 : 07DEC4 Gain Matrix

	MV1	MV2	MV3	DV1	DV2	DV3
	07PIC1098A.SP	07FIC1120.SP	07TIC1100.SP	07FIC1073.SP	00TI2004.PV	58TIC1017.SP

U07DEC4			Top debutanizer pressure	Reflux flow	Debutanizer tray 31 temp.	Feed to the column	Ambient Temperature	Hot Oil Temperature
CV1	07PIC1098A.OP	top pressure valve	-45			0.02	0.5	
CV2	07PDI1278.PV	pressure drop rectifying	-2	0.015				
CV3	07AI1119C.PV	Pentane in butane		-4	10			
CV4	DC4_REF_R.PV	Reflux to feed ratio DC4		0.002		-0.0072		
CV5	DC4_DUTY_R.PV	Hot oil to feed ratio DC4		0.004	0.5	-0.0100		
CV6	C4BOT_INF.PV	butane in top DC5 inf.			-0.005			0
CV7	07TI1108PCT.PV	bottom PCT			1			0

III.5.3.6. Unit 07 Deispentanizer (07DEC5)

The model matrix for the 07DEC5 controller is shown below.

Table III.37 : 07DEC5 Gain Matrix

07DEC5			MV1	MV2	DV1	DV2	DV3
			07FIC1192.SP	07TIC1174.SP	07FIC1168.SP	00TI2004.PV	58TIC1017.SP
			Reflux flow	tray 46 temp.	column feed	Ambient Temp.	Hot Oil Temperature
CV1	DC5_REF_R.PV	Reflux to feed ratio DC5	0.0078		-0.0176		
CV2	TVRBOT_INF.PV	TVR bottom product		-0.036		-0.00342	0
CV3	07TI1171PCT.PV	bottom PCT		2		0.19	0
CV4	DC5_DT_REB.PV	Reboiler delta temperature		-1			

III.5.3.7. Unit 16 Mixed refrigerant composition (U16MR)

The model matrix for the U16MR controller is shown below.

Table III.38 : U16MR Gain Matrix

	MV1	MV2	MV3	MV4
	16HIC1304.OP	16HIC1306.OP	16HIC1310.OP	16HIC1255.OP

U16MR			N2 to MR	C1 to MR (scrub ovhd)	C2 to MR (deetha ovhd)	C3 to MR
CV1	16AI2465A.PV	MR N2 content	0.0000380			
CV2	16AI2465B.PV	MR C1 content		0.0022000		
CV3	16AI2465C.PV	MR C2 content			0.0005900	
CV4	16AI2465D.PV	MR C3 content				0.0005200

III.5.4. Phases of an APC Project in a complex.

Impact on Gains Can Be Realized : The first phase of an APC project in a complex is to identify potential benefits and impact on gains that can be realized. Analyze current process and identify areas where APC can improve efficiency, reduce costs, and increase production. **[4]**

Preliminary Testing Steps : Conduct preliminary testing steps to validate feasibility of APC project. Test hardware and software components, and conduct simulation studies to ensure APC system can operate effectively in complex.

Functional Design Specification : Specify functional design of APC system. Define system's architecture, identify key components, and determine how they will interact with each other.

Step-Testing : Conduct step-testing of APC system. Test system in controlled environment to ensure it can operate as expected.

Identification of Process Models : Identify process models that will be used in APC system. Develop mathematical models of process that can be used to simulate and predict behavior of process.

Simulation : Conduct simulation studies using process models. Use models to simulate behavior of process under different operating conditions and evaluate performance of APC system.

Installation : Install APC system in complex. Integrate APC system with existing process control system and ensure it can operate effectively in complex.

III.5.5. Practical Steps in Building the Profit Controller

Two main steps are involved: **[5]**

Step 1: "Construction of the off-line mathematical model". This process involves a series of successive actions, including:

1. Identifying the list of model variables and selecting the type for each variable.

- CVs can be SP, PV, or OP.

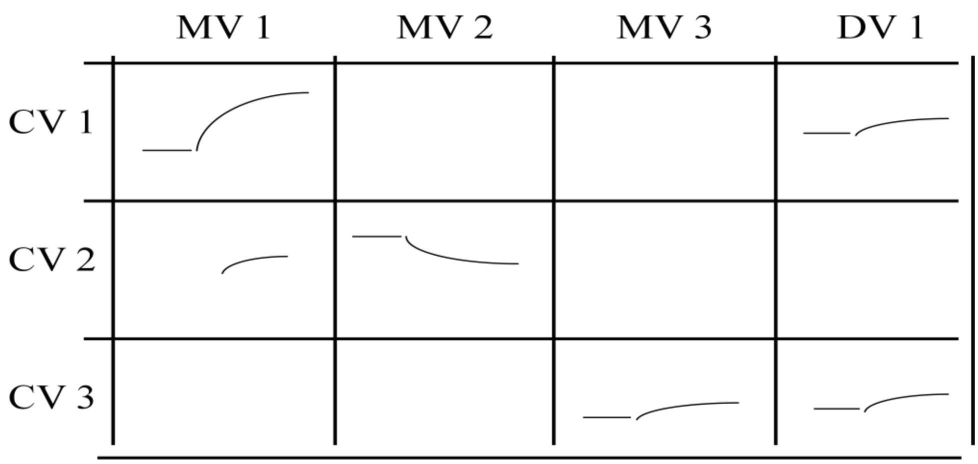
- In configuring the APC for MV, only SP or OP can be chosen because the MV is managed by the APC, which adjusts its value to keep the CV within its range.
- DV affects the CVs, but the Profit Controller cannot change it; normally, it should be a PV or SP.
- Keeping the process equipment within their operating ranges. For example, if FIC is an MV, then FIC can be included as a CV. If the lower limit of this CV is 30% (valve opening percentage), then the Profit Controller cannot change the FIC.SP to a value that does not comply with the lower limit of the CV.

2. Developing a plan for the **step tests**.

3. Conducting "off-line" Step Tests.

These **step tests** are performed online in the DCS to study the effect of the variation of each MV on the CVs, and after each test, we return to the initial operating state. Always record the transition time of the MVs and the change in the CVs.

- Before conducting the step tests, it is necessary to:
- "Break" all the cascades of the control loops (level-flow, temperature-flow, etc.) to use them in the step tests and use them in the RMPCT controller database as control variables and manipulated variables.
- Improve basic control through analysis to understand the behavior of the PID loops. Therefore, the APC working group conducts pre-tests to identify the state in which the PID loops will be configured for APC operation. For this, they have proposed recommendations to improve normal operation. Some PIDs can have their operation improved by adding a **FeedForward**. Control is then more precise thanks to the anticipatory effect of the variable put in **FeedForward** on the controlled output of the PID.



4. Using the "Profit Controller" tool for the construction and identification of the final mathematical model between each CV and the MVs, DVs that influence it in the form of a matrix.

$$CV1= f\,(MV1, MV3, DV1)$$

5. The coefficients of this model are determined from the recordings of the input values of the MVs and the response values of the CVs during the step tests. With these appropriate coefficients, the model can accurately predict the responses of the CVs.

Step 2: "Implementation of the APC (Advanced Process Control) model in the DCS"

This advanced control system must be installed on Windows, but the DCS, as we indicated in the theoretical part, is installed on Unix. Therefore, they set up a Unix-Windows station for the APC installation, with the second station remaining UNIX, and the third dedicated to alarms.For the installation of the APC, there is a server where all the components will be installed. The essential parts of the installation are the "**Profit Design Studio**" and the "**Profit Suite Operator Station**" which allow us to access the "**Profit Controller server**" to disable the "**Profit Controller human machine interface graphics**."

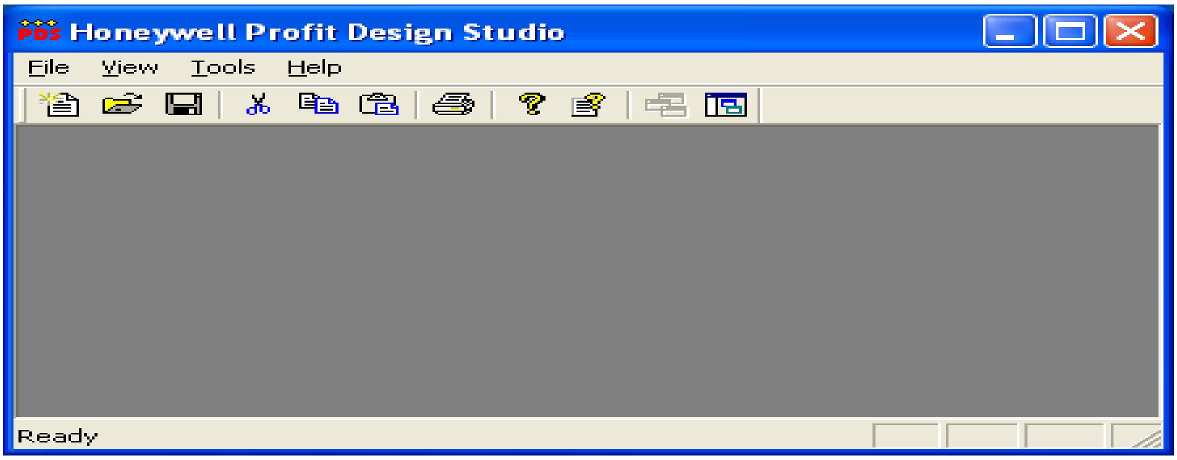


Figure III.3 : View of Profit design studio

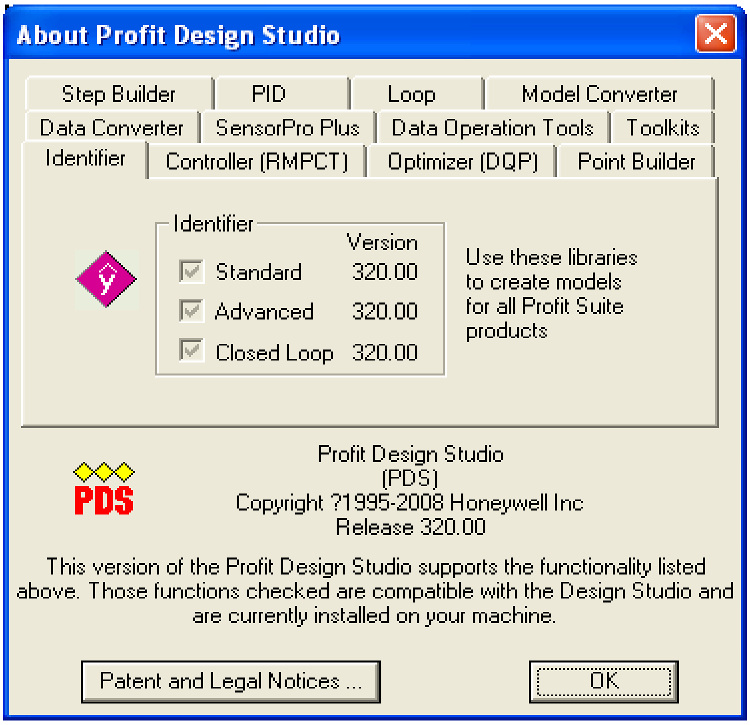


Figure III.4 : View of the Profit Design Studio window

By using the PSRS "**Profit Suite Runtime Studio**" we can define the **Profit Controller** and fill in the various sections of the following table

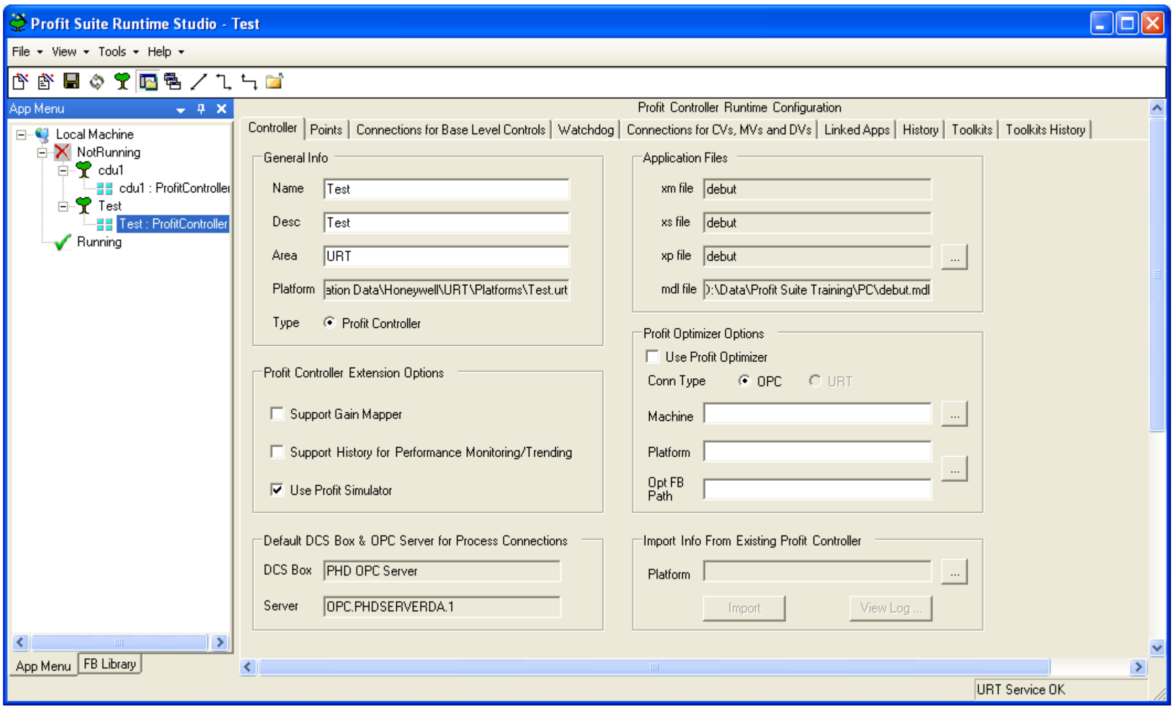


Figure III.5 :View of the "Profit Suite Runtime Studio" window

- URT Explorer is used for parameter control.
- PHD is used to access the Profit Controller Data history and parameter values.

After the complete installation of the APC, access to the new interfaces "PSOS" (Profit Suite Operator Station) is available, which engineers and operators use to replace the old interfaces.

- For modifying the final model if necessary (APC online), additional steps tests should be performed using the "Profit Stepper" tool.

III.6. Advanced Process Control (APC) Optimization of Depropanizer Column Performance

The Advanced Process Control (APC) uses manipulated variables (MVs) to maintain controlled variables (CVs) within their optimal operating ranges, considering the disturbances (DVs). For example, to control the top pressure valve (CV1), APC adjusts the top depropanizer pressure setpoint (MV1) to prevent saturation and stabilize the pressure. To ensure correct separation of butane in propane (CV2), APC modulates the reflux flow (MV2), thereby optimizing separation efficiency. The pressure-compensated temperature at tray 14 (CV3) is maintained by adjusting the depropanizer bottom temperature (MV3), ensuring precise component separation. The system also monitors the pressure drop in the rectifying section (CV4) and adjusts temperature and flow parameters to prevent flooding. The propane concentration in the top product of the debutanizer (CV5) is controlled by modulating the bottom temperature (MV3), ensuring product quality. Finally, the pressure-compensated temperature at the bottom of the column (CV6) is stabilized through adjustments to the bottom temperature (MV3), ensuring efficient separation of heavier components. By integrating these adjustments and accounting for disturbances such as feed flow (DV1), ambient temperature (DV2), and hot oil temperature (DV3), APC optimizes the overall performance of the depropanizer column.

III.6.1. Control Variables (CVs)

III.6.1.1. CV1: 07PIC1061A.OP (Top Pressure Valve)

- **Purpose :** To avoid saturation.
- **Explanation :** This variable monitors the position of the top pressure control valve. Saturation occurs when the valve is fully open or closed, which limits its ability to control the pressure effectively. This can cause instability and inefficiency in the process. By preventing valve saturation, the APC ensures stable pressure control and maintains optimal column performance.

III.6.1.2. CV2 : C4TOP_INF.PV (Butane in Propane Inferential)

- **Purpose :** To ensure correct separation of butane in propane.
- **Explanation :** This inferential measurement estimates the concentration of butane in the propane stream and is updated with real-time data from 07AI3062C.PV. Maintaining accurate control of this variable is crucial for ensuring product quality and meeting separation specifications.

III.6.1.3. CV3 : 07TI1069PCT.PV (Tray 14 Pressure Compensated Temperature)

- **Purpose :** To control temperature for accurate separation.
- **Explanation :** The pressure-compensated temperature at tray 14 is critical for the separation process. This calculated CV accounts for pressure variations, providing a more accurate temperature measure, and ensuring efficient component separation. The calculation is:
- $07TI1069PCT \ = 07TI1069.\boldsymbol{PV} - 42.1 \times ln(\frac{07PIC1061A.\boldsymbol{PV}}{12.8})$

III.6.1.4. CV4: 07PDI1319.PV (Pressure Drop Rectifying Section)

- **Purpose:** To avoid flooding.
- **Explanation:** This variable measures the pressure drop across the rectifying section of the column. A high pressure drop can indicate flooding, which disrupts the separation process. By keeping the pressure drop within acceptable limits, APC prevents flooding and ensures smooth column operation.

III.6.1.5. CV5 : C3BOT_INF.PV (Propane in Top Debutanizer Inferential)

- **Purpose:** To control propane concentration.
- **Explanation:** This inferential measurement estimates the propane concentration in the top product of the debutanizer and is updated with real-time data from 07AI1119A.PV. Controlling this variable ensures that the product meets specifications, maintaining the efficiency of the separation process.

III.6.1.6. CV6 : 07TI1060PCT.PV (Bottom Pressure Compensated Temperature)

- **Purpose:** To control the bottom temperature.
- **Explanation:** The pressure-compensated temperature at the bottom of the column is critical for separating heavier components. Accurate control ensures the bottom product meets quality standards and the column operates efficiently. The calculation is:

$$07TI1060PCT = 07TI1060.\boldsymbol{PV} - 35.0 \times ln(\frac{07PIC1061A.\boldsymbol{PV} + 1}{14})$$

III.6.2. Manipulated Variables (MVs)

III.6.2.1. MV1 : 07PIC1061A.SP (Top Depropanizer Pressure)

- **Purpose:** Maintain stable pressure control.
- **Explanation:** APC manipulates the setpoint for the top depropanizer pressure to maintain stable operation. By adjusting this setpoint, the system can avoid valve saturation, which helps in maintaining consistent pressure control and avoiding process upsets. This ensures that the pressure remains within the optimal range for efficient separation.

III.6.2.2. MV2 : 07FIC1086.SP (Reflux Flow)

- **Purpose:** Minimize energy consumption while ensuring product quality.
- **Explanation:** The reflux flow is a critical parameter for controlling the separation process. APC minimizes this flow based on the butane content in the top product. By reducing the reflux flow, the system lowers energy consumption while ensuring that the product specifications are met. This balance is crucial for operational efficiency and cost-effectiveness.

III.6.2.3. MV3: 07TIC1375.SP (Depropanizer Bottom Temperature)

- **Purpose:** Minimize temperature to reduce energy consumption.
- **Explanation:** APC aims to minimize the bottom temperature of the depropanizer to reduce energy consumption. Lowering the temperature reduces the amount of heat required, thus saving energy while maintaining effective separation. This optimization helps in reducing operational costs and improving overall process efficiency.

III.6.3. Disturbance Variables (DVs)

III.6.3.1. DV1 : 07FIC1035.SP (Feed to the Column)

- **Purpose:** To anticipate the effect of feed changes.
- **Explanation:** The feed flow to the column can vary due to upstream process changes, directly impacting the separation process. APC monitors this variable to anticipate changes in feed flow and adjust other parameters accordingly. By doing so, it ensures that the column operates smoothly and maintains optimal performance even when there are fluctuations in the feed rate. This proactive adjustment helps in maintaining product quality and process stability.

III.6.3.2. DV2 : 00TI2004.PV (Ambient Temperature)

- **Purpose:** To adapt the unit.
- **Explanation:** Ambient temperature changes can affect the overall process conditions, including heat transfer rates and equipment efficiency. APC monitors ambient temperature to account for these variations and make necessary adjustments in real-time. This helps maintain consistent operation and product quality, as the system adapts to external temperature fluctuations, ensuring that the process remains stable and efficient.

III.6.3.3. DV3 : 58TIC1017.SP (Hot Oil Temperature)

- **Purpose:** Compensate the dynamic effect.
- **Explanation:** The temperature of the hot oil used in heat exchangers can fluctuate, impacting the efficiency of the heat transfer process. APC monitors this variable to adjust other parameters and compensate for changes in hot oil temperature. This ensures that the depropanizer column operates at optimal performance levels by maintaining the necessary heat balance. By dynamically adjusting to these temperature changes, APC helps in preventing process disturbances and maintaining energy efficiency.

III.6.4. Explanation of APC Operation

III.6.4.1. Monitoring and Adjustment

APC continuously monitors CVs such as top pressure valve position, inferential measurements, and pressure-compensated temperatures to detect deviations from desired conditions. When deviations are detected, APC adjusts MVs like top depropanizer pressure, reflux flow, and bottom temperature setpoints to bring the process back within desired limits and ensures stability and optimal performance.

III.6.4.2. Predictive Control

APC uses predictive models to anticipate the effects of DVs like feed flow, ambient temperature, and hot oil temperature. For instance, if feed flow changes, the predictive model estimates the impact and adjusts MVs to maintain stability and performance.

III.6.4.3. Optimization

APC optimizes depropanizer column performance by minimizing energy consumption and maximizing efficiency. By reducing reflux flow and bottom temperature while meeting product specifications, APC lowers the energy required for the separation process. This optimization not only saves costs but also reduces the environmental impact by lowering energy consumption.

III.6.4.4. Interaction Handling

Managing multiple interacting variables in the depropanizer column is complex. APC uses techniques like the Relative Gain Array (RGA) to understand and mitigate these interactions. RGA helps identify which MVs impact CVs most significantly, allowing APC to adjust control strategies effectively, ensuring smoother operation and improved efficiency.

III.6.5. Cascade Control in the Depropanizer Column

In a cascade control system, two or more controllers are used to manage a single process variable. These controllers work in a master-slave configuration, where the master controller sets the setpoint for the slave controller. This approach allows for more precise and stable control of complex processes. Let's look at how cascade control operates in the context of the depropanizer column.

III.6.5.1. Control Loops and Roles

Master Controller: This is the primary controller responsible for maintaining the main process variable at its setpoint.

Slave Controller: This secondary controller receives its setpoint from the master controller and controls a related secondary variable.

III.6.5.2. Reboiler Cascade Control:

Master Controller (07TIC1375): Monitors the bottom temperature of the column and adjusts its setpoint accordingly.

Slave Controller (07FIC1068): Receives the setpoint from the master controller and controls the flow of hot oil to the reboiler to achieve the desired temperature.

Operation: The master controller regulates the bottom temperature, while the slave controller adjusts the hot oil flow rate to maintain this temperature. This ensures precise temperature control at the bottom of the column, critical for efficient separation.

Explanation : The Reboiler Cascade Control operates using a cascade control system to maintain the liquid temperature in the reboiler of the depropanizer column at optimal levels. Four temperature indicators play key roles in this process: 07TI1066 monitors the temperature of the liquid entering the reboiler, 07TI1075 controls the temperature of the hot oil heating the liquid, 07TI1067 measures the temperature of the propane as vapor exiting the reboiler, and 07TI1149 monitors the temperature of the heated liquid returning to the column.

The master controller, 07TIC1375, adjusts the flow rate of hot oil based on the measured liquid temperature, while the slave controller, 07FIC1068, receives the setpoint from the master controller and regulates the hot oil flow rate to achieve the target temperature in the reboiler. This cascade system ensures precise regulation of the liquid temperature in the reboiler, thereby ensuring optimal operation of the depripanizer column.

III.6.5.3. Cascade Control for product flow valve

In the depropanizer column, the product flow valve control system uses a sophisticated cascade control strategy to maintain optimal operation. This strategy involves two master controllers, 07PIC1061 and 07LIC1083, and one slave controller, 07FY1087, which operates as a comparative relay. Here's a detailed explanation of how this system works:

Master Controllers:

07PIC1061 (Pressure Controller): This controller monitors and regulates the product flow of the depropanizer column. It ensures the pressure stays within the desired setpoint (SP) by adjusting its output (OP) signal.

07LIC1083 (Level Controller): This controller manages the liquid level in the reflux drum. It ensures the level is maintained within the desired setpoint (SP) by adjusting its output (OP) signal.

Slave Controller (Comparative Relay):

07FY1087: This slave controller acts as a comparative relay that receives the output (OP) signals from both master controllers 07PIC1061 and 07LIC1083. It continuously compares these signals and selects the minimum value between them. This minimum value is then used to control the final control element, typically a valve, ensuring the system operates within safe and efficient limits.

Operation:

Pressure Control Loop : The primary objective of the 07PIC1061 pressure controller is to maintain the product flow of the depropanizer column. It adjusts its output signal based on the pressure setpoint and the actual pressure (PV - process variable).

Level Control Loop : The 07LIC1083 level controller ensures the liquid level in the reflux drum is within the desired range. It adjusts its output signal based on the level setpoint and the actual level (PV - process variable).

Comparative Relay Action: The 07FY1087 controller receives the output signals from both 07PIC1061 and 07LIC1083. It acts as a relay by selecting the lower of the two signals to ensure the process remains stable and prevents overcorrection by either controller.

This ensures that if the pressure or the level gets too high, the appropriate control action is taken without one control loop overriding the other excessively.

III.6.5.4. Bottom Product Valve cascade control:

Master Controller (07LIC1070):

Function : This Level Indicator Controller (LIC) regulates the level of the bottom product in the depropanizer column.

Operation : It adjusts the setpoint for the flow rate of the bottom product based on the liquid level in the column.

Slave Controller (07FIC1073):

Function : This Flow Indicator Controller (FIC) receives the setpoint from the master controller (07LIC1070) and regulates the flow of the bottom product.

Operation: It adjusts the flow rate of the bottom product to maintain the desired level as dictated by the master controller.

Explanation : The Bottom Product Valve Cascade Control operates using a cascade control system to maintain the liquid level in the bottom of the depropanizer column at optimal levels. The master controller, 07LIC1070, continuously monitors the liquid level in the column. When the level deviates from the setpoint, the master controller adjusts the setpoint for the flow rate of the bottom product to bring the level back to the desired value.

The slave controller, 07FIC1073, then takes this flow rate setpoint and ensures that the actual flow rate of the bottom product matches the setpoint. It does this by adjusting the control valve that regulates the flow of the bottom product. This precise regulation of the bottom product flow rate ensures that the liquid level in the column remains stable and within the optimal range, which is essential for maintaining efficient separation processes in the depropanizer column.

III.7. Operator interface

The **Profit Suite Operation Station** (PSOS) provides a user interface for Profit Controller applications. This section presents the layout of the Operators' interface.

III.7.1. Control Strategy

The operator manages the Profit Controller control applications through the **PSOS**, provided as a shortcut on the operator workstation's desktop. The **PSOS** access can be specialized depending on the user (operator or engineering access), So the operator can use only a few needed displays that will permit him to perform all the operations. The operator can see all displays but not all the fields are accessible to him because some are under engineering access. An example is the tuning fields of a controlled variable in the optimize displays under the CV page. Anytime the field is not accessible, the system informs the operator that it is impossible to modify that field. The most important **PSOS** operator displays are reported and briefly described in the following paragraphs.

III.7.2. PSOS Controller interface

A mouse click on the "Launch default view" button brings to the APC controller pages. The first page that is visualized is the “My View” page. The controller interface gives the possibility to the operator to switch on and off controller, access to the CV, MV, DV and sub- controllers pages. The main interface is reported in Figure 5. On this page the operator can access to different commands such as controller switch, and can read information on controller execution.

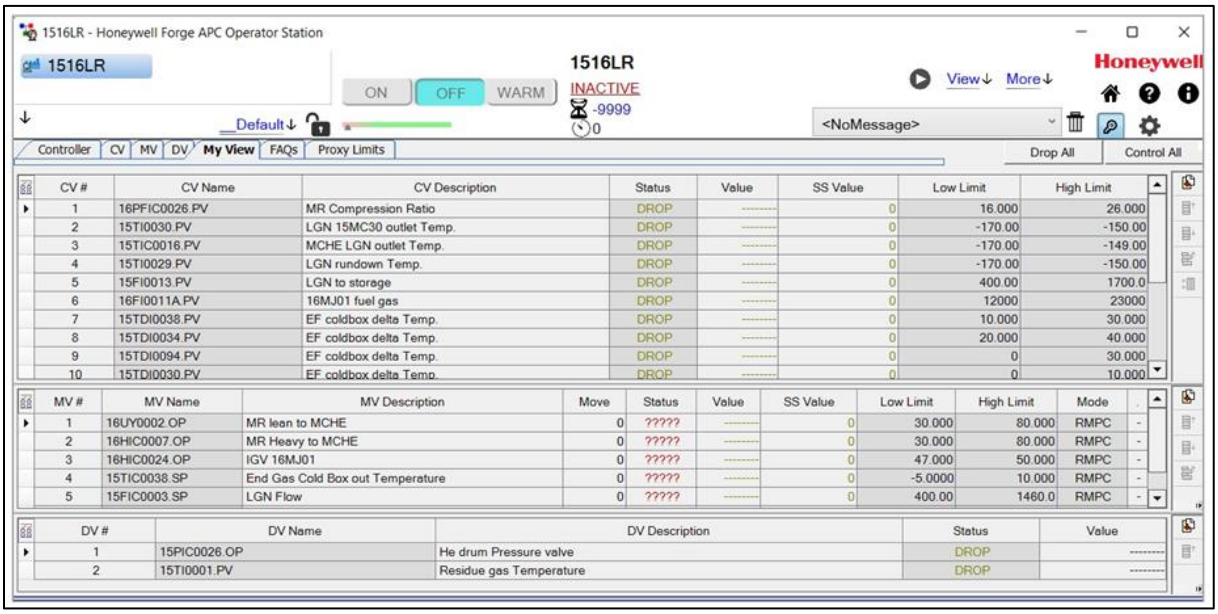


Figure III.6 : APCOS Main Page

The meaning of the controller switches in the top middle section of the page (see below) is the following:

- ON: controller is set to normal operation mode, calculating and writing moves to MV and controlling CV between limits.
- WARM: controller calculates the movements needed to respect CV and MV limits but <u>it does NOT write anything to the DCS</u>. This should be used only in order to initialize predictions after controller stop and to check the movement plan.
- OFF: controller is set to OFF. No calculation or writing is done, application remains in stand-by in the background

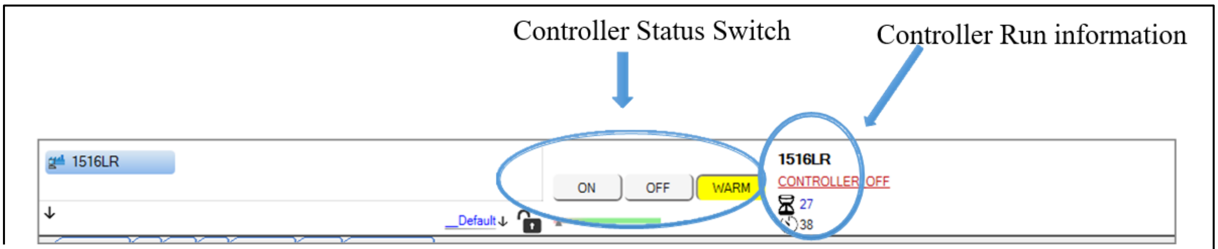


Figure III.7 : Controller status swich

In the same page, some information are accessible in the control running information, such as time to Next run in seconds, actual controller status, and number of intervals since start. Controller status can assume the following values:

- INACTIVE: Controller is inactive, and nothing is done.
- CONTROL OK: Controller running (no optimization).
- INITIALIZING: Controller is starting up aligning to current process conditions.
- SHEDDING CONNECTIONS: Controller has been shut off and it is now shedding connections to PID loops.
- WAIT FOR EXECUTION: Controller is waiting for background calculations to complete.
- HANDLING CONTRAINTS: Not all the constraints can be met at the actual execution. Check on the controller actual constraints is needed.
- OPTIMIZING: Controller is running and optimizing up to economic objectives.

In the following paragraphs the single pages of the controller interfaces are introduced.

III.7.3. Controller Pages

The description of the pages available in PSOS that are of interest to operators is reported below.

III.7.3.1. Detail page

This page reports the main information for RMPCT controller. The operator has only reading access to this page; the page is useful for operators for the last entry "Application Status" which should show a "ACTIVE" green label:

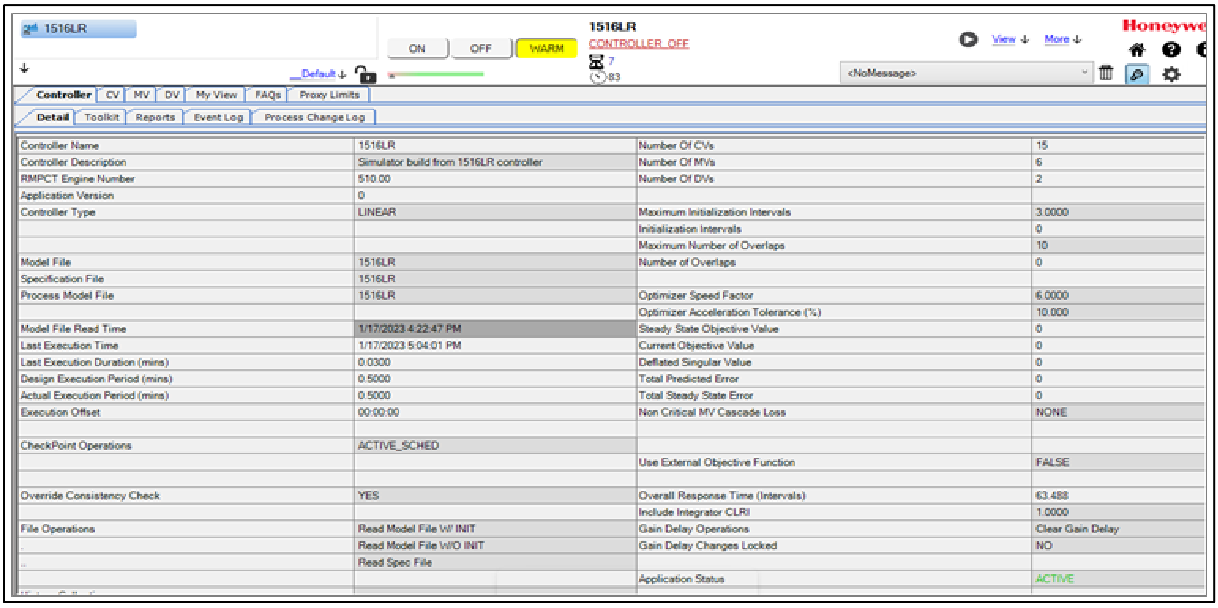


Figure III.8 : Controller Detail Page

III.7.3.2. Toolkit Page, Reports Page, and Event Log Page

The operator does not use them; they are visible but essentially used by Engineers with access to the APC application configuration files.

III.7.4. CVs Pages

The description of the CV pages available in PSOS that are of interest for operators is reported below. [6]

III.7.4.1. Summary page

The CVs page is the page that is opened by default when the controlled visualization is launched. Here, the operator can access to the high and low limits of CVs and set them to a proper value. This page shows the following column:

- CV #: the number of CV in the controller's internal list. It cannot be modified, and the controller uses it to refer to the CV, for example, to understand the reason for which an MV is moving, as it will be shown later.
- CV Name: The name of the CV.
- CV Description: The description of the CV.
- Status: CV actual status. Values that actual status can get are reported below.
- Value: CV actual value.
- Future Value: CV predicted value after controller movement plan.
- SS Value: Value that CV will reach at steady state.
- Low Limit: CV operator lower limit.
- High Limit: CV operator high limit.
- Set-point: CV set-point value (when applicable).

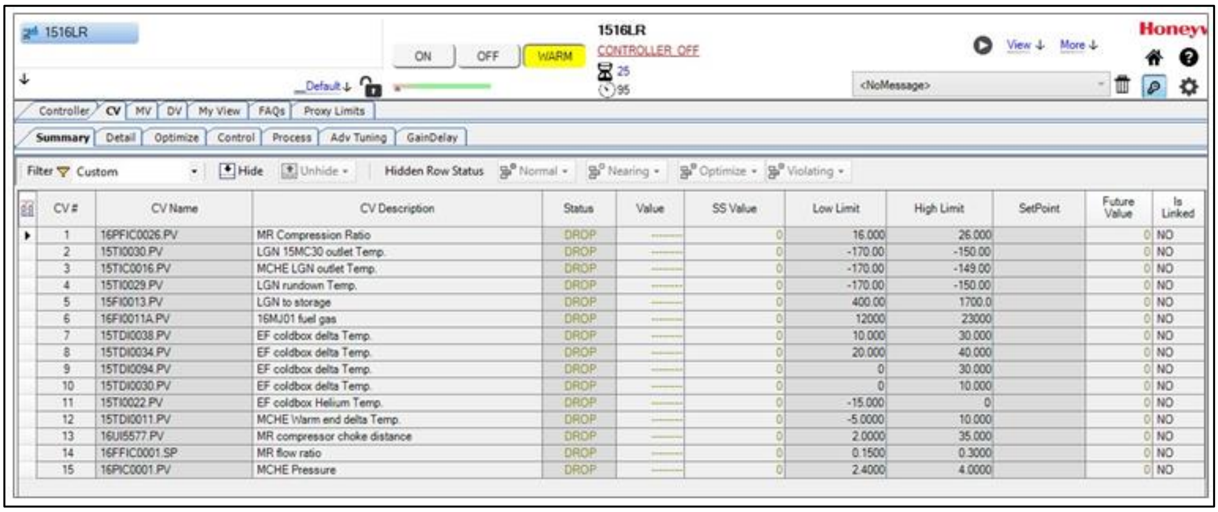
The possible values for the status of the controller are:

- GOOD: CV is OK and used for control and optimization.
- DROP: CV is out of the controller and not used inside calculations.
- INIT: Variable is initializing.
- WOUND UP: The controller cannot change that CV (all MV are constrained).
- PRED: Prediction is used instead of true value (the measured variable is BAD).
- INAC: The controller is inactive.

The meaning of the colors of the font in the Value, Future Value, and SS Value columns is the following:

- Black: Measured/Predicted value between limits (more than 0.1% inside limits).
- Yellow: Measured/Predicted value near limits (between 0.1% inside and 1% outside limits).
- Red: Measured/Predicted value violates limits (more than 1% outside limits).
- In the Low and High limit Columns, some symbols can appear depending on the actual operations of the controller:

- Green arrow (up or down): the CV is going up at the maximum speed.
- Solid or hollow yellow diamond: hard or soft constraints are reached



CV#	CV Name	CV Description	Status	Value	SS Value	Low Limit	High Limit	SetPoint	Future Value	Is Linked
1	16PFIC0026.PV	MR Compression Ratio	DROP	--------	0	16.000	26.000		0	NO
2	15TI0030.PV	LGN 15MC30 outlet Temp.	DROP	--------	0	-170.00	-150.00		0	NO
3	15TIC0016.PV	MCHE LGN outlet Temp.	DROP	--------	0	-170.00	-149.00		0	NO
4	15TI0029.PV	LGN rundown Temp.	DROP	--------	0	-170.00	-150.00		0	NO
5	15FI0013.PV	LGN to storage	DROP	--------	0	400.00	1700.0		0	NO
6	16FI0011A.PV	16MJ01 fuel gas	DROP	--------	0	12000	23000		0	NO
7	15TDI0038.PV	EF coldbox delta Temp.	DROP	--------	0	10.000	30.000		0	NO
8	15TDI0034.PV	EF coldbox delta Temp.	DROP	--------	0	20.000	40.000		0	NO
9	15TDI0094.PV	EF coldbox delta Temp.	DROP	--------	0	0	30.000		0	NO
10	15TDI0030.PV	EF coldbox delta Temp.	DROP	--------	0	0	10.000		0	NO
11	15TI0022.PV	EF coldbox Helium Temp.	DROP	--------	0	-15.000	0		0	NO
12	15TDI0011.PV	MCHE Warm end delta Temp.	DROP	--------	0	-5.0000	10.000		0	NO
13	16UI5577.PV	MR compressor choke distance	DROP	--------	0	2.0000	35.000		0	NO
14	16FFIC0001.SP	MR flow ratio	DROP	--------	0	0.1500	0.3000		0	NO
15	16PIC0001.PV	MCHE Pressure	DROP	--------	0	2.4000	4.0000		0	NO

Figure III.9 : Controller CV Summary page

III.7.4.2. CV detailed page

Clicking on one of the CVs, the detail page will appear. In this page, the main information on the CV is reported, such as description, values and configured limits.Operators can view this page but has no access to the engineering parameters. The useful information for operators in the CV Detail page is essentially:

Engineering Low Limit: this is the low limit set by engineer; operators have no right to set the low limit of the selected CV below the Engineering Low Limit. An Engineering Low Limit set equal to " " means that no limit is specified and the operator is free to choose the low limit for the selected CV.

Engineering High Limit: this is the high limit set by engineer; operators have no right to set the high limit of the selected CV above the Engineering High Limit. An Engineering High Limit set equal to " " means that no limit is specified and the operator is free to choose the high limit for the selected CV.

An example of Detail Page for one CV is reported below:

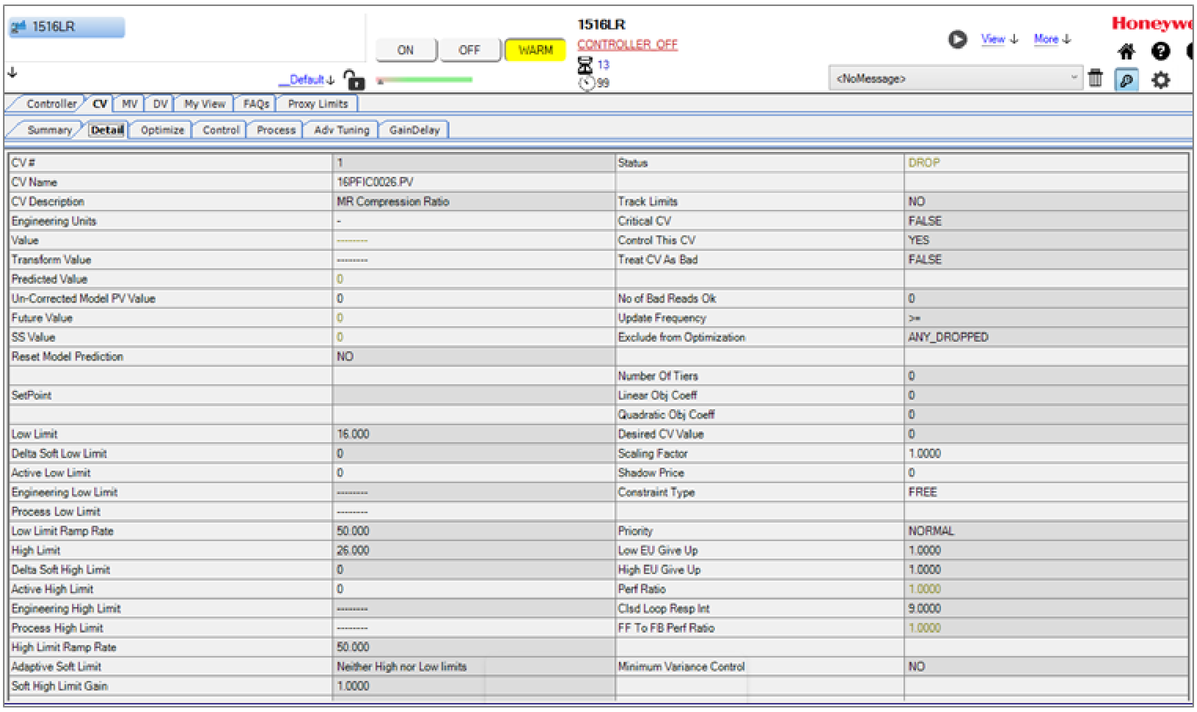


Figure III.10 : CV Details page

III.7.4.3. Optimize, Control Process and advanced tuning pages

These pages report the tuning parameters of the CVs. Operator has not writing access to them, and he can only read the value set by engineers.

III.7.4.4. Gain Delay

These pages contain the model gains for the Controllers. The table here reports, for any CV-MV couple, the value of model gain. If no gain is present (or if it is 0), no model is present. This is an only-view page for the operators. Below an example of the Gain Delay page is reported.

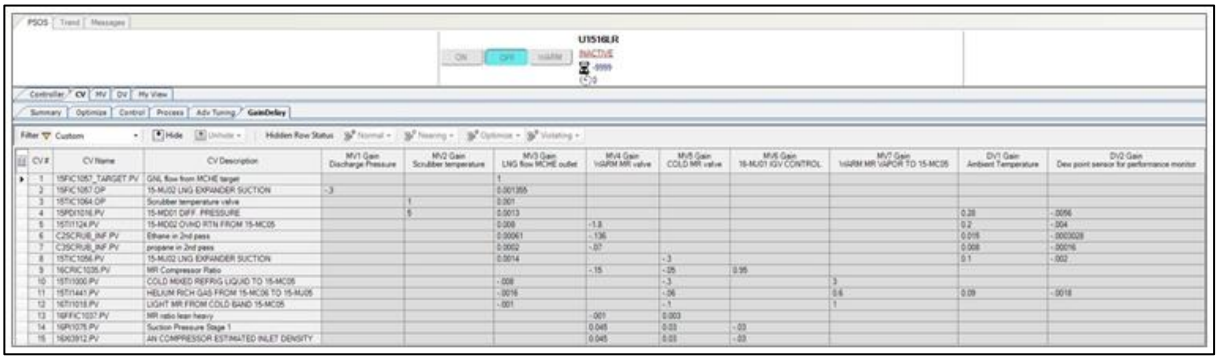


Figure III.11 : CV Gain-Delay page

III.7.5. MV page

The description of the MV pages available in PSOS that are of interest to operators is reported below. **[6]**

III.7.5.1. Summary page

Clicking on the MV tab on the upper part of the page, the operator can access to MV display pages. Here, the operator can change the high and low limit of MVs, and set them to a desired value. This page shows the following column:

- MV #: the number of MV in the controller internal list. It cannot be modified.
- MV Name: The name of the MV.
- MV Description: The description of the MV.
- Status: MV actual status. Values that actual status can get are reported below.
- Value: MV actual value.
- Future Value: MV value at the end of controller movement plan.
- SS Value: Value that MV will reach at steady state.
- Low Limit: MV operator lower limit
- High Limit: MV operator high limit
- Move Cause: the aim of moving the CV (Optimization, Control, Interaction-Opt, Not Determined).
- Main CV#: the number of the CV that principally causes the movement of the MV (onlyif Move cause is not "Optimization" nor "Not Determined").
- Mode: Can be set by the operator. If set to RMPC the variable is available for Profit Controller. If it is set to OPER, the MV can adopt the value FFWD (treated as a feed forward disturbance, used for prediction but not for control) or DROP (MV is droppedfrom both prediction and control). FFWD or DROP depends on engineering tuning of the controller, and operator cannot change it.

The possible values for status are:

- ON: MV is OK and the controller is ON.
- READY: MV is OK and the controller is OFF or WARM.
- INIT: Variable is initializing.
- FFWD: MV is used as a feed-forward variable.
- LOW: MV is clamped low or it is at its low limit.
- HIGH: MV is clamped low or it is at its low limit.
- SERV: Communication with the process is lost, non critical MV is ignored
- INAC: The controller is inactive.

The meaning of the colours of the font in the Value and SS Value columns is the following:

- Black: Measured/Predicted value between limits
- Yellow: Measured/Predicted value near limits.

- Red: Measured/Predicted value violates limits.

In the Low and High limit Columns some symbols can appear depending on actual operationsof the controller:

- Green arrow (up or down): the MV is going up at the maximum speed.
- Solid or hollow yellow diamond: hard or soft constraints are reached

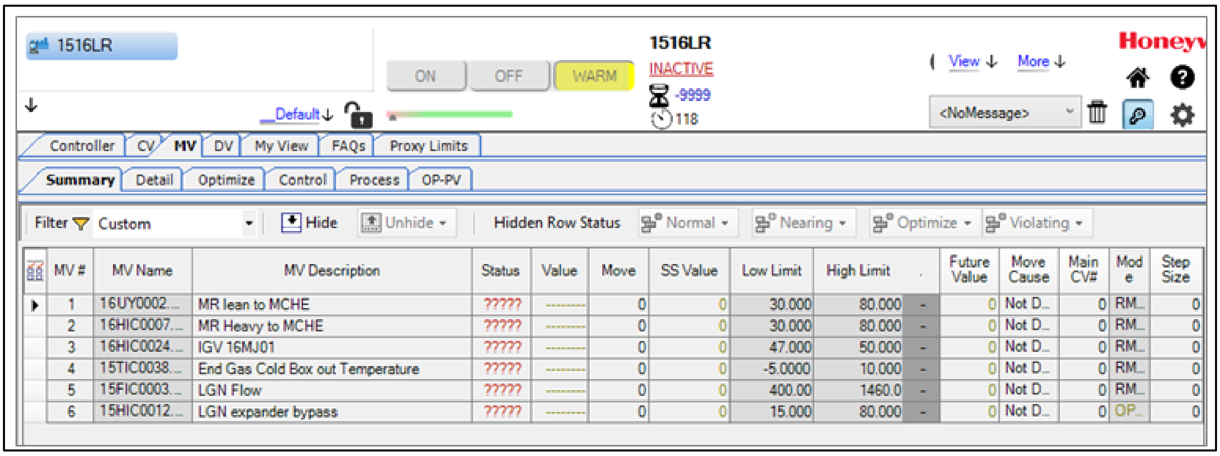


Figure III.12 : Controller MVs Summary Page

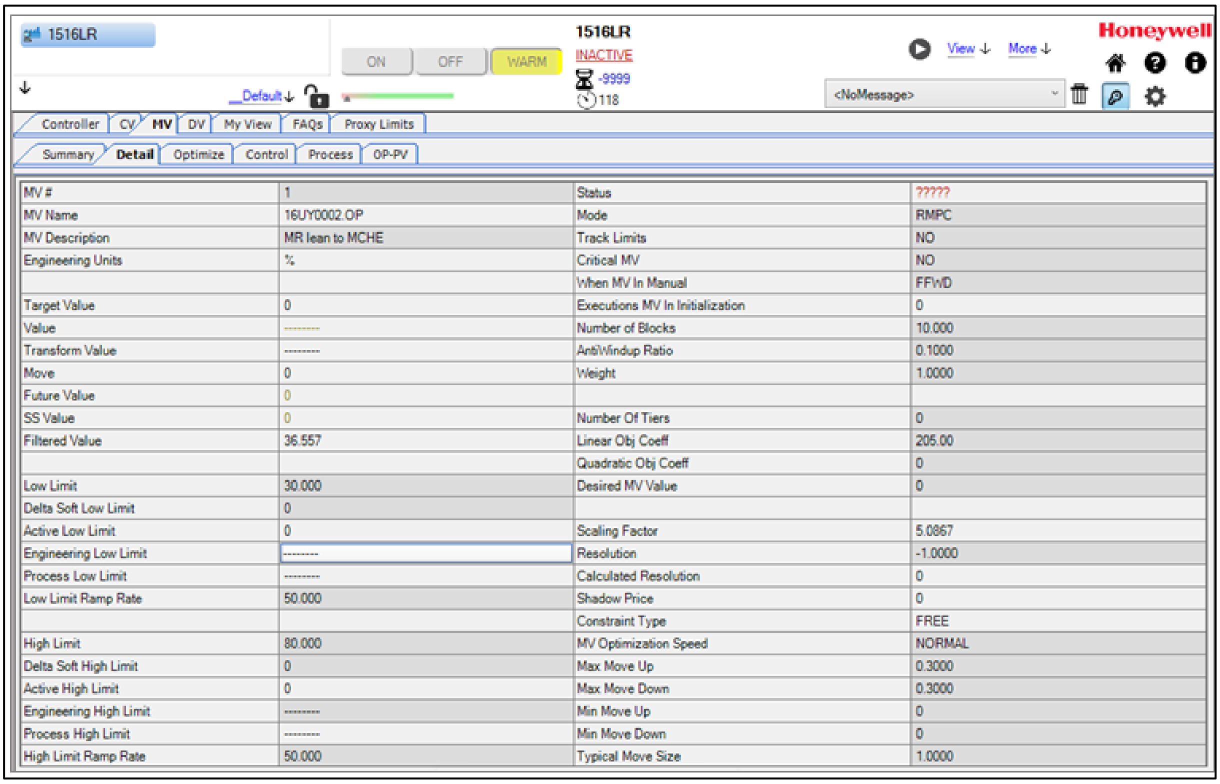
III.7.5.2. MV Detail Page

Clicking on one of the MVs, the detail page will appear. In this page, the main information on the MV is reported, such as description, values and configured limits.

Operators can view this page but has no access to the engineering parameters. The useful information for operators in the MV Detail page is essentially:

Engineering Low Limit: this is the low limit set by engineer; operators have no right to set the low limit of the selected MV below the Engineering Low Limit. An Engineering Low Limit set equal to " " means that no limit is specified and the operator is free to choose the low limit for the selected MV.

Engineering High Limit: this is the high limit set by engineer; operators have no right to set the high limit of the selected MV above the Engineering High Limit. An Engineering High Limit set equal to " " means that no limit is specified and the operator is free to choose the high limit for the selected MV. **[6]**



MV #	1	Status	?????
MV Name	16UY0002.OP	Mode	RMPC
MV Description	MR lean to MCHE	Track Limits	NO
Engineering Units	%	Critical MV	NO
		When MV In Manual	FFWD
Target Value	0	Executions MV In Initialization	0
Value	--------	Number of Blocks	10.000
Transform Value	--------	AntiWindup Ratio	0.1000
Move	0	Weight	1.0000
Future Value	0		
SS Value	0	Number Of Tiers	0
Filtered Value	36.557	Linear Obj Coeff	205.00
		Quadratic Obj Coeff	0
Low Limit	30.000	Desired MV Value	0
Delta Soft Low Limit	0		
Active Low Limit	0	Scaling Factor	5.0867
Engineering Low Limit	--------	Resolution	-1.0000
Process Low Limit	--------	Calculated Resolution	0
Low Limit Ramp Rate	50.000	Shadow Price	0
		Constraint Type	FREE
High Limit	80.000	MV Optimization Speed	NORMAL
Delta Soft High Limit	0	Max Move Up	0.3000
Active High Limit	0	Max Move Down	0.3000
Engineering High Limit	--------	Min Move Up	0
Process High Limit	--------	Min Move Down	0
High Limit Ramp Rate	50.000	Typical Move Size	1.0000

Figure III.13 : MV Detail Page

III.8. Conclusion

In this chapter, we explained how APC technology works and provided key insights for its effective use. APC is a powerful tool for improving performance, product quality, and profitability in industrial processes. Understanding how it operates and its best practices helps engineers and operators make the most of APC and tackle current industrial challenges.

We also discussed how APC controls the depropanizer column using Controlled Variables (CVs), Manipulated Variables (MVs), and Disturbance Variables (DVs) within the RMPCT framework. Additionally, we examined how APC manages the cascades of the column. This comprehensive approach ensures a solid understanding of APC applications in depropanizer column control, enhancing overall system efficiency and stability.

CHAPTER.IV. Comparative Interactions study between PID and APC

IV.1. Introduction

In industrial engineering, the automatician's perspective on processes revolves around dynamic systems with inputs and outputs that must be controlled to meet quality standards and product specifications. As most industrial systems are multivariable, controlling them has become a significant focus in scientific research. Various control techniques have been developed to enable the complete or partial decomposition of multivariable systems while meeting the necessary requirements.

IV.2. Multivariable systems

IV.2.1. Definition

A system is a collection of parts, objects, or entities that perform a specific operation. It is defined by its inputs and outputs, which connect it to the external environment.

A multivariable system is defined as a process that has more than one input (setpoint, control) $U = (u_1, u_2, \dots\dots, u_p)$and/or more than one output (effects, measurements) $Y = (y_1, y_2, \dots\dots, y_q)$. An output can be affected by more than one input (interaction), and the system may be subject to disturbances or noise.

Vector notations:

$$U^T = [u_1, u_2, \dots\dots, u_p] \text{ , } Y^T = [y_1, y_2, \dots\dots, y_q]$$

With :

U: Command vector (inputs, setpoints) of dimension (p×1).

Y: Output vector (measurement) of dimension (q×1).

p: Number of system commands.

q: Number of system outputs.

Mathematically, a multivariable system can be represented by a set of differential equations or transfer functions with multiple inputs and outputs. The interactions between variables are described by the off-diagonal elements in the transfer function matrix.

IV.2.2. Multivariable Regulation

The design of a control system suitable for a multivariable industrial process obviously poses a number of problems. Among these, the problem of the existence of interactions between the input-output variables of the system is the main cause for which the synthesis and operation of the system in multi-loop, loop by loop, are difficult because a change in an input variable results in changes in several output variables; which makes it difficult to maintain the performance of each loop. Moreover, the performance of a control loop can be strongly affected by the parameters of the regulators of the other loops. **[9]**

In the general context of multivariable system control, considerable attention has been paid to the concept of interaction analysis. In this perspective, the aim is often to compensate the system so that:

- Each input affects only one output;
- Disturbance on one output, with inputs being zero, affects only that single output.

IV.2.3. Multiloop control

The multiloop control technique provides an acceptable level of performance in the majority of cases. The synthesis of a multiloop control system is carried out in two steps:

Step 1 : Determination of the control configuration by selecting input-output pairs (each input must be looped with a single, well-defined output, introducing a well-designed regulator).

Step 2 : Choice of the control law and determination of the regulator parameters for each loop ensuring the desired performance.

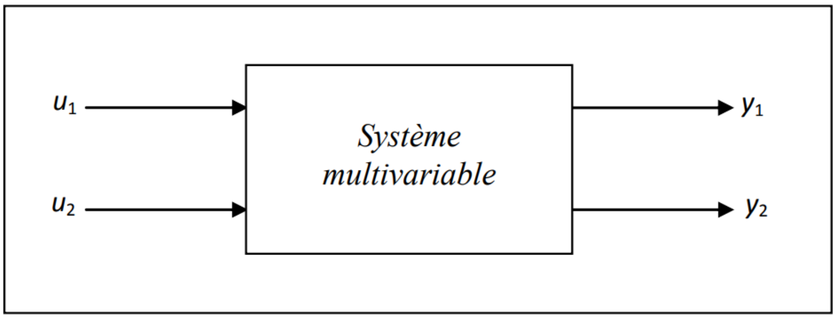


Figure IV.1 : Functional diagram of multivariable system

In the first step, the choice of the appropriate control configuration, i.e., the configuration in which the interactions between the resulting control loops are very weak, is guided by the use of an interaction analysis method which also allows evaluating the level For a two-input, two-output multivariable system, as illustrated in the figure:

Two control configurations are possible:

- U_1 controls Y_1 and U_2 controls Y_2, configuration designated by
- U_1 controls Y_2 and U_2 controls Y_1, configuration designated by

For the multiloop control of a system, the most important step is the choice of the control configuration (input-output pairs). This is determined by analyzing the interactions present in the system. The choice is based on the configuration with very low levels of interaction between the different control loops, while ensuring the stability of each loop and that of the overall system (closed-loop system). **[9]**

IV.2.4. Definition of interaction in a multivariable system

The control loops in a multivariable system are said to be interactive if a control action $U_k(S)$ in the k-th loop (resulting from a disturbance $Z_k(S)$ or a setpoint change $C_k(S)$) causes a control action $u_l(S)$ $(l \neq k)$ in one or more loops, with the aim of maintaining the output variables $y_l(S)$ $(l \neq k)$ assigned to them at their setpoints.

IV.2.5. Explanation of the interaction phenomenon

To clarify the interaction phenomenon in a multivariable system, represented in the figure IV.2.

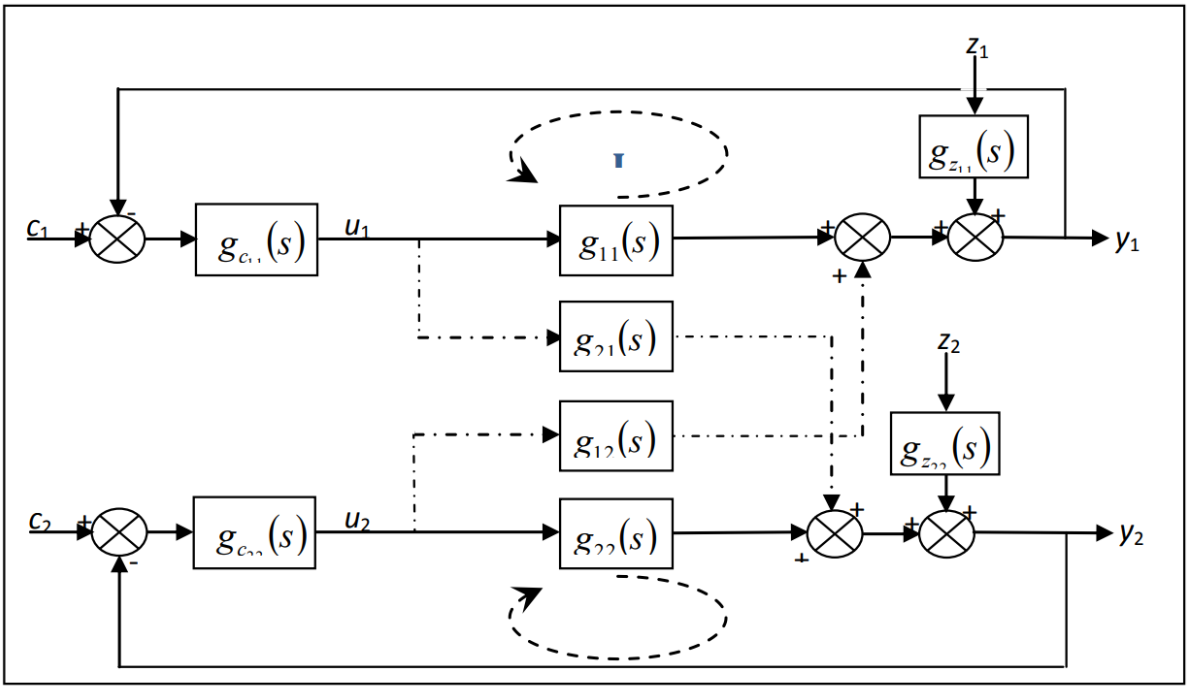


Figure IV.2 : Distributed control of a system

When the disturbance Z_1 affects the output Y_1, causing it to deviate from its setpoint C_1, the controller $g_{c11}(S)$ generates a control signal U_1 to eliminate this deviation (solid line). However, the generated control signal U_1 also affects the output Y_2 through the transfer function $g_{c21}(S)$ (dashed line), causing Y_2 to deviate from its setpoint C_2. This necessitates the controller $g_{c22}(S)$ to generate a control signal U_2 to maintain Y_2 at the desired setpoint C_2. The corrective action of the controller $g_{c22}(S)$ of the second loop (II) (the control signal U_2) also affects the output Y_1 through the transfer function $g_{12}(S)$. Thus, maintaining both Y_1 and Y_2 at their desired positions despite the disturbance Z_1, which should be canceled by the controller $g_{c11}(S)$, becomes a challenging task.

In this example, we demonstrated that changing one control variable U_1 has two effects:

1. **A direct effect** : on Y_1, the measured variable of the loop.
2. **An indirect effect** : on Y_2, through loop interactions.

If there is a disturbance affecting one output, it propagates through the system and affects other outputs. This is due to the interactions between control loops.

In this example, the transfer functions g_{21} and g_{12} are responsible for the interactions between the two loops. These interactions have a detrimental effect on the robustness of control strategies. The phenomenon of interaction poses a constraint for controlling multivariable systems, which is why interaction analysis is essential for synthesizing a control system for a multivariable system.

IV.2.6. Interaction analysis method

Interaction analysis is a crucial step in designing effective control strategies for multivariable systems. It involves studying the interactions between the inputs and outputs of a system to determine the best loop pairings and mitigate the effects of interactions.

The choice of interaction analysis method depends on the specific characteristics of the multivariable system and the control objectives Once the interactions are characterized, appropriate control strategies can be designed, such as decentralized control with decoupling compensators or multivariable control techniques that explicitly consider the interactions.**[9]**

IV.2.7. Relative Gains Matrix

The Relative Gain Array (RGA) method was introduced by Bristol in 1966. This method is a powerful tool for selecting control pairs and is easy to implement, which is why it is frequently used in process engineering. It helps us identify a control configuration with a low level of interaction. **[10]**

We consider a process with two controls and two outputs to introduce the concept of relative gain

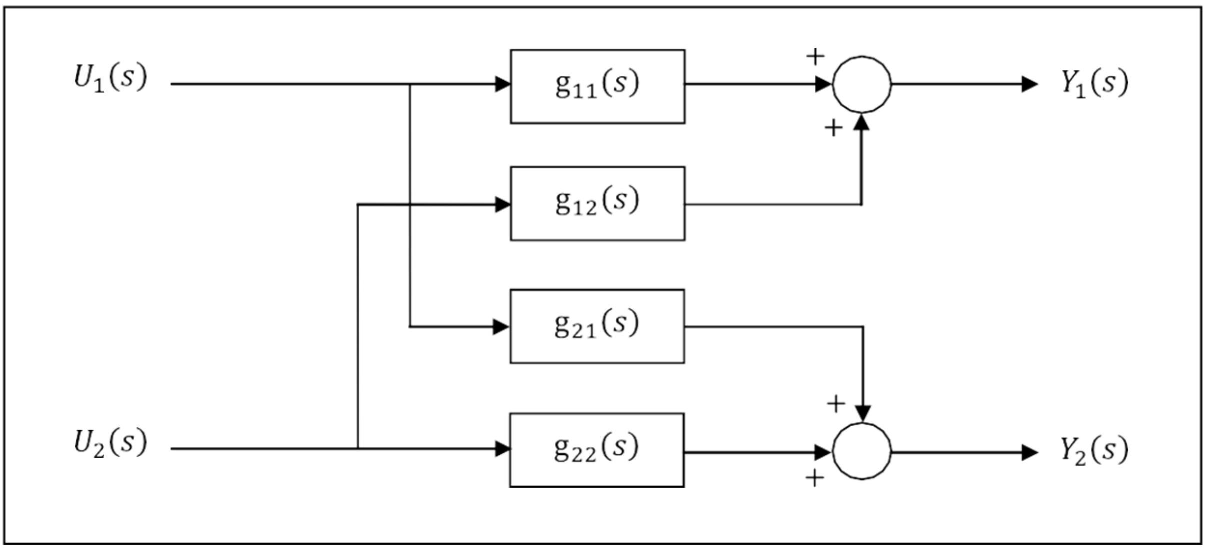


Figure IV.3 : Block diagram of a multivariable system (2×2)

$$Y_1(S) = g_{11}(S)U_1(S) + g_{12}(S)U_2(S)$$

$$Y_2(S) = g_{21}(S)U_1(S) + g_{22}(S)U_2(S)$$

The transfer function between control U_1 and output Y_1 in open loop is:

$$\left(\frac{Y_1(s)}{U_1(s)}\right)_{bo} = g_{11}(s)$$

Suppose that we regulate only the output Y_2 using a controller whose transfer function is $g_{c22}(S)$, as shown in the figure IV.4.

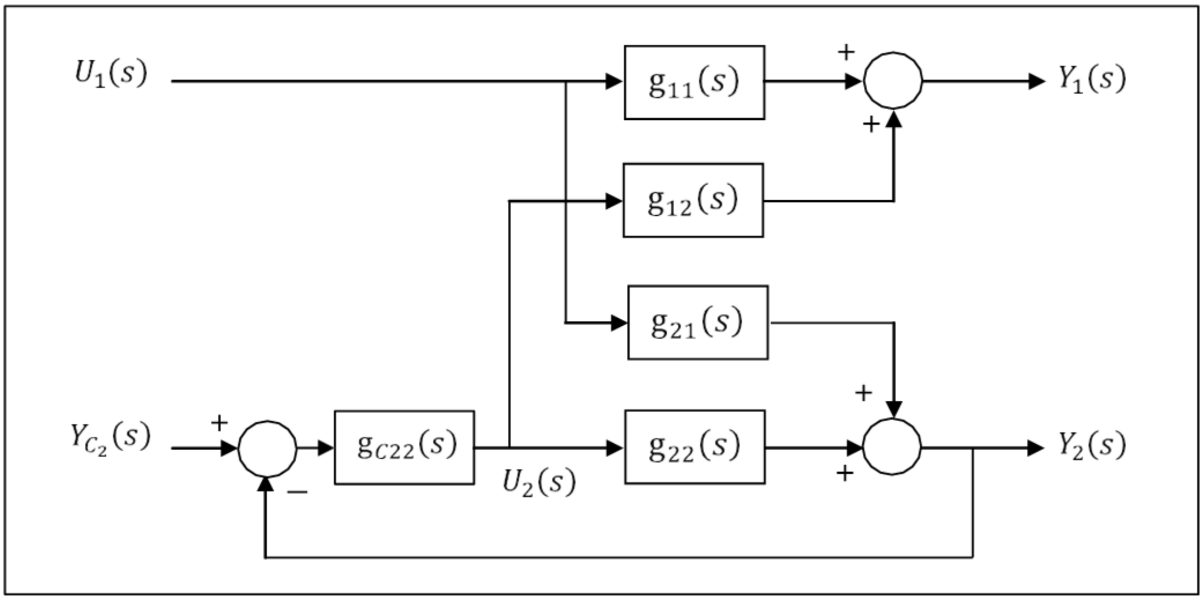


Figure IV.4 : (2×2) system with a controller on the second loop

We will have :

$$Y_1(s) = \left[g_{11}(s) - \frac{g_{12}(s)g_{C22}(s)g_{21}(s)}{1 + g_{22}(s)g_2(s)}\right]U_1(s) + \frac{g_{12}(s)g_{C22}(s)}{1 + g_{22}(s)g_{c22}(s)}Y_{c2}(s)$$

$$Y_2(s) = \frac{g_{21}(s)}{1 + g_{22}(s)g_{C22}(s)}U_1(s) + \frac{g_{22}(s)g_{c22}(s)}{1 + g_{22}(s)g_{C22}(s)}Y_{c2}(s)$$

The transfer function between the control input U_1 and the output Y_1 in the closed loop is:

$$\left(\frac{Y_1(s)}{U_1(s)}\right)_{b_{f22}} = \left[g_{11}(s) - \frac{g_{12}(s)g_{C22}(s)g_{21}(s)}{1 + g_{22}(s)g_{C22}(s)}\right]$$

b_{f22} : indicates that the loop is closed between Y_2 and U_2.

The ratio $\mu_{11}(S)$ between the open-loop and closed-loop transfer functions expresses the influence of the first loop on the second loop which is :

$$\mu_{11}(s) = \frac{g_{11}[1 \quad g_{22}g_{C22}]}{\boldsymbol{Y_1}(s) \quad g_{C22}(s)[g_{11}(s)g_{22}(s) + g_{12}(s)g_{21}(s)]}$$

The existence of an integrator being common in controllers implies that when $g_{c22}(S) = +\infty$, the measure of interaction $\mu_{11}(S)$, denoted as λ_{11}, depends only on the static gains of the transfer function $G(S)$. **[10]**

Which is :

$$\lambda_{11} = \frac{g_{11}(0)g_{22}(0)}{\boldsymbol{Y_1}(0)g_{22}(0) - g_{12}(0)g_{21}(0)}$$

Similarly, we find the other relative gains $(\lambda_{12}, \lambda_{21}, \lambda_{22})$. The result can then be represented using a matrix of relative gains :

$$\Lambda = \begin{bmatrix} \lambda_{11} & \lambda_{12} \\ \lambda_{21} & \lambda_{22} \end{bmatrix}$$

For a system (2×2) :

$$\Lambda = \begin{bmatrix} \lambda_{11} & 1 - \lambda_{11} \\ 1 - \lambda_{11} & \lambda_{11} \end{bmatrix}$$

A generic element of the relative gain matrix Λ of a multivariable system of higher dimension is equal to:

$$\lambda_{ij} = \frac{\left(\frac{dY_i}{dU_j}\right)_{U_k=0, k\neq j}}{\left(\frac{dY_i}{dU_j}\right)_{Y_l=0, l\neq j}}$$

IV.2.7.1. General procedure for calculating RGA

For a full-rank matrix G, the relative gain array is defined as follows:

$$\boldsymbol{\Lambda} = \boldsymbol{K_s}.\times (\boldsymbol{K_s^{-1}})^{\boldsymbol{T}}$$

This relation in matrix form was demonstrated by Skogestad in 1987 to facilitate calculations for systems larger than (2x2). **[10]**

with:

$$\Lambda = [\lambda_{ij}.\, i,j = 1, \ldots., m].$$

$$K_s = [k_{ij}.\, i,j = 1, \ldots., m].$$

$.\times$: is the product of Hadamard (the element-wise product)

K_s : is the matrix of static gains.

K_{ij} : is the static gain between u_i and y_j.

The elements of K_s are calculated as follows :

$$K_{ij} = \lim_{s\to\infty} g_{ij}(S)$$

IV.2.7.2. Properties of the Relative Gain Array (RGA)

The algebraic sum of the elements of the RGA along a row i or a column j is equal to 1.

$$\sum_{j=1,i=cst}^{m} \lambda_{ij} = 1$$

$$\sum_{j=cst,i=1}^{m} \lambda_{ij} = 1$$

- Permutations on a row or column of G give the same permutations in RGA .
- If G is triangular, the relative gain matrix reduces to the identity matrix.

- RGA (G) = RGA (D_1GD_2) if D_1 and D_2 are diagonal.

IV.2.7.3. Interpretation of the Relative Gain Array Matrix

$\lambda_{ij} = 1$: The open-loop and closed-loop gains are identical. Interactions do not affect the input-output pair $[U_j - Y_i]$, meaning that the other loops do not influence loop $[U_j, Y_i]$.

$\lambda_{ij} = 0$: The open-loop gain is equal to zero, meaning U_j does not affect Y_i.

$\lambda_{ij} < 0$: The open-loop and closed-loop gains have different signs. In this case, the corresponding loop may change its direction of variation (inverse response system) if the other loops are closed. Most importantly, negative elements on the diagonal can introduce integral instability.

$\lambda_{ij} \gg 1$: Control is very difficult due to strong interactions.

$\lambda_{ij} = 0.5$: The degree of interaction is high, and the other loops have the same effect on the output Y_j as the input U_i.

In summary, choosing the best control configuration involves selecting input-output pairs such that λ_{ij} is closest to 1 while avoiding pairs with negative relative gains.

Remark :

Relative gains are independent of scale and units of measurement. In the case where the system is not square, meaning the number of inputs p is different from the number of outputs q, the calculation of the RGA is done using the pseudo-inverse of K_s^+.

IV.3. Correctors in multivariable control

In the context of multivariable control systems, correctors play a crucial role in managing interactions between system variables and enhancing overall stability and performance. These correctors, such as PID controllers, are designed to adjust control inputs based on feedback and system conditions to achieve desired setpoints.

Correctors, also known as controllers, are essential components in multivariable control systems. They are tasked with continuously monitoring system variables and making real-time adjustments to maintain optimal operation.

In multivariable systems, interactions between different variables can complicate control efforts. For instance, changes in one variable may influence others due to process dynamics and coupling effects. Correctors mitigate these interactions by adjusting control actions in a coordinated manner, ensuring that each variable responds appropriately to achieve overall system stability and performance objectives.

IV.3.1. Interest of Correctors

A controller aims to reduce the error between the setpoint and the measured output to a value very close to zero, regardless of disturbances in the case of regulation affecting the system or changes in setpoint in the case of tracking.

Sizing a controller allows imposing desired behavior on the controlled system, which involves meeting several criteria often specified in specifications:

- Ensuring stability of the closed-loop system,
- Canceling or reducing steady-state error to a setpoint change,
- Minimizing external disturbances,
- Ensuring the system responds quickly to setpoint changes (reducing response time),
- Minimizing the impact of process parameter variations,
- Ensuring the controller is robust and does not require frequent adjustments.

IV.3.2. Proportional Integral Derivative (PID) Controller

The PID controller is the combination of three actions: Proportional, Integral, and Derivative. The structure of this controller depends on the three coefficients K_p, K_i, and K_d. The presence of the integral action increases the system's type, and the presence of the derivative action improves stability.

PID corrector transfer function is :

$$G(S) = K_p\left(1 + \frac{1}{T_i S} + \frac{T_d S}{1 + \frac{T_d}{N}}\right)$$

IV.3.3. Setting PID Controller Parameters

To meet the requirements imposed by a specification, one must calculate the parameters of the standard PID controller, which encompasses all actions used in control systems. The calculation of these parameters, known as controller synthesis, involves finding the constants K_p, K_i, and K_d of the PID controller. These parameters ensure the desired behavior or performance in closed-loop control, as specified by the specification. For example, these may include maximum overshoot, response time, rise time, etc. There are several methods available for calculating these parameters. For our study, we focus on the Ziegler-Nichols method for single-variable systems and the BLT method for multivariable systems.

IV.3.3.1. Practical tuning using the Ziegler and Nichols method

This method, established around 1942, is the most well-known practical method for tuning control loops. It applies to both self-regulating processes and integrator processes, provided that it is possible to induce sustained oscillations. To achieve the limit of oscillation, a proportional controller is placed in the closed loop, and the gain K_p of this controller is gradually increased until sustained oscillations (pumping phenomenon) are obtained. The critical gain K_{cr} that brings the system to the stability limit and the period T_{cr} of the obtained oscillations are noted.**[11]**

The settings recommended by Ziegler and Nichols, based on the regulator structure used, are given in the following table:

Table IV.1 : Settings recommended by Ziegler and Nichols

Régulateur	K_p	T_i	T_d
P	$0.5\ K_{cr}$		
PI	$\frac{K_{cr}}{2,2}$	$\frac{T_{cr}}{1,2}$	
PID	$0,6\ K_{cr}$	$0,5\ T_{cr}$	$0,125\ T_{cr}$

These adjustments result in a relatively aggressive tuning, characterized by a step response with a first overshoot D_1 of around 30% to 40%.

These values might not meet the specifications if the first overshoot D_1 is too significant, necessitating slight modifications to these adjustments.

IV.3.3.2. Biggest Log-Modulus Tuning (BLT) Method

The Ziegler and Nichols method allows for quick determination of PID controller parameters, but its application is limited to single-variable systems only. For the regulation of multivariable systems, the Biggest Log-Modulus Tuning (BLT) method has been proposed. This method is a generalization of the Ziegler and Nichols method, proposed by Luyben in 1990. It allows for the calculation of each controller's parameters and ensures system stability. The synthesis of PID controllers in a multivariable environment using the BLT method involves the following steps: **[11]**

- **Step 1**

Calculate the PI controller parameters using the Ziegler and Nichols method for each individual loop. Begin by determining the critical frequency W_{cr}, which is the frequency corresponding to the $-\pi$ phase shift. Identify the critical gain K_{cr} corresponding to the critical frequency, then calculate the Ziegler and Nichols parameters using the following formulas:

$$K_{ZN} = \frac{K_{cr}}{2.2}$$

$$T_{ZN} = \frac{2\pi}{1.2W_{cr}}$$

- **Step 2**

We choose an adjustment factor $1.5 < F < 4$, we divide the gains of the correctors by F :

$$K_c = \frac{K_{ZN}}{F}$$

and the time constants of the controllers are multiplied by the factor F :

$$T_i = T_{ZN}F$$

- **Step 3**

Consider the following function:

$$W(j\omega) = -1 + Det[I + G(j\omega)G_c(j\omega)]$$

With :

$G(S)$: transfer function matrix of the system.

$G_c(S)$: diagonal transfer function matrix of the controllers.

I : identity matrix

$$G_c(S) = \begin{bmatrix} g_{c11}(S) & 0 & \cdots & 0 \\ 0 & g_{c22}(S) & \cdots & 0 \\ \vdots & \vdots & \ddots & \vdots \\ 0 & 0 & \cdots & g_{cnn}(S) \end{bmatrix}$$

The transfer function of the controllers is of type (PI) :

$$g_{cii} = K_p\left(1 + \frac{1}{T_iS}\right)$$

To bring the system to the stability limit, we bring the function $W(j\omega)$ to the Nyquist point (-1,0). By analogy with single-variable systems, we define the multivariable closed-loop module.

$$L^{cm}(j\omega) = 20\log\left|\frac{W}{1+W}\right|$$

- **Step 4**

We vary the tuning factor F until L^{cm}_{max} is equal to $2n$, where n is the order of the multivariable system. For a 2×2 system, the factor $L^{cm}_{max} = 2 \times 2 = 4$. This method allows for the design of a stable multiloop control system with reasonable performance. Equation [III.17] is the characteristic equation of the closed-loop multivariable system, which rigorously determines the stability of the closed-loop system.

Rule :

A unit feedback control system is stable, as shown in Figure IV.6.a, if, when plotting the Nyquist plot $A(j\omega)$ of the open-loop transfer function in the direction of increasing frequencies, the critical point (coordinates (-1,0)) is to its left. It is unstable, as shown in Figure IV.5.b, if this is not the case.**[12]**

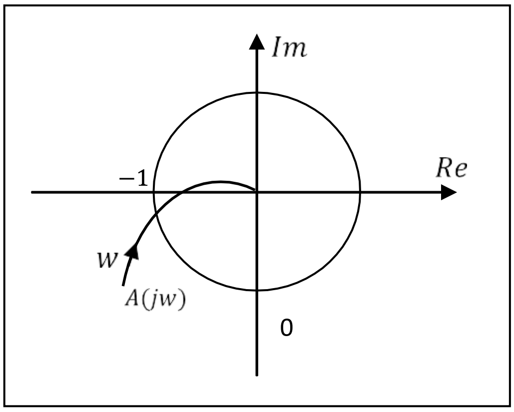


Figure IV.6 : a.Stable system

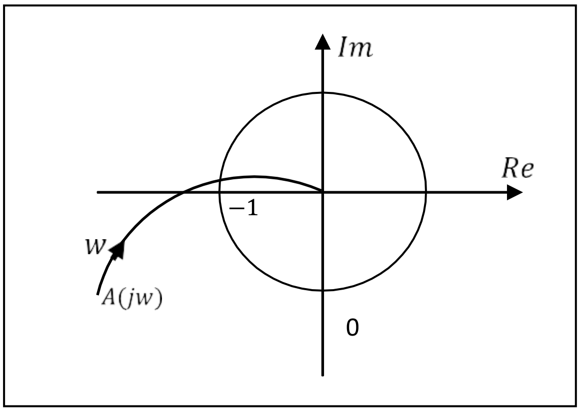


Figure IV.5 : b.Unstable system

It is important to confirm the shape of the Nyquist plot of the function W when applying this method, because function simply indicates the proximity to the critical point. An unstable system could be regulated at a maximum gain of 2N, as it surrounds the point (-1,0i), but is not close to it.

IV.4. Multivariable control of a distillation column

Distillation columns represent complex multivariable systems where precise control of temperature, pressure, flow rates, and composition is essential for efficient separation of components. Maintaining optimal operating conditions and ensuring consistent product quality require advanced control strategies and techniques.

In this context, a PID (Proportional-Integral-Derivative) controller, tuned using the Biggest Log-Modulus Tuning (BLT) method, is employed to regulate the distillation column. The BLT method, a generalization of the Ziegler-Nichols method, is particularly suitable for multivariable systems like distillation columns.**[10]**

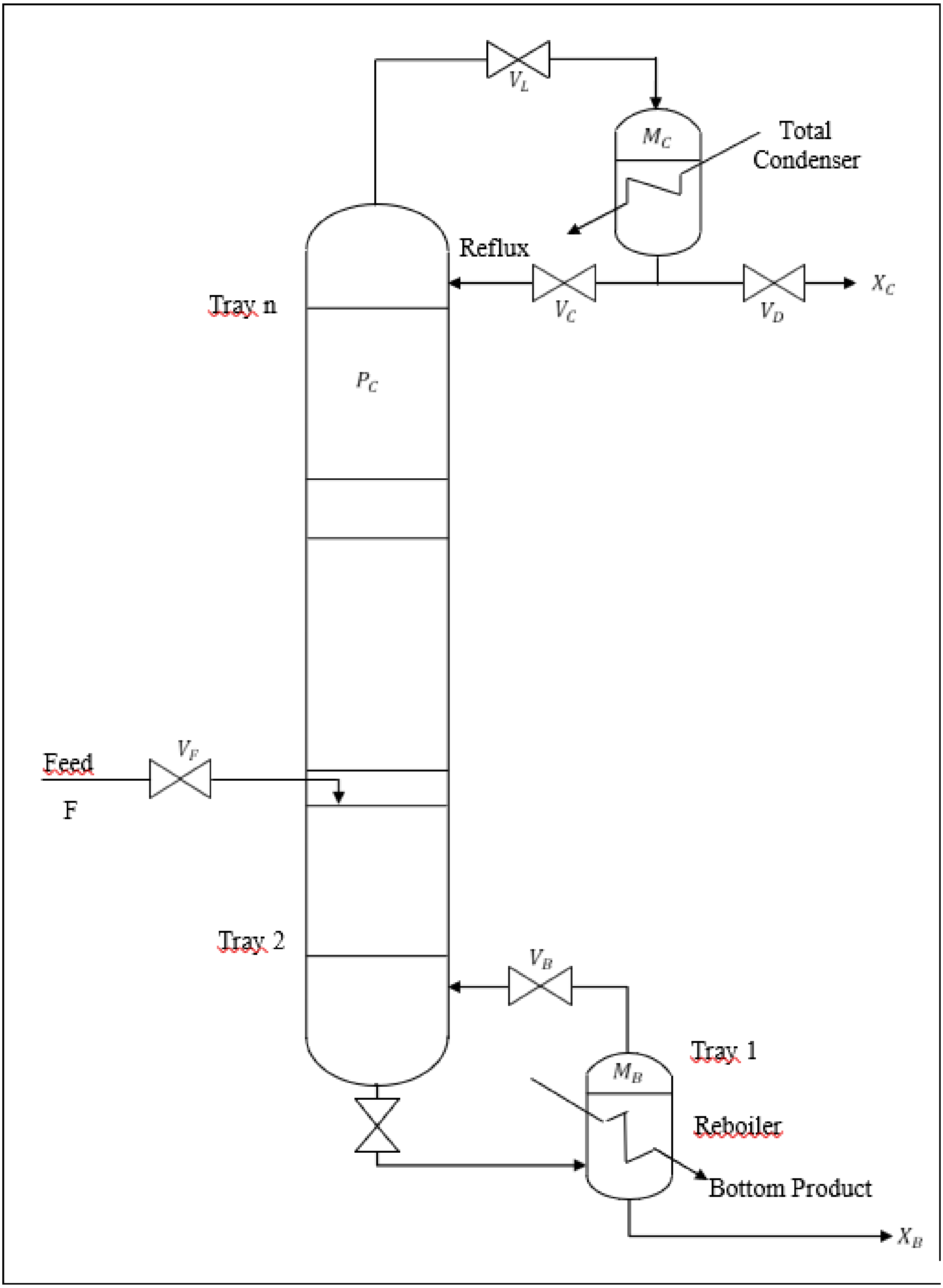


Figure IV.7 : Distillation Column

IV.4.1. Mathematical Model of the Distillation Column

The main role of a distillation column is to separate liquid mixtures and bring the column compositions to desired specifications, for example, by regulating the bottom vapor pressure (BVP) of the column, while adjusting the temperature, level, and pressure of the column. To model the column, the system is represented in a block diagram form, and its inputs and outputs are defined.

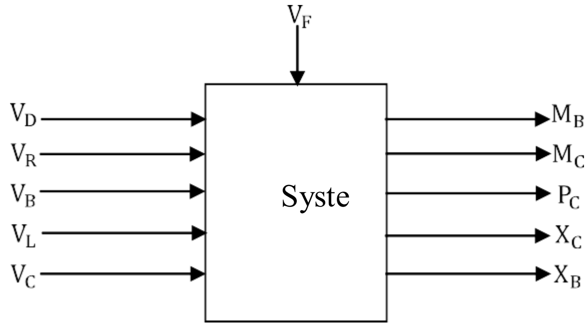


Figure IV.8 : Column block diagram

The distillation column shown in Figure IV.7 is represented as a block diagram in Figure IV.8, containing five control inputs, five outputs, and one disturbance variable. The five control variables correspond to the five manipulable valves of the column:

1. distillate flow rate from the condenser V_D.
2. residue flow rate from the reboiler V_R.
3. reflux flow rate at the top of the column V_L.
4. vapor flow rate at the reboiler V_B.
5. vapor flow rate at the top of the column V_C.

The five output variables that need to be controlled to maintain operational stability are:

1. reboiler hold-up M_B.
2. condenser hold-up M_C.
3. column hold-up or pressure P_C.
4. top column composition X_C.
5. bottom column composition X_B.

The disturbances in the column are generally related to the feed.

- Feed enthalpy, characterized by the flow rate F.

Obtaining a reduced linear model is often sought. It can be done at various levels of complexity, depending on the objective. For this system, the goal is to maintain the desired specifications of the compositions at the bottom and top of the column, so the two outputs X_B and X_C are controlled by the two inputs V_L and V_B and disturbed by V_F. Therefore, the simplified system is of order (2×2), as shown in figure IV.9.

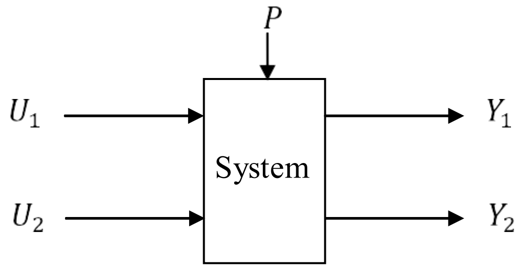


Figure IV.9 : Schéma bloc d'un système (2×2)

The inputs $U_1 = V_L$, $U_2 = V_B$, the outputs $Y_1 = X_C$, $Y_2 = X_B$, and the disturbances $P = F$.

Authors have identified the transfer functions of the distillation column considered as a system (2×2) with inputs being the reflux and vapor flow rates to the reboiler, and outputs being the mole fractions of the distillate from the top and bottoms from the reboiler. Subsequently, the influence of disturbances from the feed flow rate was added to this model. The complete model of the system is a transfer function matrix.

$$\begin{bmatrix} Y_1 \\ Y_2 \end{bmatrix} = \begin{bmatrix} \frac{12,8}{1+16,7s} e^{-s} & \frac{-18,9}{1+21s} e^{-3s} \\ \frac{6,6}{1+10,9s} e^{-7s} & \frac{-19,4}{1+14,4s} e^{-3s} \end{bmatrix} \begin{bmatrix} U_1 \\ U_2 \end{bmatrix} + \begin{bmatrix} \frac{0,7}{1+14,4s} e^{-5s} \\ \frac{1,3}{1+12s} e^{-3s} \end{bmatrix} P$$

With :

$Y_1(S)$: The first output of the system.

$Y_2(S)$: The second output of the system.

$U_1(S)$: The first input of the system.

$U_2(S)$: The second input of the system.

$P(S)$: The disturbance.

To design the control system, we will first use the Relative Gain Array (RGA) method to choose a configuration characterized by low interaction levels. Then, using the Biggest Log-modulus Tuning (BLT) method, we will calculate the parameters of the PI controllers.

IV.4.2. Choosing the system configuration

The system variables exhibit interactions among themselves. To analyze these interactions, we will use the Relative Gain Array (RGA) matrix. For our system, which is second-order with two inputs and two outputs, there are two possible configurations. The first configuration involves controlling the first output with the first input and the second output with the second input.

$$U_1 \rightarrow Y_1$$

$$U_2 \rightarrow Y_2$$

The second configuration involves controlling the first output with the second input and the second output with the first input.

$$U_2 \rightarrow Y_1$$

$$U_1 \rightarrow Y_2$$

- Calculation of the steady-state gain matrix of the model :

$$K_s = \lim_{S\to\infty} G(S)$$

$$K_s = \begin{bmatrix} \lim\limits_{S\to\infty} g_{11}(S) & \lim\limits_{S\to\infty} g_{12}(S) \\ \lim\limits_{S\to\infty} g_{21}(S) & \lim\limits_{S\to\infty} g_{22}(S) \end{bmatrix}$$

After Calculation :

$$K_s = \begin{bmatrix} 12.8 & -18.9 \\ 6.6 & -19.4 \end{bmatrix}$$

- Calculation of the inverse of the matrix K_s :

$$K_s^{-1} = \begin{bmatrix} 0.1570 & -0.1529 \\ 0.0534 & -0.1036 \end{bmatrix}$$

- Calculating the transpose of the matrix K_s^{-1} :

$$[K_s^{-1}]^T = \begin{bmatrix} 0.1570 & 0.0534 \\ -0.1529 & -0.1036 \end{bmatrix}$$

- Calculation of the relative gains matrix :

$$RGA = K_S . \times [K_S^{-1}]^T$$

$$RGA = \begin{bmatrix} 2.0094 & -1.0094 \\ -1.0094 & 2.0094 \end{bmatrix}$$

The calculated relative gain matrix shows λ_{11} a value greater than 1, indicating that this input/output pairing results in a stable system but with strong interactions. In the case of the second configuration, it is noted that λ_{12} is negative, indicating that this pairing makes the system unstable. Therefore, the best configuration for this system is the first one.

$$U_1 \rightarrow Y_1$$

$$U_2 \rightarrow Y_2$$

IV.4.3. Controller Synthesis

After choosing the configuration, we will calculate the parameters of the PI controllers using the Biggest Log-Modulus Tuning (BLT) Method.

The chosen configuration specifies the two elements of the transfer function matrix to be controlled.

$$g_{11}(S) = \frac{12.8}{1 + 16.7S} e^{-S}$$

$$g_{22}(S) = \frac{-19.4}{1 + 14.4S} e^{-3S}$$

According to the BLT method, we will calculate a PI controller for each individual loop, specifically for the subsystems $g_{11}(S)$ and $g_{12}(S)$.

To calculate the various parameters of the PI controllers, we have implemented the the Biggest Log-Modulus Tuning (BLT) Method. using MATLAB as follows:

The Matlab function `margin` allows us to calculate the critical gain and the corresponding frequency.

For $g_{11}(S)$ we will have :

$$\omega_{cr1} = 1.60\ rad/s$$

$$K_{cr1} = 2.09$$

For $g_{12}(S)$ we will have :

$$\omega_{cr2} = 1.58\ rad/s$$

$$K_{cr2} = 1.17$$

Then we calculate the Ziegler-Nichols parameters using the following relationships:

$$K_{ZN} = \frac{K_{cr}}{2.2}$$

$$T_{ZN} = \frac{2\pi}{1.2W_{cr}}$$

Our results are :

- Controller $g_{C11}(S)$:

The proportional gain of the controller $K_{c_{11}} = 0.31$ (without unit)

The integration constant of the controller $T_{i_{11}} = 10.17_{mn}$

- Controller $g_{C12}(S)$:

The proportional gain of the controller $K_{c_{22}} = -0.17$ (without unit)

The integration constant of the controller $T_{i_{22}} = 10.31_{mn}$

The two obtains controllers are :

$$g_{C11}(S) = \frac{3.11S + 0.31}{10.17S}$$

$$g_{C22}(S) = -\frac{1.77S + 0.17}{10.31S}$$

The obtained detuning factor is $F = 3.1223$, which gives the maximum loop gain value of the system $L_{max} = 4.0002$.

IV.4.4. Simulation results

IV.4.4.1. Simulation diagram

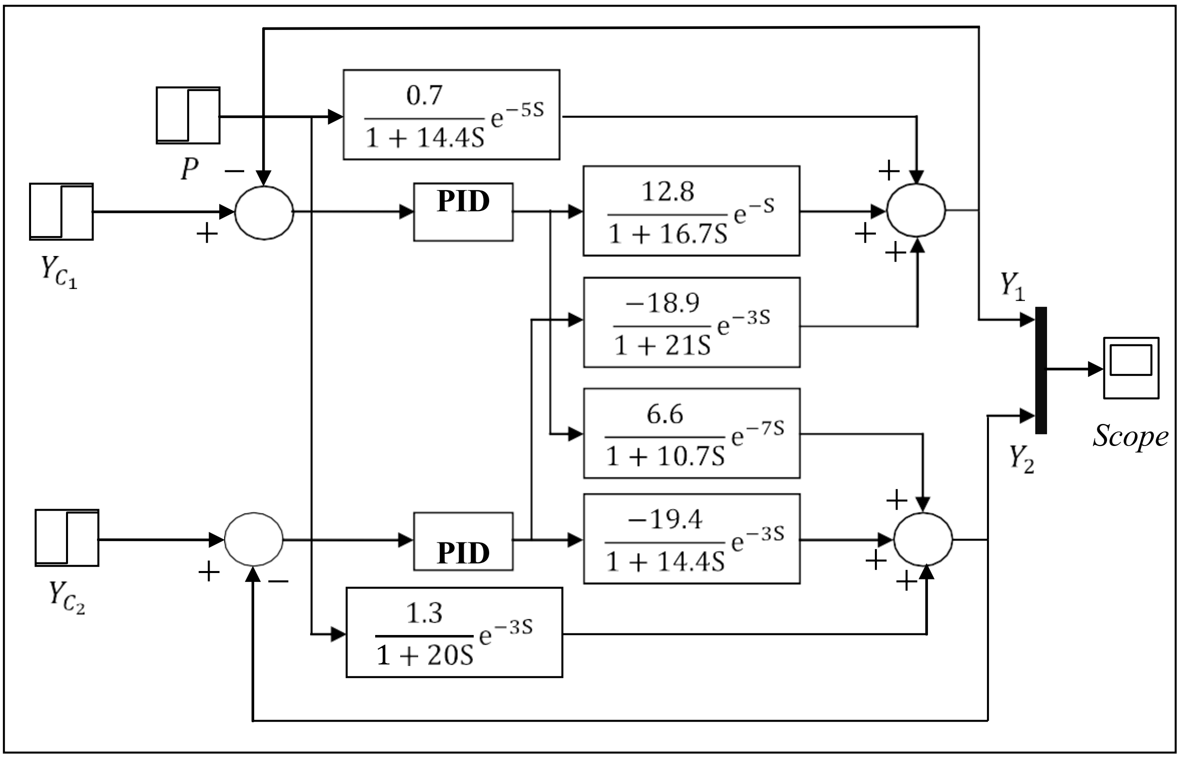


Figure IV.10 : Simulink simulation block diagram

The simulation of the system allowed us to observe the system's behavior and highlight the effect of interactions between the characteristic variables.

IV.4.4.2. Effects of interactions

IV.4.4.2.1. Change in the first setpoint

To observe the effect of setpoint Y_{c1} on the second output Y_{c2}, we increase Y_{c1} by a step of 1 unit at time t = 20 seconds while keeping the second setpoint and disturbances at zero. The responses, setpoints, and controller outputs of the system for this setpoint change are shown in figures IV.11 and IV.12.

The change in Y_{c1} setpoint results in a change in control signal u_1, which brings the output Y_1 to the desired value Y_{c1}. The response Y_1 exhibits some oscillations during the transient regime, lasting about 80 seconds, with maximum overshoots of 6%. The steady-state response is established by t = 100 seconds.

Manipulating the first setpoint affects not only the first output but also the second. The steady-state establishment of the second output occurs approximately 100 seconds after the setpoint change.

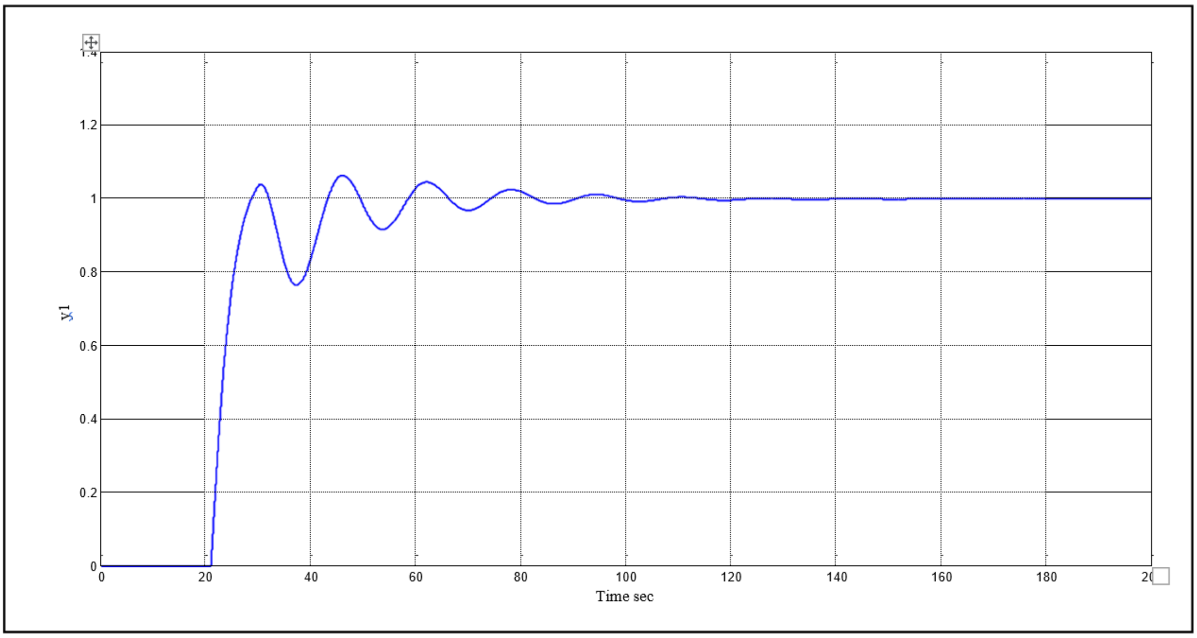


Figure IV.11 :Behavior of the 1st loop upon application of Yc1

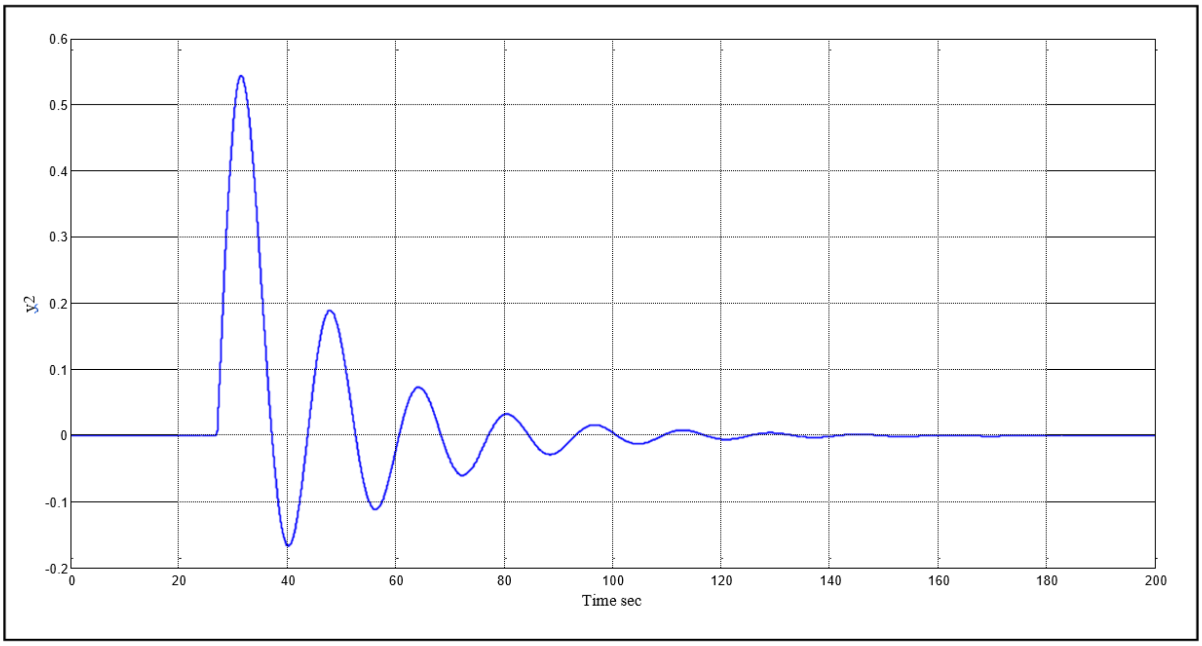


Figure IV.12 : Behavior of the 2nd loop upon application of Yc1

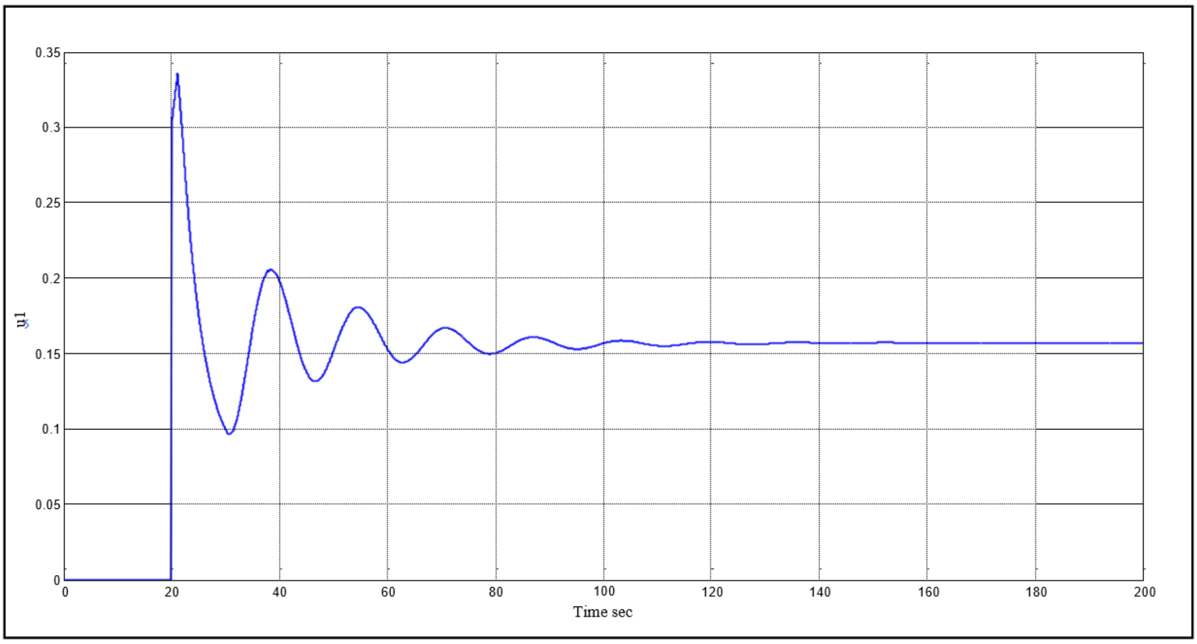


Figure IV.13 : The control u1 upon application of Yc1

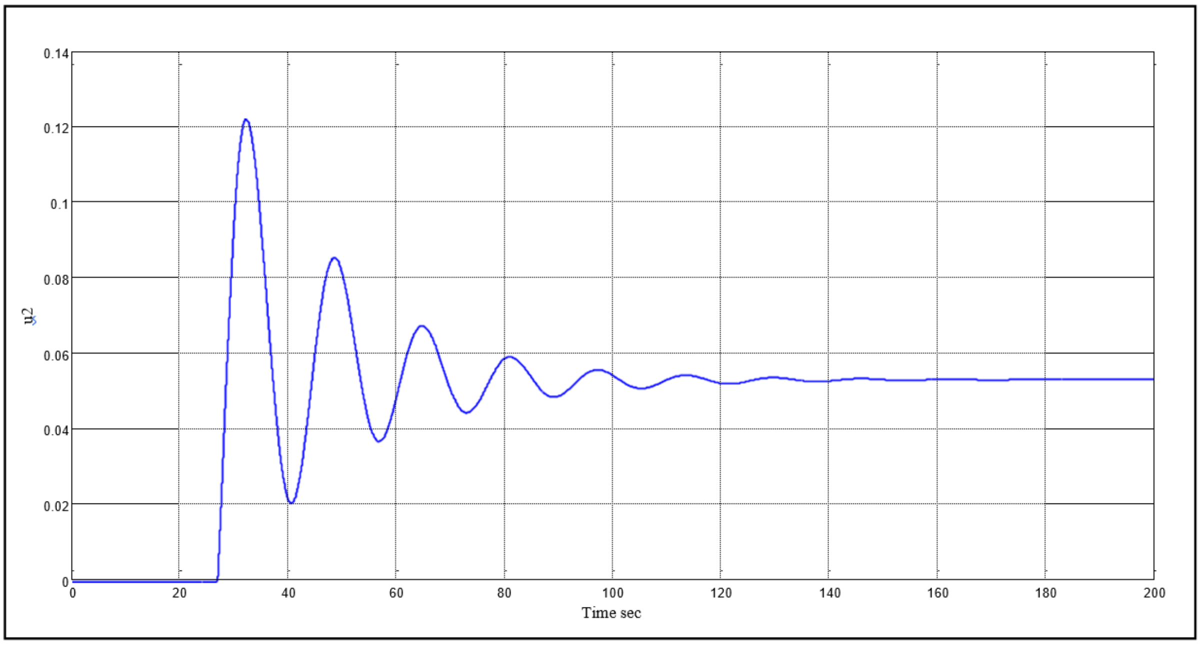


Figure IV.14 : The control u2 upon application of Yc1

IV.4.4.2.2. Change in the second setpoint

To observe the effect of setpoint Y_{c2} on the first output Y_1, we apply a unit step $Y_{c2} = 1$ at $t = 20$ seconds. The responses, setpoints, and control signals of the system are shown in figures IV.15 and IV.16.

The step change in the second input Y_{c2} causes a change in control signal u_2, which brings the output Y_2 to the desired value. During the transient regime, the response Y_2 is somewhat more oscillatory than the first response, with a maximum overshoot of 30%, and the steady state is established at $t = 120$ seconds.

The variation in control signal u_2 causes a disturbance in output Y_1, which reaches its steady state at $t = 100$ seconds, and the effect of the interaction disappears.

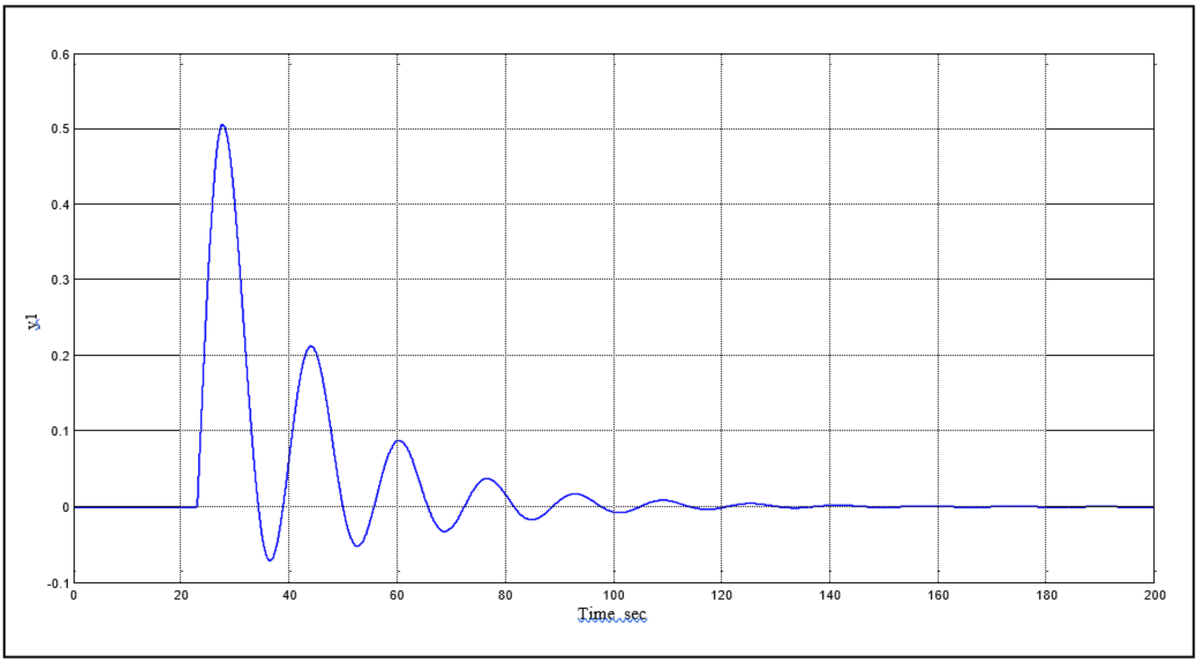


Figure IV.15 : Behavior of the 1st loop upon application of Yc2

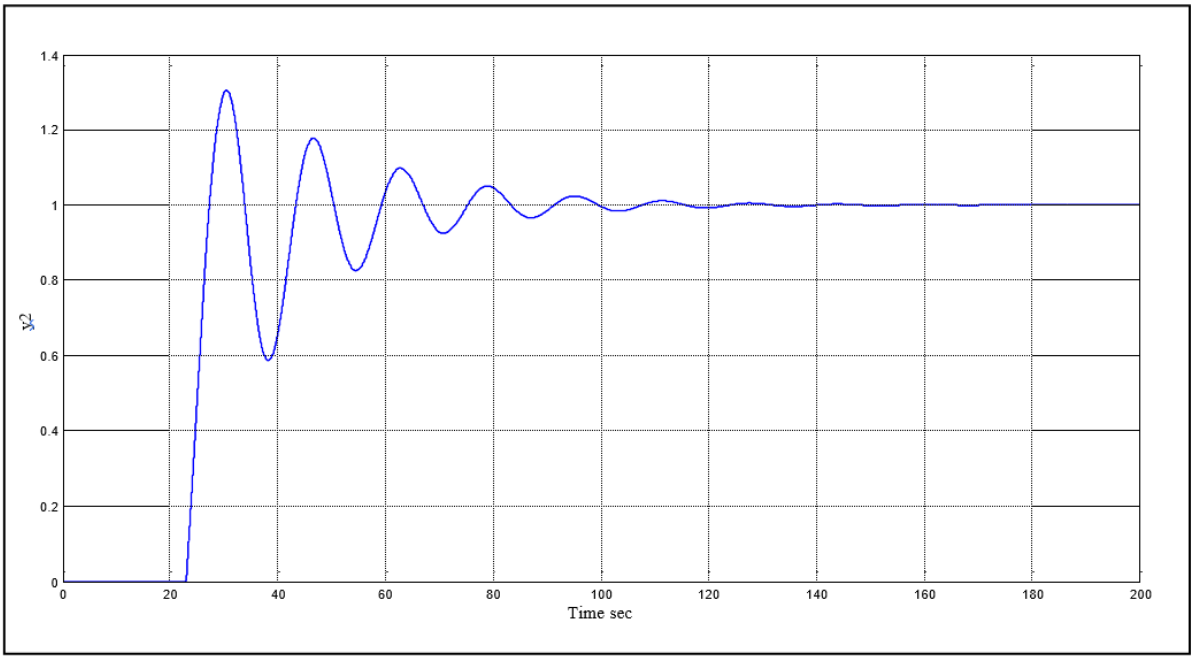


Figure IV.16 : Behavior of the 2nd loop upon application of Yc2

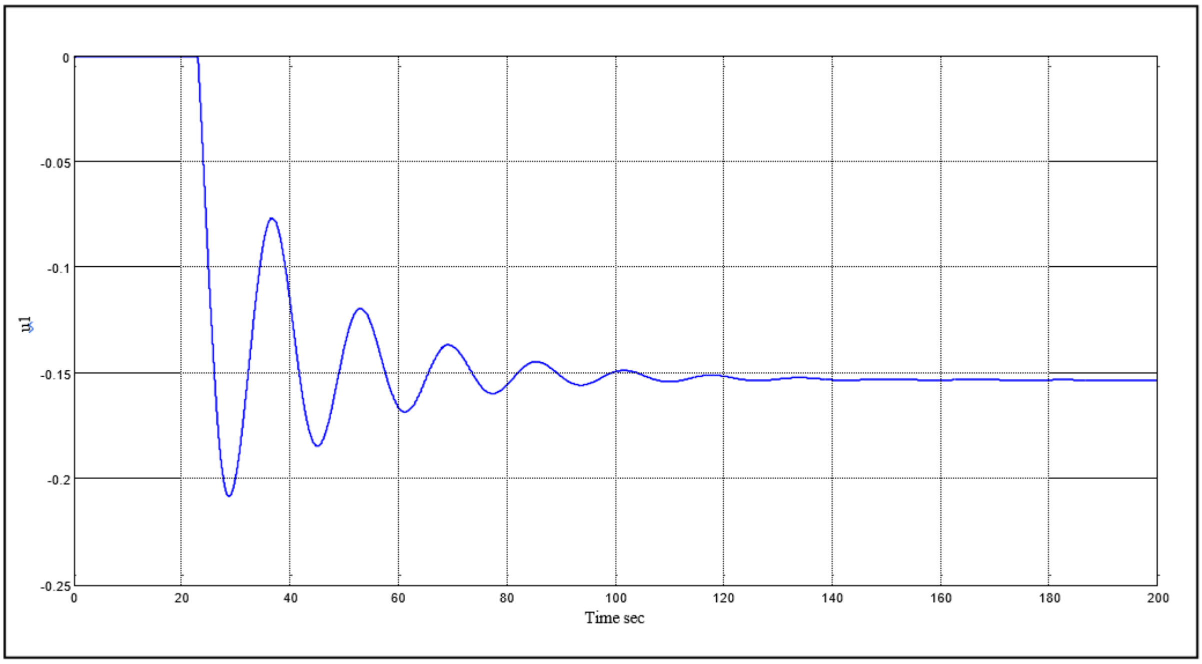
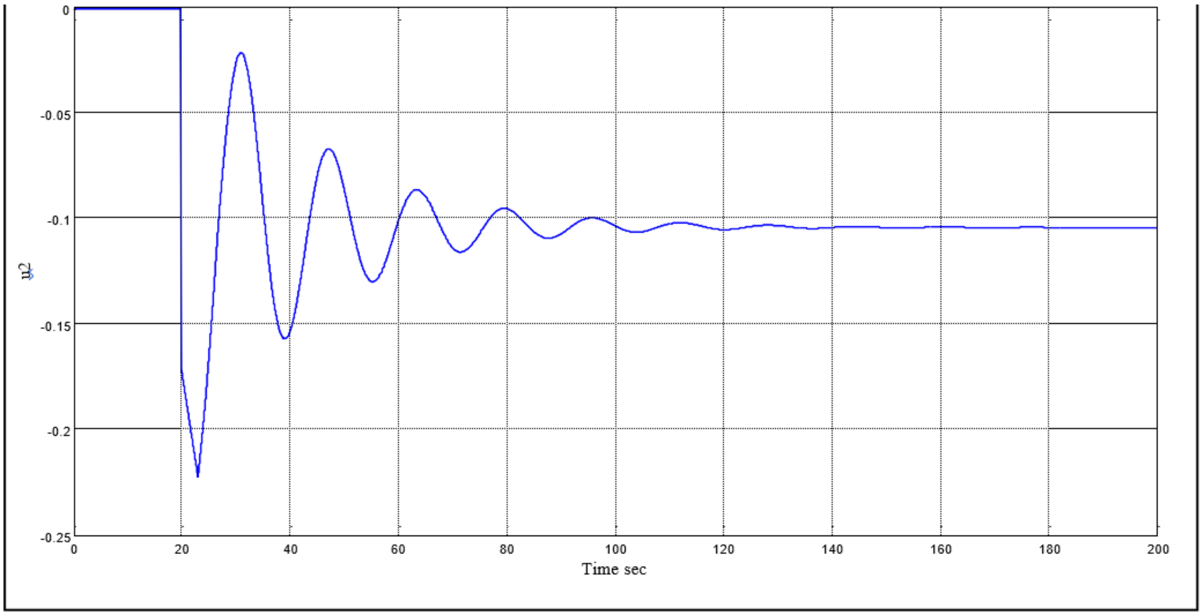


Figure IV.17 : The control u2 upon application of Yc2

Figure IV.18 : The control u2 upon application of Yc2



IV.4.4.2.3. Effects of disturbances

To evaluate the system's reaction to disturbances, we apply two unit step inputs to both entries at $t = 0$ seconds. When the steady state is fully established at $t = 150$ seconds, we apply a step disturbance $P = 0.5$ (unit of measurement). The system's behavior in response to the disturbance is shown in figures IV.19 and IV.20.

The output Y_1 is affected by the disturbance with a small amplitude and for a duration of 50s. At $t = 200$ seconds, the disturbance is completely rejected, and the output reaches the desired value $Y_1 = Y_{c1}$.

The second output Y_2 is also affected by the disturbance, with a slightly higher amplitude than the first output. By $t = 200$ seconds, the effect of the disturbance is canceled on the second output Y_2.

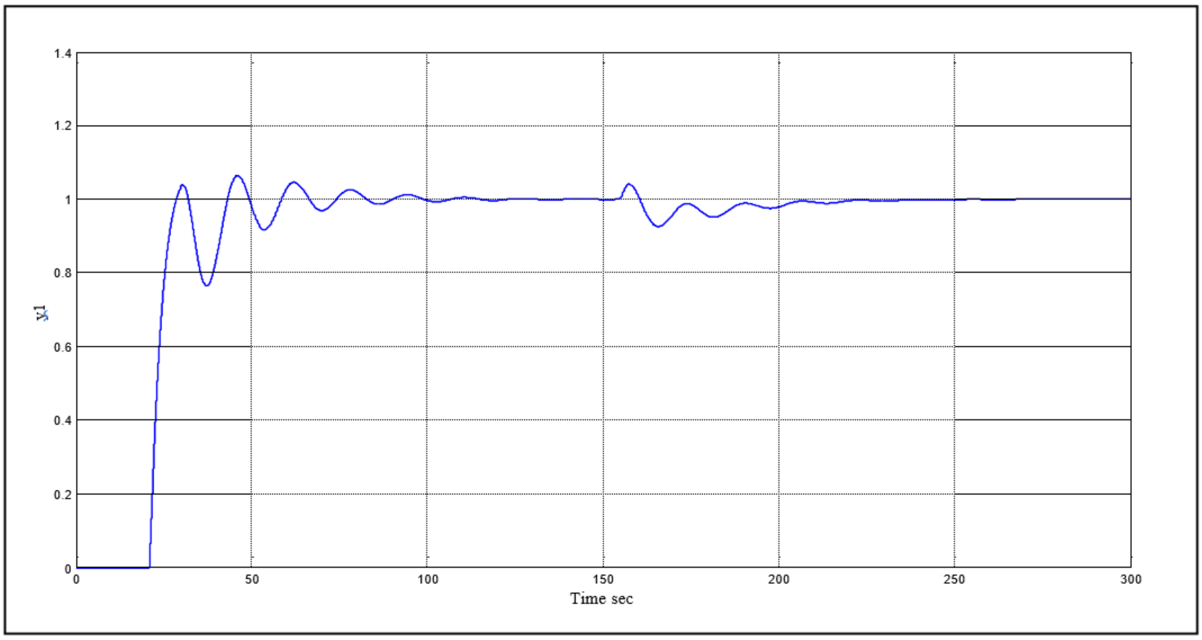


Figure IV.19 : Behavior of the 1st loop in response to disturbances

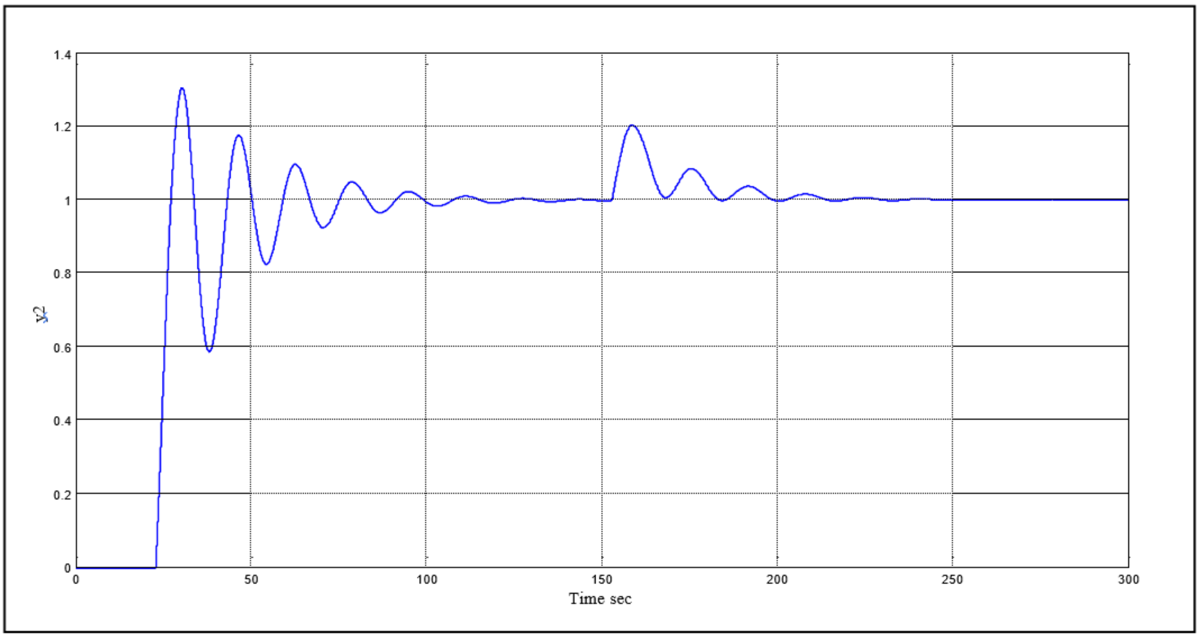


Figure IV.20 : Behavior of the 2nd loop in response to disturbances

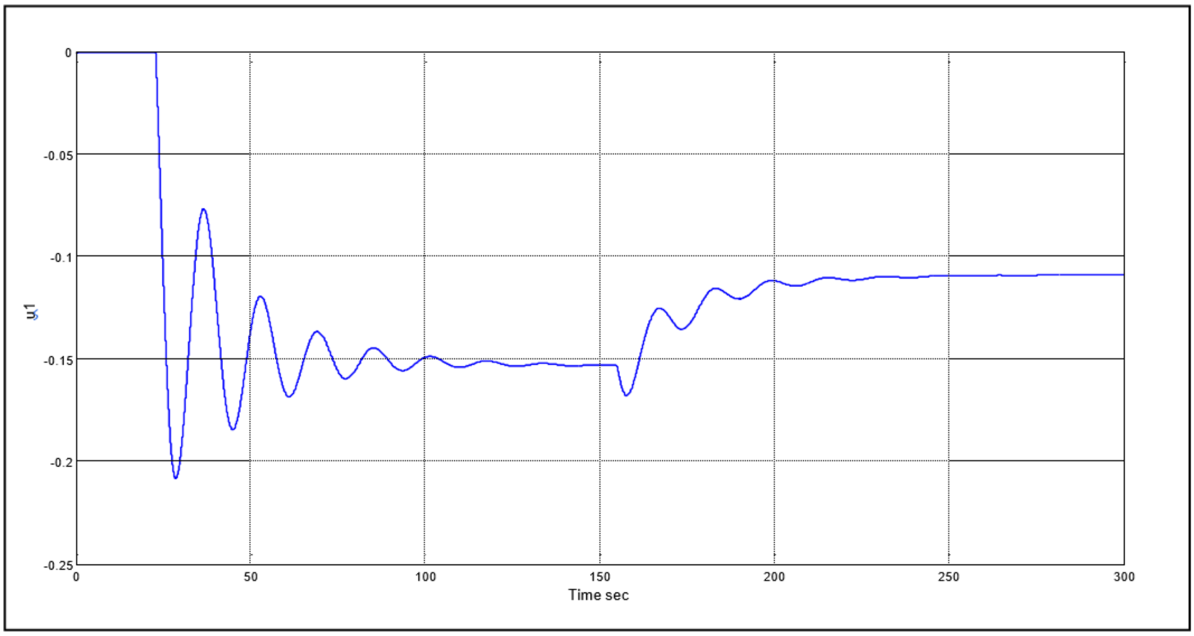


Figure IV.21 : The control u1 in response to disturbances

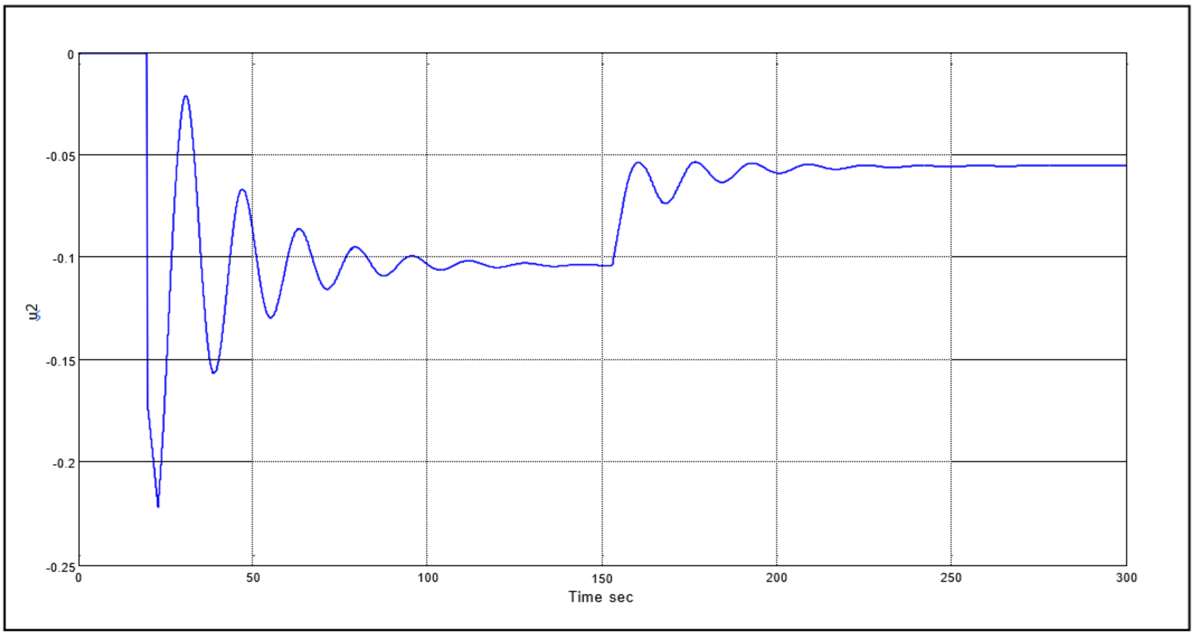


Figure IV.22 : The control u2 in response to disturbances

IV.4.4.3. Intrepretation and comments :

In analyzing the multivariable control of a distillation column, we observe significant interactions between outputs when setpoints change and disturbances occur. When the first setpoint Y_{c1} is increased by a unit step at t = 20 seconds, the first control signal $u_1$adjusts to bring Y_1 to the new setpoint, causing Y_1 to exhibit oscillations with a 6% overshoot, stabilizing by t = 100 seconds. The second output Y_2, although its setpoint Y_{c2} remains unchanged, is affected by this change and stabilizes around the same time. Conversely, applying a unit step change to Y_{c2} at t = 20 seconds results in $u_2$adjusting to bring Y_2 to the new setpoint, leading to a 30% overshoot and stabilization by t = 120 seconds. This change in Y_{c2} also disturbs Y_1 which reaches steady state by t = 100 seconds. The control signals u_1 and $u_2$dynamically interact to manage these setpoint changes. When evaluating the system's reaction to disturbances, a step disturbance P = 0.5 applied at t = 150 seconds initially causes small amplitude deviations in Y_1 and slightly higher amplitude changes in Y_2 . Both outputs manage to reject the disturbance by t = 200 seconds, demonstrating the control system's robustness.

Despite the application of PID controllers tuned via the BLT method, the complexity of the distillation column control and the significant interactions highlight the necessity for advanced control strategies like Advanced Process Control (APC), which will be explored in the next section. In this upcoming section, we will calculate the Relative Gain Array (RGA) between Manipulated Variables (MVs) and Controlled Variables (CVs) of the APC in the depropanizer column, and validate that the APC uses the RGA to adjust control strategies effectively, ensuring smoother operation and improved efficiency.

IV.4.5. Controlling Depropanizer with CVs, MVs and DVs

Table IV.2 : Transfer Fonctions Between CVs and MVs

	MV1	**MV2**	**MV3**
CV1	$G(s) = -35\frac{8.57S+1}{10.5S^2+6.49S+1}e^{-0S}$		
CV2		$G(s) = -0.000171\frac{1}{150S^2+22S+1}e^{-60S}$	
CV3		$G(s) = -0.018\frac{1}{100S^2+16S+1}e^{-16S}$	

CV4		$G(s) = -0.01\frac{1}{100S^2 + 16S + 1}e^{-10S}$	
CV5			$G(s) = -0.5\frac{1}{8S + 1}e^{-0S}$
CV6			$G(s) = 1.2\frac{1}{100S^2 + 16S + 1}e^{-0S}$

Table IV.3 : Transfer Fonctions Between CVs and DVs

	DV1	**DV2**	**DV3**
CV1		$G(s) = 0.5\frac{1}{5S + 1}e^{-0S}$	
CV2	$G(s) = 3.5e^{-5}\frac{1}{5S + 1}e^{-0S}$		
CV3	$G(s) = 0.0032\frac{1}{5S + 1}e^{-0S}$		
CV4			
CV5			$G(s) = 0.0005\frac{13S + 1}{8S^2 + 7.3S + 1}e^{-0S}$
CV6			$G(s) = 0.002\frac{13S + 1}{8S^2 + 7.3S + 1}e^{-0S}$

IV.4.5.1. APC Transfer Function Diagrams for Controlling Depropanizer Column

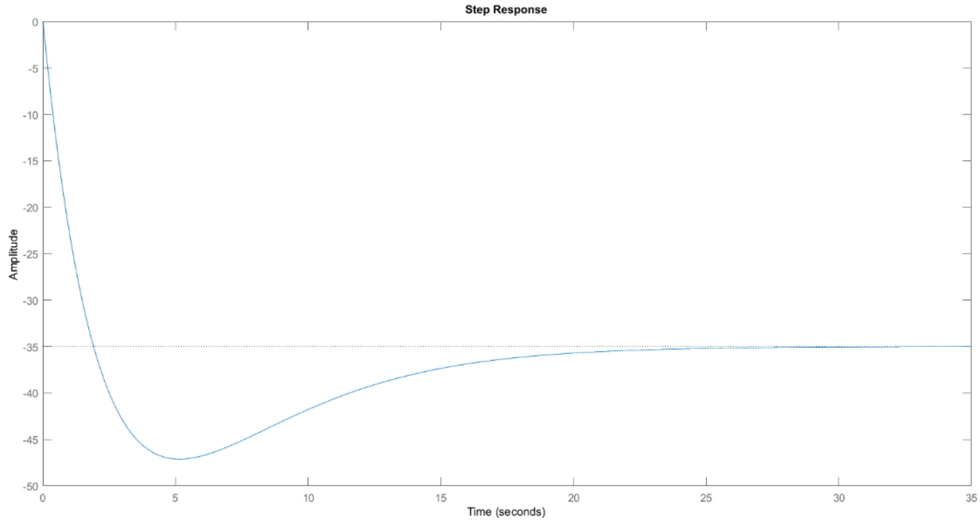


Figure IV.23 : CV1 , MV1 Transfer Function Diagram

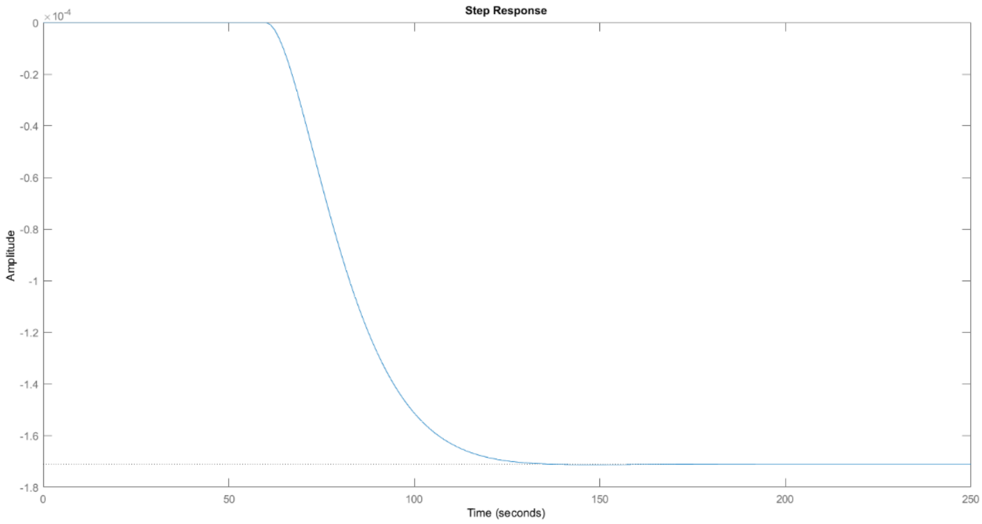


Figure IV.24 : CV1 , MV1 Transfer Function Diagram

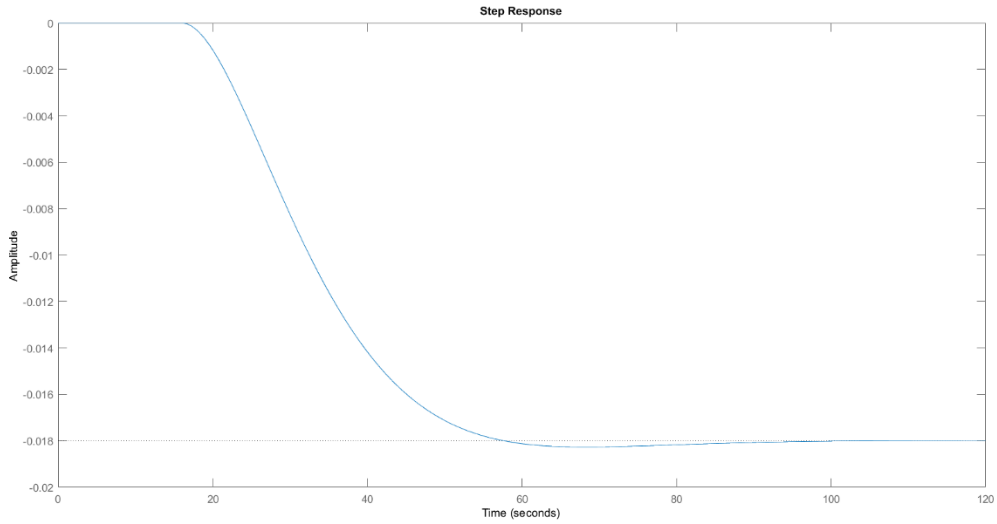


Figure IV.25 : CV3 , MV2 Transfer Function Diagram

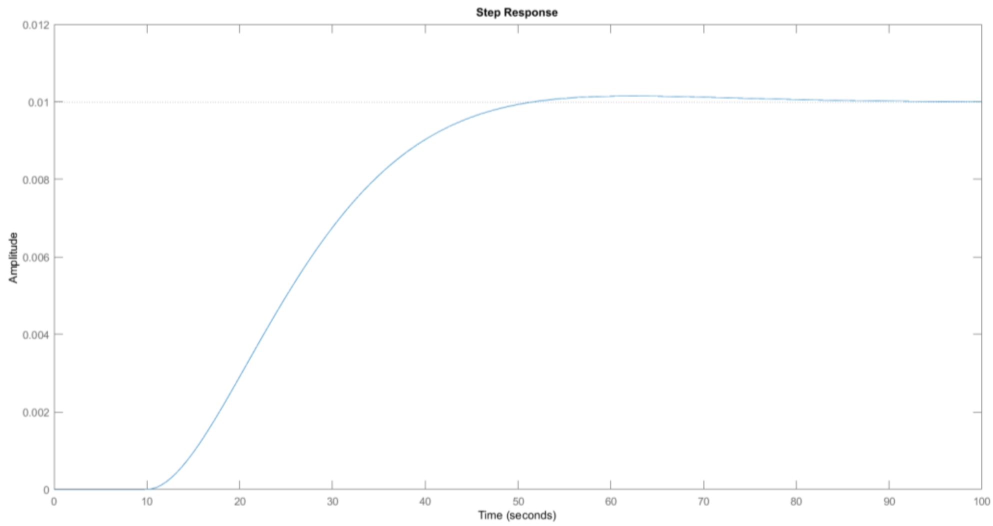


Figure IV.26 : CV4 , MV2 Transfer Function Diagram

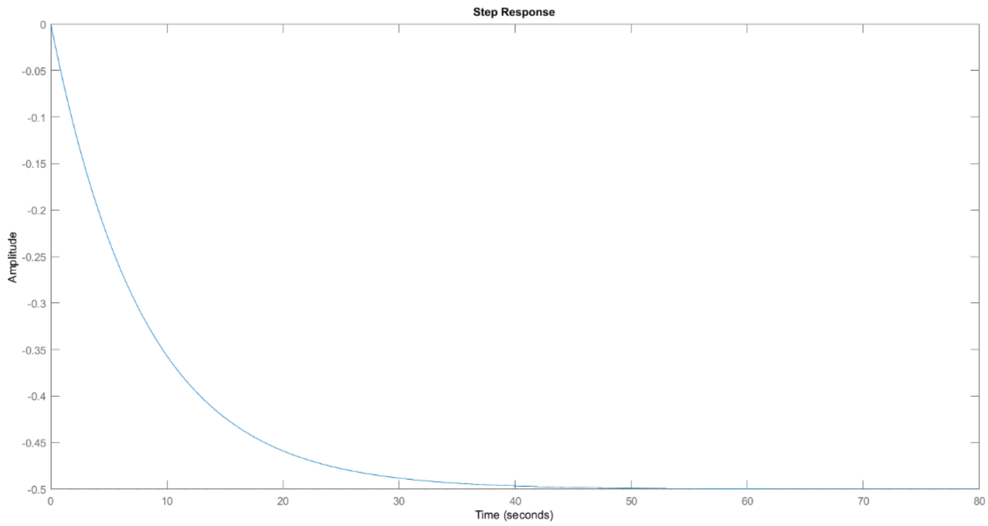


Figure IV.27 : CV5 , MV3 Transfer Function Diagram

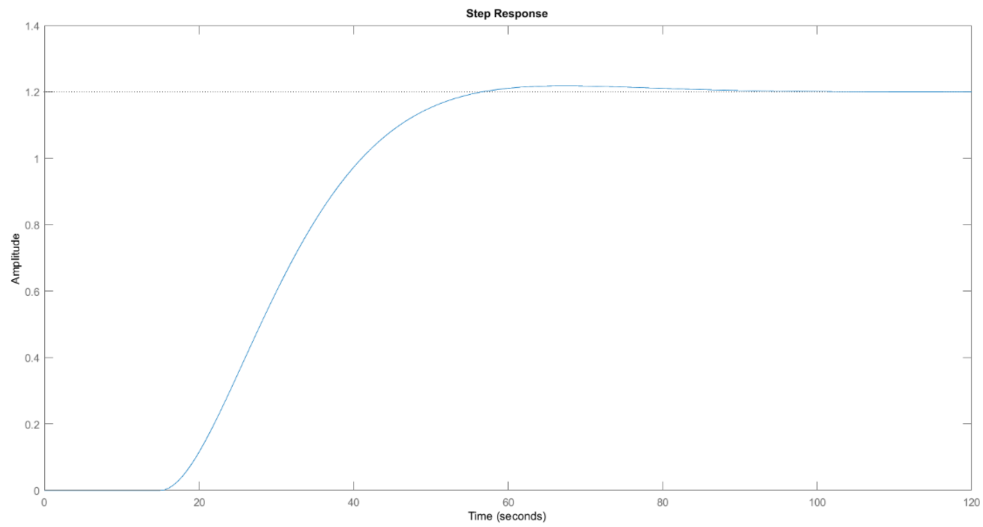


Figure IV.28 : CV6 , MV3 Transfer Function Diagram

IV.4.5.2. Calculating RGA between MV1,MV2 and CV1, CV2

$$G(S) = \begin{bmatrix} -35\frac{8.57S+1}{10.5S^2+6.49S+1}e^{-0S} & 0 \\ 0 & -0.000171\frac{1}{150S^2+22S+1}e^{-60S} \end{bmatrix}$$

- Calculation of the steady-state gain matrix of the model :

$$K_s = \lim_{S\to\infty} G(S)$$

$$K_s = \begin{bmatrix} \lim_{S\to\infty} g_{11}(S) & \lim_{S\to\infty} g_{12}(S) \\ \lim_{S\to\infty} g_{21}(S) & \lim_{S\to\infty} g_{22}(S) \end{bmatrix}$$

After Calculation :

$$K_s = \begin{bmatrix} -35 & 0 \\ 0 & -0.000171 \end{bmatrix}$$

- Calculation of the inverse of the matrix K_s :

$$K_s^{-1} = \begin{bmatrix} -0.0285714285714285714 28 & 0 \\ 0 & -5847.9532163742690058 \end{bmatrix}$$

- Calculating the transpose of the matrix K_s^{-1} :

$$[K_s^{-1}]^T = \begin{bmatrix} -0.028571428571428571428 & 0 \\ 0 & -5847.9532163742690058 \end{bmatrix}$$

- Calculation of the relative gains matrix :

$$RGA = K_s . \times [K_s^{-1}]^T$$

$$RGA = \begin{bmatrix} 1 & 0 \\ 0 & 1 \end{bmatrix}$$

IV.4.5.3. Calculating RGA between MV1,MV2 and CV1, CV3

$$G(S) = \begin{bmatrix} -35\frac{8.57S+1}{10.5S^2+6.49S+1}e^{-0S} & 0 \\ 0 & -0.018\frac{1}{100S^2+16S+1}e^{-16S} \end{bmatrix}$$

- Calculation of the steady-state gain matrix of the model :

$$K_s = \lim_{S\to\infty} G(S)$$

After Calculation :

$$K_s = \begin{bmatrix} -35 & 0 \\ 0 & -0.018 \end{bmatrix}$$

- Calculation of the inverse of the matrix K_s :

$$K_s^{-1} = \begin{bmatrix} -0.0285714285714285714 28 & 0 \\ 0 & -55.555555555555555555 \end{bmatrix}$$

- Calculating the transpose of the matrix K_s^{-1} :

$$[K_s^{-1}]^T = \begin{bmatrix} -0.0285714285714285714 28 & 0 \\ 0 & -55.555555555555555555 \end{bmatrix}$$

- Calculation of the relative gains matrix :

$$RGA = K_s.\times [K_s^{-1}]^T$$

$$RGA = \begin{bmatrix} 1 & 0 \\ 0 & 1 \end{bmatrix}$$

IV.4.5.4. Calculating RGA between MV1,MV2 and CV1, CV4

$$G(S) = \begin{bmatrix} -35\dfrac{8.57S+1}{10.5S^2+6.49S+1}e^{-0S} & 0 \\ 0 & -0.01\dfrac{1}{100S^2+16S+1}e^{-10S} \end{bmatrix}$$

- Calculation of the steady-state gain matrix of the model :

$$K_s = \lim_{S\to\infty} G(S)$$

After Calculation :

$$K_s = \begin{bmatrix} -35 & 0 \\ 0 & -0.01 \end{bmatrix}$$

- Calculation of the inverse of the matrix K_s :

$$K_s^{-1} = \begin{bmatrix} -0.0285714285714285714 28 & 0 \\ 0 & 100 \end{bmatrix}$$

- Calculating the transpose of the matrix K_s^{-1} :

$$[K_s^{-1}]^T = \begin{bmatrix} -0.0285714285714285714 28 & 0 \\ 0 & 100 \end{bmatrix}$$

- Calculation of the relative gains matrix :

$$RGA = K_s.\times [K_s^{-1}]^T$$

$$RGA = \begin{bmatrix} 1 & 0 \\ 0 & 1 \end{bmatrix}$$

IV.4.5.5. Calculating RGA between MV1,MV3 and CV1, CV5

$$G(S) = \begin{bmatrix} -35\frac{8.57S+1}{10.5S^2+6.49S+1}e^{-0S} & 0 \\ 0 & -0.5\frac{1}{8S+1}e^{-0S} \end{bmatrix}$$

- Calculation of the steady-state gain matrix of the model :

$$K_s = \lim_{S\to\infty} G(S)$$

After Calculation :

$$K_s = \begin{bmatrix} -35 & 0 \\ 0 & -0.5 \end{bmatrix}$$

- Calculation of the inverse of the matrix K_s :

$$K_s^{-1} = \begin{bmatrix} -0.0285714285714285714 28 & 0 \\ 0 & -2 \end{bmatrix}$$

- Calculating the transpose of the matrix K_s^{-1} :

$$[K_s^{-1}]^T = \begin{bmatrix} -0.028571428571428571428 & 0 \\ 0 & -2 \end{bmatrix}$$

- Calculation of the relative gains matrix :

$$RGA = K_s.\times [K_s^{-1}]^T$$

$$RGA = \begin{bmatrix} 1 & 0 \\ 0 & 1 \end{bmatrix}$$

IV.4.5.6. Calculating RGA between MV1,MV3 and CV1, CV6

$$G(S) = \begin{bmatrix} -35\dfrac{8.57S+1}{10.5S^2+6.49S+1}e^{-0S} & 0 \\ 0 & 1.2\dfrac{1}{100S^2+16S+1}e^{-15S} \end{bmatrix}$$

- Calculation of the steady-state gain matrix of the model :

$$K_s = \lim_{S\to\infty} G(S)$$

After Calculation :

$$K_s = \begin{bmatrix} -35 & 0 \\ 0 & 1.2 \end{bmatrix}$$

- Calculation of the inverse of the matrix K_s :

$$K_s^{-1} = \begin{bmatrix} -0.0285714285714285714 28 & 0 \\ 0 & 0.83333333333333333333 \end{bmatrix}$$

- Calculating the transpose of the matrix K_s^{-1} :

$$[K_s^{-1}]^T = \begin{bmatrix} -0.028571428571428571428 & 0 \\ 0 & 0.83333333333333333333 \end{bmatrix}$$

- Calculation of the relative gains matrix :

$$RGA = K_s.\times [K_s^{-1}]^T$$

$$RGA = \begin{bmatrix} 1 & 0 \\ 0 & 1 \end{bmatrix}$$

IV.4.5.7. Calculating RGA between MV2,MV3 and CV2, CV5

$$G(S) = \begin{bmatrix} -0.000171 \dfrac{1}{150S^2 + 22S + 1} e^{-60S} & 0 \\ 0 & -0.5 \dfrac{1}{8S+1} e^{-0S} \end{bmatrix}$$

- Calculation of the steady-state gain matrix of the model :

$$K_s = \lim_{S \to \infty} G(S)$$

After Calculation :

$$K_s = \begin{bmatrix} -0.000171 & 0 \\ 0 & -0.5 \end{bmatrix}$$

- Calculation of the inverse of the matrix K_s :

$$K_s^{-1} = \begin{bmatrix} -5847.95 & 0 \\ 0 & -2 \end{bmatrix}$$

- Calculating the transpose of the matrix K_s^{-1} :

$$[K_s^{-1}]^T = \begin{bmatrix} -5847.95 & 0 \\ 0 & -2 \end{bmatrix}$$

- Calculation of the relative gains matrix :

$$RGA = K_s . \times [K_s^{-1}]^T$$

$$RGA = \begin{bmatrix} 1 & 0 \\ 0 & 1 \end{bmatrix}$$

IV.4.5.8. Calculating RGA between MV2,MV3 and CV2, CV6

$$G(S) = \begin{bmatrix} -0.000171\frac{1}{150S^2+22S+1}e^{-60S} & 0 \\ 0 & 1.2\frac{1}{100S^2+16S+1}e^{-15S} \end{bmatrix}$$

- Calculation of the steady-state gain matrix of the model :

$$K_s = \lim_{S\to\infty} G(S)$$

After Calculation :

$$K_s = \begin{bmatrix} -0.000171 & 0 \\ 0 & 1.2 \end{bmatrix}$$

- Calculation of the inverse of the matrix K_s :

$$K_s^{-1} = \begin{bmatrix} -5847.95 & 0 \\ 0 & -2 \end{bmatrix}$$

- Calculating the transpose of the matrix K_s^{-1} :

$$[K_s^{-1}]^T = \begin{bmatrix} -5847.95 & 0 \\ 0 & -2 \end{bmatrix}$$

- Calculation of the relative gains matrix :

$$RGA = K_s.\times [K_s^{-1}]^T$$

$$RGA = \begin{bmatrix} 1 & 0 \\ 0 & 1 \end{bmatrix}$$

IV.4.5.9. Calculating RGA between MV2,MV3 and CV3, CV5

$$G(S) = \begin{bmatrix} -0.018\frac{1}{100S^2+16S+1}e^{-16S} & 0 \\ 0 & -0.5\frac{1}{8S+1}e^{-0S} \end{bmatrix}$$

- Calculation of the steady-state gain matrix of the model :

$$K_s = \lim_{s\to\infty} G(S)$$

After Calculation :

$$K_s = \begin{bmatrix} -0.018 & 0 \\ 0 & -0.05 \end{bmatrix}$$

- Calculation of the inverse of the matrix K_s :

$$K_s^{-1} = \begin{bmatrix} -55.5556 & 0 \\ 0 & -2 \end{bmatrix}$$

- Calculating the transpose of the matrix K_s^{-1} :

$$[K_s^{-1}]^T = \begin{bmatrix} -55.5556 & 0 \\ 0 & -2 \end{bmatrix}$$

- Calculation of the relative gains matrix :

$$RGA = K_s.\times [K_s^{-1}]^T$$

$$RGA = \begin{bmatrix} 1 & 0 \\ 0 & 1 \end{bmatrix}$$

IV.4.5.10. Calculating RGA between MV2,MV3 and CV3, CV6

$$G(S) = \begin{bmatrix} -0.018\dfrac{1}{100S^2+16S+1}e^{-16S} & 0 \\ 0 & 1.2\dfrac{1}{100S^2+16S+1}e^{-15S} \end{bmatrix}$$

- Calculation of the steady-state gain matrix of the model :

$$K_s = \lim_{S\to\infty} G(S)$$

After Calculation :

$$K_s = \begin{bmatrix} -0.018 & 0 \\ 0 & 1.2 \end{bmatrix}$$

- Calculation of the inverse of the matrix K_s :

$$K_s^{-1} = \begin{bmatrix} -55.5556 & 0 \\ 0 & 0.833333 \end{bmatrix}$$

- Calculating the transpose of the matrix K_s^{-1} :

$$[K_s^{-1}]^T = \begin{bmatrix} -55.5556 & 0 \\ 0 & 0.833333 \end{bmatrix}$$

- Calculation of the relative gains matrix :

$$RGA = K_s.\times [K_s^{-1}]^T$$

$$RGA = \begin{bmatrix} 1 & 0 \\ 0 & 1 \end{bmatrix}$$

IV.4.5.11. Calculating RGA between MV2,MV3 and CV4, CV5

$$G(S) = \begin{bmatrix} -0.01\frac{1}{100S^2 + 16S + 1}e^{-10S} & 0 \\ 0 & -0.5\frac{1}{8S + 1}e^{-0S} \end{bmatrix}$$

- Calculation of the steady-state gain matrix of the model :

$$K_s = \lim_{S\to\infty} G(S)$$

After Calculation :

$$K_s = \begin{bmatrix} -0.01 & 0 \\ 0 & -0.5 \end{bmatrix}$$

- Calculation of the inverse of the matrix K_s :

$$K_s^{-1} = \begin{bmatrix} 100 & 0 \\ 0 & -2 \end{bmatrix}$$

- Calculating the transpose of the matrix K_s^{-1} :

$$[K_s^{-1}]^T = \begin{bmatrix} 100 & 0 \\ 0 & -2 \end{bmatrix}$$

- Calculation of the relative gains matrix :

$$RGA = K_s.\times [K_s^{-1}]^T$$

$$RGA = \begin{bmatrix} 1 & 0 \\ 0 & 1 \end{bmatrix}$$

IV.4.5.12. Calculating RGA between MV2,MV3 and CV4, CV6

$$G(S) = \begin{bmatrix} -0.01\frac{1}{100S^2+16S+1}e^{-10S} & 0 \\ 0 & 1.2\frac{1}{100S^2+16S+1}e^{-15S} \end{bmatrix}$$

- Calculation of the steady-state gain matrix of the model :

$$K_s = \lim_{S\to\infty} G(S)$$

After Calculation :

$$K_s = \begin{bmatrix} -0.01 & 0 \\ 0 & 1.2 \end{bmatrix}$$

- Calculation of the inverse of the matrix K_s :

$$K_s^{-1} = \begin{bmatrix} 100 & 0 \\ 0 & 0.833333 \end{bmatrix}$$

- Calculating the transpose of the matrix K_s^{-1} :

$$[K_s^{-1}]^T = \begin{bmatrix} 100 & 0 \\ 0 & 0.833333 \end{bmatrix}$$

- Calculation of the relative gains matrix :

$$RGA = K_s.\times [K_s^{-1}]^T$$

$$RGA = \begin{bmatrix} 1 & 0 \\ 0 & 1 \end{bmatrix}$$

IV.4.6. Intrepretation and comments :

The Relative Gain Array (RGA) results for the depropanizer column demonstrate how Advanced Process Control (APC) effectively manages interactions between manipulated variables (MVs) and controlled variables (CVs) to maintain stability, even in the presence of perturbations. The consistent RGA matrices, which show strong, independent interactions between specific MV-CV pairs without cross-interaction, illustrate APC's capability to handle complex variable relationships. For instance, MV1 interacts exclusively with CV1, and MV2 with CV2, ensuring precise control without interference. This means that each MV only affects the CV it interacts with, and no other CV is influenced by that MV. Other pairs, such as MV1 with CV1 and MV3 with CV5 or MV2 with CV2 and MV3 with CV6, also exhibit distinct, effective control. This independence simplifies control system design, enhancing the stability and efficiency of the depropanizer column's operation.

This robust control strategy is a key aspect of the technology used in Real-time Multivariable Predictive Control Technology (RMPCT) that APC employs to control the distillation column. RMPCT leverages these insights to optimize process parameters, maintaining optimal performance even with hard disturbances. By effectively handling the interactions between CVs and MVs, APC ensures that the system remains stable and efficient, underscoring its value in modern industrial operations. This technology highlights APC's role in achieving precise, efficient, and stable process control, making it an essential tool for managing complex distillation processes.

IV.5. Economic Impact

IV.5.1. Economic Impact of Advanced Process Control (APC) on Propane Production

IV.5.1.1. Production Data

Table IV.4 : Production Data Without/ With APC

Parameter	Without APC (March 1-15, 2021)	With APC (October 6-12, 2023)
Total Production (TM)	3,131	1,494.105
Number of Days	15	7
Average Daily Production (TM/day)	208.73	213.44
TP/Design (%) Average	33.18	33.93

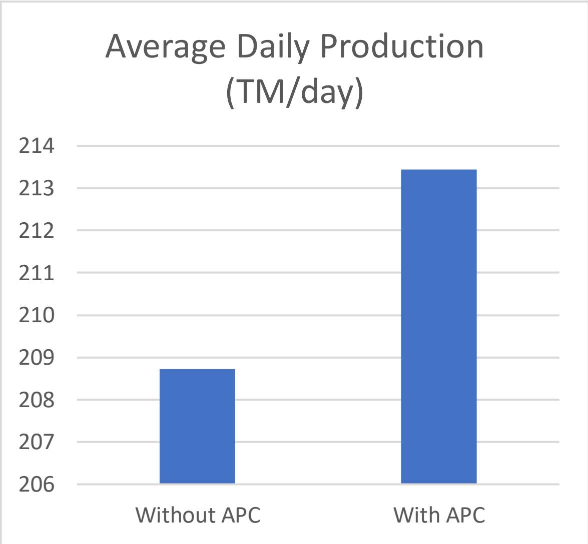


Figure IV.29 : Average daily production

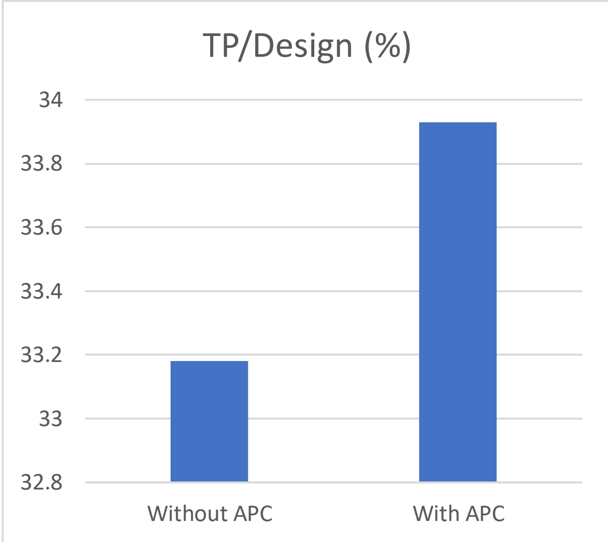


Figure IV.30 : TP/Design

IV.5.1.2. Impact of APC Implementation on Propane Production and Revenue

The implementation of APC has led to a significant increase in propane production. The average daily production increased by 4.71 TM, and if APC had been implemented over 15 days, the projected total production would be 3,201.6 TM. This represents an increase of 70.6 TM over the period. Assuming a market price of $500 per TM of propane, the additional production would result in an extra $35,300 in revenue.

Table IV.5 : Projected Production and Revenue Increase with APC Implementation

Parameter	Value
Increase in Average Daily Production (TM/day)	4.71
Projected Total Production With APC Over 15 Days (TM)	3,201.6
Increase in Production Over 15 Days (TM)	70.6
Market Price of Propane ($/TM)	500
Additional Revenue Over 15 Days ($)	35,300
Improvement in TP/Design (%)	0.75

IV.5.1.3. Economic Impact of Advanced Process Control (APC) on the Hot Oil Flow of the Depropanizer

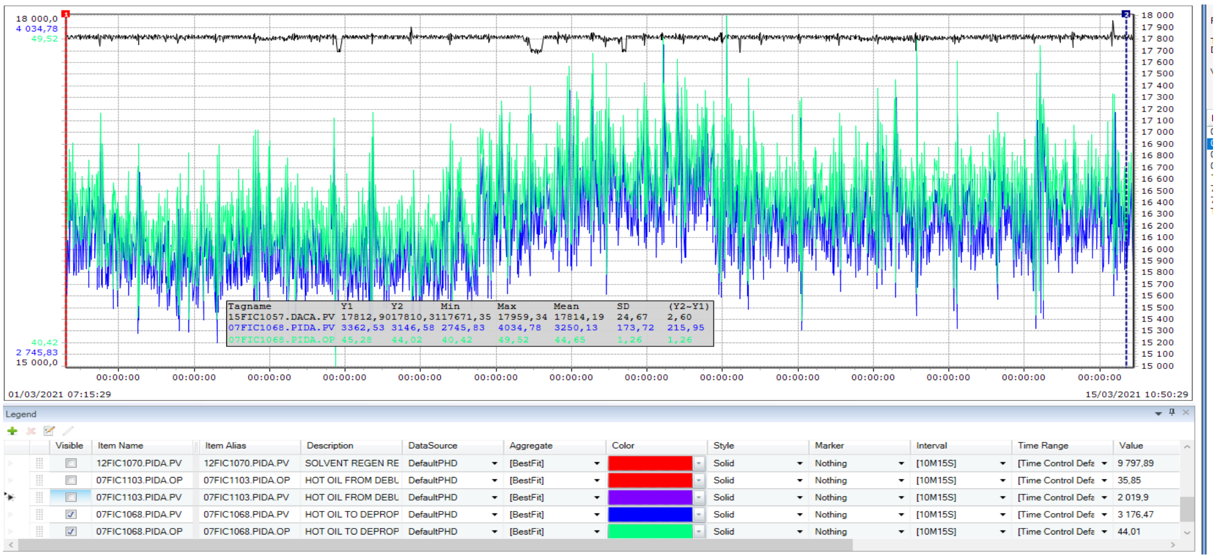


Figure IV.31 : Oil Flow Rate Trend of the 07MC04 Reboiler in the Depropanizer Column Before APC Installation

Figure IV.32 : Oil Flow Rate Trend of the 07MC04 Reboiler in the Depropanizer Column After APC Installation

IV.5.1.4. Stability and Efficiency

The introduction of APC in the depropanizer column has resulted in a more stable and efficient operation. The flow variability has been significantly reduced, leading to several economic benefits, including lower energy consumption, reduced mechanical stress on equipment, and improved product quality.

Table IV.6 : Comparison of Operational Parameters in the Depropanizer Column Before and After APC Implementation

Paramerter	Without APC(green Graph)	With APC (Pink Graph)
Flow Variability	High	Low
Energy Consumption	Increased	Reduced
Mechanical Stress	High	Low
Product Quality Variability	High	Low

IV.5.1.5. Quantifying Economic Benefits

1- Energy Savings :

- Reduction in energy consumption from 100 units/hour to 90 units/hour.
- Annual energy savings amount to 87,600 units.
- With an energy cost of $0.10 per unit, the annual cost savings are $8,760.

2- Maintenance Savings :

- With APC, mechanical stress is reduced, lowering maintenance costs.
- If annual maintenance costs without APC are $50,000, a 20% reduction results in savings of $10,000 per year.

3- Increased Product Quality :

- Improved product quality can lead to a 5% increase in revenue.
- If annual revenue without APC is $1,000,000, the additional revenue from improved quality is $50,000.

Table IV.7 : Cost and Benefit Analysis of APC Implementation in the Depropanizer Column

Parameter	Without APC	With APC	Saving/benifits
Energy Consumption (units/hour)	100	90	10 units/hour
Annual Energy Consumption (units/year)	876,000	788,400	87,600 units/year
Cost of Energy ($/unit)	$0.10	$0.10	-
Annual Energy Cost ($/year)	$87,600	$78,840	$8,760
Annual Maintenance Cost ($/year)	$50,000	-	$10,000 (20% reduction)
Annual Revenue ($/year)	$1,000,000	-	$50,000 (5% increase)

IV.5.1.6. Total Economic Impact

Table IV.8 : Annual Economic Benefits of APC Implementation

Aspect	Value
Energy Savings	$8,760
Maintenance Savings	$10,000
Increased Revenue	$50,000
Total Annual Economic Benefit	$68,760

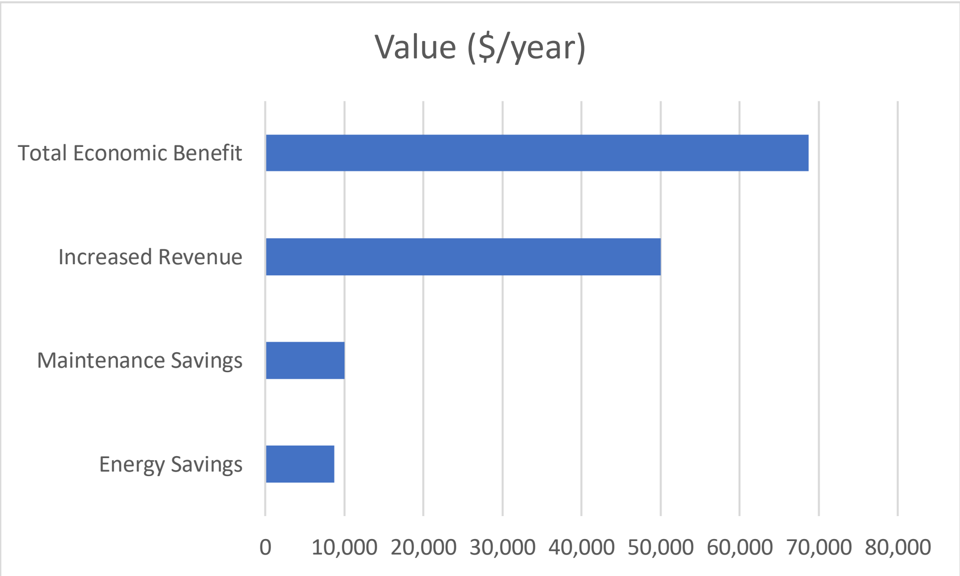


Figure IV.33 : Economic Benefits of Implementing Advanced Process Control (APC) in the Depropanizer Column

IV.5.1.7. Summary of Economic Benefits

Table IV.9 : Annual Economic Benefits of Implementing Advanced Process Control (APC)

Benifit	Value ($/year)
Energy Savings	$8,760
Maintenance Cost Reduction	$10,000
Additional Revenue from Improved Product Quality	$50,000
Total Economic Impact	$68,760

The implementation of Advanced Process Control (APC) at the Sonatrach GLIK complex has had a significant economic impact on both propane production and the flow of the depropanizer column. By stabilizing flow rates and improving efficiency, APC has increased the average daily production of propane by 4.71 TM, resulting in an additional 70.6 TM over a 15-day period, translating to $35,300 in additional revenue. In the depropanizer column, APC has reduced energy consumption, saving $8,760 annually, and lowered maintenance costs by $10,000 per year. Furthermore, the improved product quality has led to an additional $50,000 in annual revenue. Overall, the total economic benefit of implementing APC is estimated at $68,760 per year, demonstrating the substantial financial and operational advantages of advanced process control technologies.

IV.6. Conclusion :

This chapter provided a practical, mathematical, and economical study of controlling a distillation column. We focused on using PID controllers and the BLT method for tuning. Despite improvements with BLT-tuned PID controllers, high interaction levels and sensitivity to disturbances remain challenging, showing that PID control alone is often insufficient.

Our interaction study revealed that each manipulated variable (MV) only affects its corresponding control variable (CV). The economic study comparing the depropanizer column's performance with and without APC demonstrated significant benefits. APC stabilizes the system, reduces energy consumption, lowers maintenance costs, and improves product quality.

Advanced strategies like APC are essential for managing complex dynamics and disturbances in distillation columns, proving their superiority in achieving stable and efficient operations.

General Conclusion

General Conclusion :

In this work, we conducted a comprehensive study on the control and optimization of a distillation column, focusing on practical, mathematical, and economical aspects. We began with the implementation and tuning of PID controllers using the BLT method, which provided a foundational understanding of managing system interactions and enhancing stability. Despite the improvements, the high interaction levels and sensitivity to disturbances highlighted the limitations of PID control alone.

We then explored Advanced Process Control (APC), specifically the use of Relative Gain Array (RGA) analysis, to handle the complex interactions between control variables (CVs) and manipulated variables (MVs). Our findings showed that each MV affects only its corresponding CV, validating the effectiveness of APC in maintaining stable operations even with significant perturbations.

The economic study comparing the depropanizer column's performance with and without APC demonstrated substantial benefits. With APC, the system achieved better stability, reduced energy consumption, lower maintenance costs, and improved product quality, underscoring the economic advantages of advanced control strategies.

This memory highlights the importance of integrating advanced control methods like APC to manage the intricate dynamics of distillation columns. By combining practical implementations, mathematical analyses, and economic evaluations, we showcased a holistic approach to optimizing distillation processes, proving the superiority of APC in achieving efficient and stable operations.

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