Sudan University of Science and Technology

College of Petroleum Engineering and Technology

Department of Transportation and Refining Engineering

Simulation and analysis of cryogenic air separation process (case study Khartoum refinery company)

محاكاة وتحليل وحدة فصل الهواء بالتبريد العالي (دراسة حالة:شركة مصفاة الخرطوم)

Submitted in partial fulfillment of the requirement of for the (B.Sc) Degree in transportation and refining Engineering

Prepared by:

Adam Ahmed Abdalla Mustafa

Alhassan Abdalrhman Hassan Abdalla

Ehab Mustafa Abdalgader Mohammed

Supervised by:

Mrs: Zeinab Abdalla Mohamed Khalel

September 2014

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The project is accepted by college of petroleum engineering and technology to transportation and refining department.

Project Supervisor:- Mrs/ Zienab Abdallah Mohammed Khliel

Signature:-...

Head of Department:Mrs/ Zienab Abdallah Mohammed Khliel

Signature:...

Dean of College:Dr/ Somya Abdualmonem Mohamed

Signature:...

Date: / /2014

الإستهلال

قال تعالى:–

بِسْمِ اللَّهِ الرَّحْمَنِ الرَّحِيمِ
(اقْرَأْ بِاسْمِ رَبِّكَ الَّذِي خَلَقَ (١) خَلَقَ الْإِنْسَانَ مِنْ عَلَقٍ
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(العلق ١-٥)

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Abstract

Cryogenic air separation is a process by which highly purified gases or liquids are produced. This is achieved by taking large volumes of air from the atmosphere, which is then compressed, cooled, and liquefied. Through a process of distillation the air is then separated into its major components. The applications of air component in refinery are FCC Enrichment and NOx Reduction, Sulfur Recovery and Acid Gas Plants, Streamlining, Pressure and Leak Testing, Purging and Blanketing, Process Chilling and Process Equipment Cooling. Process simulations enable us to find stable operation point with optimal condition for the process examined. HYSYS programme is used to simulate the process to examine the effect of change reflux ratio and tray number on the performance and then analyze the results obtained.

Key words: cryogenic air separation, simulation, material and energy balance, rectification, design, heat exchanger, cost estimation, process analysis, cold box, distillation column

التجريد

عملية فصل الهواء بالتبريد العالي يتم من خلالها إنتاج سوائل أو غازات بنقاوة عالية.العملية تتم بسحب كمية كبيرة من الهواء الجوي والذي يضغط ،يبرد، يسيل خلال عملية تقطير الهواء ثم يفصل الي مكوناته الرئيسية. تطبيقات مكونات الهواء في المصفاة تشمل التكسيير بالتمييع الحفزي وتقليل المكونات النتروجينية وإستخلاص الكبريت وفي عمليات الغازات الحمضية، وفي إختبار الضغط والتسرب وكالعزل والتغطية ،وفي عمليات التبريد وتبريد معدات العمليات .محاكاة العملية تساعد في إيجاد الظروف التشغيلية المستقرة والمثلي للعملية.برنامج المحاكاة هايسس أستخدم لمعرفة أثر تغيير عدد الصواني ونسبة الراجع علي أداء الوحدة وتحليل النتائج التي تم الحصول عليها.

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Chapter 1

Introduction and literature review

1. Introduction and literature review

1.1 introduction:

An air separation plant separates atmospheric air into its primary components, typically nitrogen and oxygen sometimes also argon and rarely other inert gases.

World first air separation plant was in 1902 for recovery oxygen after that it developed to recover oxygen and nitrogen, in 1910 was the first air separation plant using the double column rectification process. After that in 1990 the first tele-controlled air separation plant with unmanned operation and produce pure argon by rectification.

The main objective of the project:

Simulate the process, analyze the result of changing parameters, and evaluate the process performance.

1.2. Air separation technologies

There are various technologies that are used for the separation process:-

1.2.1. Cryogenic air separation:

Cryogenic air separation is a process by which highly purified gases or liquids are produced. This is achieved by taking large volumes of air from the atmosphere, which is then compressed, cooled, and liquefied. Through a process of distillation the air is then separated into its major components. After the air is compressed, impurities must be removed. Depending on user requirements, different variations can be used in cryogenic air separation cycles to produce industrial gas products. (chemsystems, 2012)

1.2.2. Pressure swing adsorption or PSA:

Rely on the fact that under pressure gases tend to be attracted to solid surfaces or "adsorbed". The higher the pressure, the more gas is adsorbed; when the pressure is reduced, the gas is released, or desorbed. PSA processes can be used to separate gases in a mixture because different gases tend to be attracted to different solid surfaces more or less strongly .(chemsystem,2012).

1.2.3. Vacuum swing adsorption (VSA):

Differs from pressure swing adsorption (PSA) techniques due to the fact that it operates at near-ambient temperatures and pressures. VSA may actually be best described as a subset of the larger category of PSA. It differs primarily from PSA in that PSA typically vents to atmospheric pressures, and uses a pressurized gas feed into the separation process.(chemsystem,2012).

1.2.4. Membrane:-

Membrane air separation is based on the principle that different gases have different permeation rates through the polymer film. Oxygen (plus water vapour and carbon dioxide) is considered a "fast gas" which diffuses more rapidly through the tube walls than the "slow gases," argon and nitrogen. This “fast gas/slow gas” diffusion allows dry air to be converted into a product that is an inert mix of mostly nitrogen gas and argon, and a low-pressure "permeate" or waste gas that is enriched in oxygen (plus water vapour and carbon dioxide) and vented from the shell(chemsystem,2012).

Table 1.1 Compression between the cryogenic separation and non- cryogenic ((Carnegie,1998)

Process	Cryogenic separation	Non-cryogenic
Advantages	Low amount of electricity per unit nitrogen	Cost-effective nitrogen and quick installation and start-up
	Produce very high purity nitrogen	Production of relatively high purity
	Can generate liquid nitrogen	Low to moderate capital cost
Disadvantages	Large side space and utility requirements	High maintains equipment
	High capital cost	Noisy operation
	Limited scalability in production	Limited scalability
	Long start-up and shut-down	Short start up and shut down

1.3 Gases and their application in refinery:

1.3.1 Oxygen:-

- FCC Enrichment and NOx Reduction:

Oxygen can be added to the regenerator of a fluid catalytic cracker (FCC) to improve yields, assist with coke burning and heat balance, and to reduce cyclone Velocities. This can also reduce NOx and CO emissions.

Additionally, oxygen is used to produce ozone, which is required in the NOx removal equipment incorporated into many FCC units (Matheson, 2011).

-Sulfur Recovery and Acid Gas Plants:-

The use of Oxygen in a sulfur recovery unit (SRU) can increase plant capacity and help with contaminants and low hydrogen sulfide streams (Matheson, 2011).

Oxygen can also be used to boost temperature. Oxygen use can be intermittent or continuous.

- Streamlining:

Oxygen can be used to enrich the air in oxidation processes to increase throughput and improve productivity (Matheson, 2011).

1.3.2 Nitrogen:-

- Pressure and Leak Testing:

An inert gas such as Nitrogen can be used to pressurize new, repaired, or modified tanks, pipelines, vessels, and process equipment in order to check their integrity and leak tightness.

Dry, high purity nitrogen will not contaminate the system with moisture lubricants or oxygen (whereas pressure testing with compressed air or water can easily lead to system contamination) (Matheson, 2011).

- Pipe Freezing:-

Liquid Nitrogen can be used to freeze a section of a pipeline’s contents. The frozen section, or plug, permits work such as valve repairs, the addition of tees or tap-offs, or the extension of existing pipelines without draining of the entire system (Matheson,2011).

Only liquids which freeze at the liquid gas temperature, such as water, glycol, and some oils can be handled by the process (Matheson,2011).

- Pressure Transfer: -

Nitrogen can be used to transfer liquid or powder products to and from tanker trucks, or storage vessels without requiring pumps, mechanical compressors or external power sources.

Dry inert nitrogen is ideal for transferring toxic fluids, highly viscous fluids, ethical pharmaceuticals, flammable products, and materials which become corrosive when contacted with moisture (Matheson, 2011).

- Purging and Blanketing:-

Inert gas, typically Nitrogen, can be used to displace air, flammable vapors, and contaminants from storage tanks, pipelines, process equipment, railcars, ocean-bound containers, or other vessels.

Gas can be used to maintain an inert atmosphere above a liquid or powdered product inside a storage tank, silo, reactor, process equipment, or other container.

The gas will help prevent product degradation from moisture and oxygen, control volatile emissions, and safeguard against fires and explosions (Matheson, 2011).

- Process Chilling and Process Equipment Cooling: -

_Liquid Nitrogen can be used as a refrigerant for temperature control in a variety of processes, including solvent recovery and chemical processes.

Liquid gases can achieve a lower temperature and provide a faster cool down rate than mechanical refrigeration systems(Matheson,2011).

1.4 Literature review:

Cryogenic air separation was pioneered by Dr. Carl von Linde in the early 20th century and is still used today to produce high purity gases.

After that many modifications was precede done for process to increase the profits or reduce cost of operation.

Also In1981 lees gaumar modified the system to use a single pressure distillation column and burns a nitrogen-oxygen waste stream to provide power for the air compressor(mechanical resarch,2012).

In 1988 Joseph M abrardo modified process by addition of a small second column which purifies a portion of the oxygen enriched liquid from the nitrogen generator overhead condenser. Reboiling for the second column is provided by condensing part of the nitrogen overhead from the nitrogen generator. This condensed nitrogen is used as reflux in the nitrogen generator (mechanical resarch,2012).

In 2010 Lewis yan and etal discovered that the equipment deficiency is overcome in the new process. Also the nitrogen expansion and the structure packing column are added while the pinch analysis is done for the new process it resulted the energy consumption can be 7.55MW lower than the original process. The total Energy efficiency can be raised by 27.21% (mechanical resarch,2012).

Hua Zhou in 2012 proposed a novel single-column atmospheric cryogenic air separation process to reduce energy consumption through the implementation of thermal pump technique to verify the validity of the proposed separation process for typical configurations of the single-column processes are constructed and simulated on the aspen plus (mechanical resarch,2012).

They arrived to result that the process can save energy up to 23% and produce the products purity in industrial standard (kolbjoern, 2010).

In Norwegian university at 2011An integrated gasification combined cycle (IGCC) is one of the most promising alternatives (mechanical resarch,2012).

It is the mission of a European collaboration project called DECARB it to enable the commercial use of this type of power plant. One of the main process units of an IGCC is the air separation unit (ASU). It provides both oxygen and nitrogen to the gasifies, and nitrogen to the gas turbine. The main objective of this thesis is to improve the Second law efficiency of a cryogenic ASU, with a focus on the use of novel distillation.

At 2012 in college of engineering in India some student make Simulation of medium purity oxygen cryogenic air separation plant integrated with biomass gasifies is carried out by using Aspen plus. Such cryogenic air separation plants which produce oxygen in a range between 85-98% can be used economically for gasification (mechanical resarch,2012).

In 2013 Zeinab A. M. Khalel etal developed a new cryogenic air separation process by using flash separator wish used to replace the turbine required to recover a portion of the energy in the double-column air separation process.

The new process decreases the energy consumption and increase the productivity.

The table 1.2 summary of modified applied to cryogenic unit.

Author name	Year	Modification proposed	Result of the modification
Lees gaumar	1981	Use a single pressure distillation column	provide power for the air compressor
Joseph M abrardo	1988	Add small second column	Increase oxygen purity
Lewis yan and etal	2010	Add nitrogen expansion	raise the total energy
Hua Zhou	2012	Add thermal pump	reduce energy consumption and increase purity
Zeinab Abdulla	2013	replace the turbine with flash separator	decreases the energy consumption and increase the productivity

Chapter 2

Process description and simulation

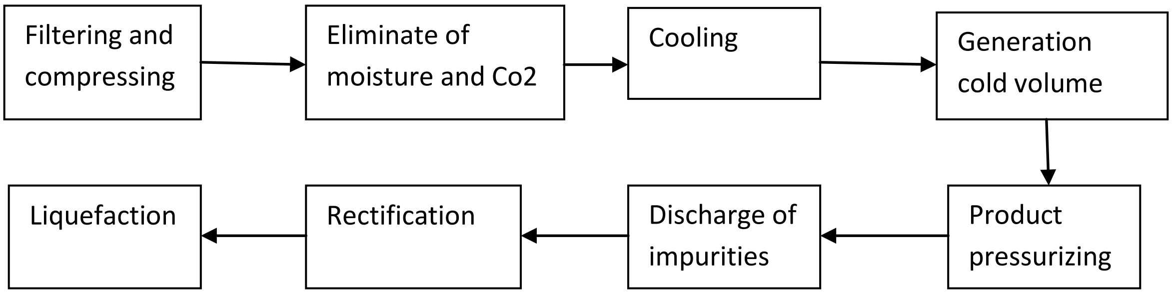
2.Process description and simulation

Basic principle of air separation:

The basic principle of air separation is to use the difference of boiling points between different major ingredients in the liquefied air, such as oxygen, nitrogen, etc, and leverage the rectification technique to separate individual ingredients apart (KRC,2011)

2.1 Operation process of air separation system:

Productions of the air separation facility primarily include the following eight processes.



2.1.1 Filtering and compression of air:

The air in the atmospheric first of all goes through the air filter to eliminate its mechanical impurities such as dust, and then is compressed by air compressor to the pressure needed. The heat generated during compression is taken away by the cooling water.(KRC,2011).

2.1.2 Elimination of moisture and CO2 from air:

If the moisture and CO2 in the air begin processed goes into the low temperature zone of the air separation equipment, then they can form water ice and carbon ice, which may block the channels in the heat exchanger and the tiny orifice in the in the column plate.

Therefore, the air precooler set and molecular sieve absorber are equipped to pre-eliminate the moisture and CO2 from the air. The molecular sieve absorber are equipped in pairs and used alternatively, with one operate while the other regenerating.(KRC,2011)

2.1.3 Cooling the air to liquefying temperature:

Cooling of air is done inside the main heat exchanger, where the air is cooled down to the liquefied temperature by the back flow gas. At the same time, the low temperature back flow gas is re-heated.(KRC,2011).

2.1.4 Production and generation of cold volume:

Due to the losses in heat insulation, the loss in insufficient reheating in the heat exchanger, and the cold loss generate during the process when the cold box directly releases low temperature liquid to the outside, the distillation tower need to have nonstop supply of could volume.

The cold volume need by the distillation tower will be obtain through the expansion and isothermal throttling effect of oxygen-enriched air (from the dirty nitrogen in the tower) in the expander.(KRC,2011).

2.1.5 Liquefaction:

During the startup stage of the equipment, the air being processed is liquefied in the main heat exchanger and the sub cooler when it exchange heat with the back flowing cold air stream; while during the normal operation process, the process of heat exchanger between oxygen-enriched liquefied air will take place inside the condensing evaporator; because the pressure of two types of fluids are different, so that the nitrogen will be liquefied while the oxygen-enriched liquefied air will be evaporated, and this is prerequisite condition that must be available in order to ensure smooth on-going of the rectification process.(KRC,2011).

2.1.6 Rectification:

Table 2.1 The main component in the air and their physical properties.(KRC,2011)

Name	Chemical symbol	Volume percent (%)	Weight percent (%)
Nitrogen	N_2	78.09	75.5
Oxygen	O_2	20.95	23.1
Argon	Ar	0.0932	1.29
Carbon dioxide	CO_2	0.03	0.05
Helium	He	0.00046	0.00006
Neon	Ne	0.0016	0.0011
Krypton	Kr	0.00011	0.00032
Xenon	Xe	0.00008	0.00004
Hydrogen	H_2	0.0001	0.0005848

Table 2.2 physical properties of main Components in Air (KRC,2011)

Name	Chemical symbol	Boiling Point C	Fusion temperature C	Specific weight		Critical point	
				Kg/m3	Kg/L	C	10^-1 Map
Nitrogen	N2	-195.8 C	-209.86 C	1.25	0.81	-147	34.5
Oxygen	O2	-183 C	-218.4 C	1.43	1.14	-119	51.3
Argon	Ar	-185.7 C	-189.2 C	1.782	1.4	-122	49.59
Helium	He	-268.9 C	-272.55 C	0.18	0.15	-267.7	2.335
Neon	Ne	-246.1 C	-248.6 C	0.748	1.204	-228.7	28.13
Krypton	Kr	-153.2 C	-157.2 C	1.735	2.155	-63.7	56
Xenon	Xe	-108 C	-111.8 C	1.6640.81	3.25	16.6	60.1

99.04% of air is oxygen and nitrogen, with 0.932% being argon, and they basically stay unchanged.

The percentage of hydrogen, carbon dioxide, and hydrocarbon are subject to change in a certain range depending on the geographical area and environment.

The content of vapor in the air will change significantly as affected by the saturation temperature and geographical and environmental condition.

Vapor and carbon dioxide have much different properties from air; under the atmospheric pressure, when the temperature of vapor reach 0 C and the temperature of carbon dioxide reach -79 C, they will transfer to water ice and carbon ice, respectively, thereby blocking of the channel plate heat exchanger and the tiny orifice on the sieve play or trays. Therefore, these component must be eliminated before enter the cold box. The dangerous impurities in air are hydrocarbons, especially acetylene.

During the rectification process, if the acetylene in the liquefied air should concentrate to a certain degree, then they would be the possibility of explosion, therefore it is require that the acetylene content in the liquefied air shall never exceed 0.1 ppm, and this must be given more attention and consideration.

For those non-condensing gases in the rare gases such as neon and helium, because their condensation temperature are very law, so they always concentrate in the condensing evaporator in the form of gas, thereby occupying the heat exchanger area and affecting the effectiveness of heat exchanger process, therefore it is also necessary to discharge them regularly.

Rectification of air is simply leveraging the fact that various component that make up air have different volatility, i.e. at the same temperature the vapor pressure of various component are different, to achieve the goal of separation by allowing liquefied air to be partially evaporated and partially condensed for multiple time. When oxygen-nitrogen mixture gas of the condensing temperature contact with and goes through the oxygen –nitrogen liquid of a lower temperature, the heat and mass exchange will simultaneously take place between the gas phase and liquid phase, therefore the gas will be partially condensed and transformed into liquid and release condensing latent heat, while the liquid will absorb heat and be partially evaporated.

Within the rectification tower, this process will a accomplish on the sieve plate, because the boiling point of oxygen and nitrogen component are Different, nitrogen is easier to volatize than oxygen, while oxygen is easier to condense than nitrogen, therefore when gas passes from top to bottom through the column plate one by one, the concentration of nitrogen will continue to increase, and if there is a sufficient number of plates, then it is possible to obtain high purity nitrogen at the top of the tower (KRC,2011).

2.1.7 Discharge of dangerous impurities:

The dangerous impurities in the air are hydrocarbons, especially acetylene. During the rectification process, if the acetylene in liquefied air should concentrate to a certain degree, then there would be the possibility of explosion, therefore it is necessary to discharge the liquid regularly, in order to insure that the acetylene concentration in the liquid inside the tower will not exceed the standard limit.

Within the condensing evaporator, the purity of oxygen-enriched liquefied air is approximately 58.68% O2, but for the sake of safety, it is still required that the acetylene concentration not exceed 0.1 ppm (KRC, 2011).

2.1.8 Product pressurizing and output discharge:-

The 0.7Mpa low pressure nitrogen leaving the tower goes through the pipeline to be directly supplied into the nitrogen pipeline network; the 0.7Mpa liquefied nitrogen is filled into the liquefied nitrogen storage tank, is then pressurize by the liquefied nitrogen pump to the pressure of 2.5 MPa, and finally gasified with the gasify to the normal temperature and sent into the medium pressure nitrogen pipeline; the 0.02MPa oxygen leaving the tower is pressurized by the oxygen compressor to ~ 14.7 MPa and filled into the high pressure seamless steel bottle, and then will be supplied to the user (KRC,2011).

2.2 Process description:

The raw material air from the atmospheric is sucked into the screw compressor or centrifugal compressor after being filtered by the anti-desert-storm filter, and is then compressed to 0.8 ~0.88MPa.

the compressed air is cooled by the circulate water and then enter the precooler, where it cooled down to 5 ~8C, and then goes through the primary/secondary dual-level filter to get rid of free from moisture and oil mist, enter the purifier, and then move via the molecular sieve adsorber to adsorb and eliminate impurities in the air, including moisture, carbon dioxide, and acetylene, etc, so that the air will be purified to become the process air.

The process air enters into the top heat exchanger and bottom heat exchanger of the distillation tower to exchange heat with the backflow low temperature gas, is then partially liquefied and enters the rectification tower, where the vapor rise to the tower top to be condensed within the condensing evaporator, and the condensed liquid will act as the backflow liquid to be rectified together with the rising vapor, so that it is possible to obtain high purity nitrogen from the tower top and liquefied air from the tower bottom; while the liquefied air is throttle via throttle valve and sent into the condensing evaporator to condense nitrogen (KRC,2011).

The gasified oxygen enriched liquefied air is reached via the bottom heat exchanger and enters into the turbo expander to be expanded and rectified, thereby providing the enter equipment with the great majority of cold volume.

The expanded oxygen-enriched air is then reheated via the bottom heat exchanger and the top heat exchanger to the normal temperature, and is sent to the purifier to be available for regeneration.

The high purity nitrogen obtain from the top of the nitrogen tower is then reheated and sent into the nitrogen pipe network as the low pressure nitrogen; at the same time, apportion of the liquefied nitrogen is taken from the tower top and feed into the liquefied nitrogen storage tank as the product.

Within the condensing evaporator, apportion of oxygen-enriched liquefied air is taken as backflow liquid for the oxygen tower; apportion of nitrogen is taken from the top of the nitrogen tower to serve as the heating air source for the reboiler, heat and mass exchanger will take place within the oxygen tower, product oxygen is then obtain from the bottom of the tower and is reheated to be discharge from the distillation tower; dirty nitrogen is obtain from the top of the oxygen tower and is mixed with the expanded oxygen-enriched air to be discharge as the regenerated air for the purifier.

The process has two kind of operating condition; when no oxygen is needed, it is possible to close the feed valve of the oxygen tower; when the process is used to produce either nitrogen or liquefied nitrogen alone, its yield of nitrogen or liquefied nitrogen can all increase (KRC, 2011).

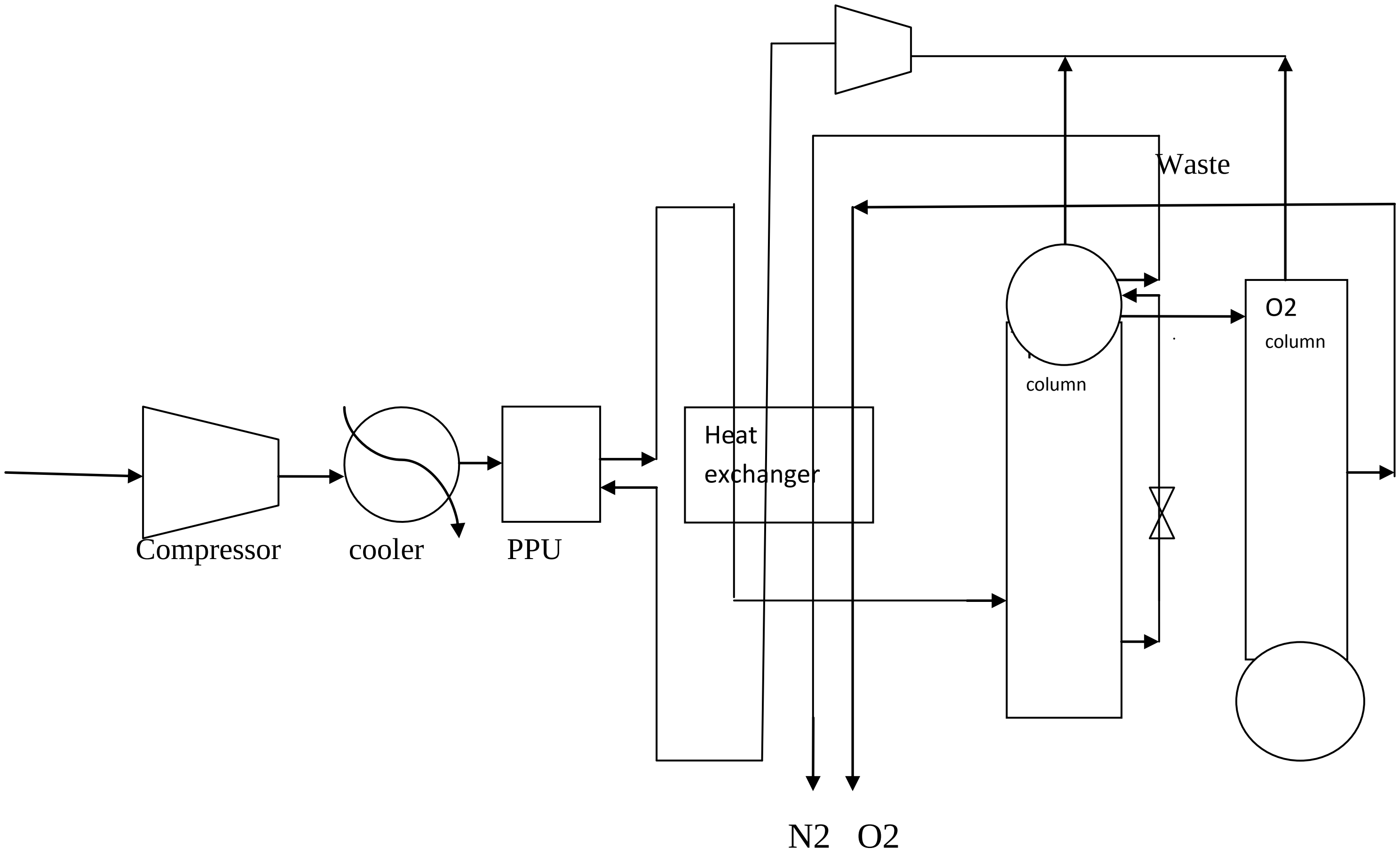


Figure 2.1 air separation unit

2.3 Major equipment:-

2.3.1 Compressor:-

Comprise of the following equipment:

- Inlet air filter.
- Compressor and self-contained electric motor.
- Cooler-equipped supercharger cooling system.
- Separation system.
- Inlet flow control system equipped with step motor.
- Motor startup control system.
- Instrument.
- Safety device.
- After cooler.
- Water separator and water drainer.

The air goes through the inlet filter to eliminate mechanical impurities and dust, and then enter the compressor; the compressed air enter the gas-liquid separator to eliminate oil, then it cooled by the after cooler, enters the drain sum to eliminate water, and then leaves the machine set to become available for use in the precooler set of the downstream air separation facility.

The lubricating oil separated from the gas-liquid separation tank return to the pressure lubrication system to be reused (KRC, 2011).

2.3.2 Precooler:

It gives the initial cool to the feed air by using water.

2.3.3 Purifier:

Comprise of two molecular sieve adsorption cylinders.

Function: the molecular sieve is used to adsorb impurities in air, such as moisture, carbon dioxide, and hydrocarbons, etc, in order to clean the air entering the cold box, prevent clogging in the low temperature channels of the main heat exchanger, and ensure normal operation of the air separation facility.

The working process: air leave the precooling system, with the free form water being separate out, such air then enters the primary/ secondary dual-level filter to further separate out the moisture and oil mist carried, next, the air enters the molecular sieve adsorber, with the help of the adsorption properties of the molecular sieve, the moisture, carbon dioxide, and hydrocarbon in the compressed air will be adsorbed.

After being used for a certain cycle of time, the molecular sieve will need to be reheated and regenerated; during the regeneration process the adsorbent is first of all reversely heated and regenerated by the dirty nitrogen that has been heated by the electric heater so that such impurities adsorbed as moisture, carbon dioxide, and hydrocarbons will be taken out of the molecular sieve adsorber; next, it’s blow cooled by the dirty nitrogen gas coming from the distillation tower until reaching the temperature; the two sets of molecular adsorption cylinder will be working alternatively, when one cylinder is working to adsorb, the other will be regenerated and reactivated (KRC,2011).

2.3.4 Distillation tower:

Wholly sealed cold box comprise of the main heat exchanger, the nitrogen rectification tower, the condensing evaporator, the oxygen rectification tower, the reboiler, the sub cooler, and gas bearing turbo expander, etc.

Function: the cleaned and purified compressed air is cooled and liquefied in the main heat exchanger of the distillation tower, is rectified in the nitrogen rectification tower and separated into nitrogen and rich oxygen, a portion of the oxygen- enriched liquefied air is taken from the condensing evaporator on the top of the nitrogen rectification tower and sent to the oxygen rectification tower, within the oxygen rectification tower, it is rectified and separated into oxygen and dirty nitrogen, where the cold volume is provide by the expander (KRC,2011).

2.3.5 Main heat exchanger:

Structure: multiple-layer plate-type finned structure, and comprise of three serially connected plate heat exchangers.

Function: it can conduct heat exchange between multiple streams of fluids, and the cold and hot air flows in multiple channels can conduct good heat exchange via the fins and baffle plates. The cold air flow will be reheated to the normal temperature, while the hot air flow will be cooled to the needed temperature.

Method of use: the purified compressed air moves from top to bottom to enter the main heat exchanger to conduct heat exchange, the air is cooled until it reaches the liquefaction temperature, approximately 5% of air will be liquefied; while the low temperature products discharged from the rectification tower, such as nitrogen, oxygen, and oxygen-enriched air, etc, will move from the bottom to top to enter the multiple channels of the main heat exchanger and enters the turbo expander to be expand, the expanded air then enters the bottom of the main heat exchanger to be reheated to the normal temperature (KRC,2011).

2.3.6 Nitrogen rectification tower:

Structure: Vertical cylinder shapes, instilled inside are sieve pore column plates, which is of the single-overflow convection structure.

Function: utilize the fact that the boiling points of multiple components in air are different, and using the rectification process to separate air into the needed ingredient.

Method of use: during the rectification process, the liquid moves from top to bottom to flow across each column plate; given the function of the overflow weir, a certain height of liquid layer will be created on the column plates, so that when the air moves from bottom to top to cross the tiny orifices in the sieve plates, it will contact with the liquid to generate bubble, in this way, the area of gas-liquid contact will be enlarged, and the heat exchanger process will progress efficiently (KRC,2011).

During the heat-mass exchange process, those components with relatively lower boiling points will gradually vaporize, and those with relatively higher boiling point.

2.3.7 Condensing evaporator:

Structure: a set of plate-type finned heat exchanger that is installed inside the external cylinder of the condensing evaporator.

Function: this is used for condensing of nitrogen and vaporizing of liquefied air, in order to maintain progress of the rectification process in the rectification tower.

Method of use: for the nitrogen taken out from the top of the rectification tower, one portion enters to the nitrogen channel of the plate-type finned heat exchanger and is condensed by the throttled liquefied air into liquefied nitrogen; one part of it will serve as the product liquefied nitrogen for output, but the majority of it will serve as the backflow liquid; the liquefied air enter into the evaporation side of the condensing evaporator to be evaporated, and will then be outputted as the air source for the expander (KRC,2011).

2.3.8 Oxygen rectification tower:

Structure: vertical cylinder shape, installed inside are sieve pore column plates, which is of the single overflow convection structure, the bottom of it is equipped with a reboiler.

Function: utilizing the fact that the boiling point of oxygen, nitrogen, and argon in air are different, and using the rectification process to separate air into the needed ingredients.

Method of use: during the rectification process, the liquid move from top to bottom to flow across each column plate; given the function of overflow weir, a certain high of the liquid layer will be created on the column plates, so that when the air moves from bottom to top to cross the tiny orifices in the sieve plates, it will contact with the liquid to generate bubbles, in this way, the area of gas-liquid contact will be enlarged, and the heat exchanger process will progress efficiently.

During the heat-mass exchanger process, those components with relatively lower boiling points will gradually vaporize, and those with relatively higher boiling points will gradually liquefy; in this way, as long as there is a sufficient number of a column plates, it is possible to obtain relatively high purity oxygen (the ingredient with a higher boiling point) from the bottom of the tower, and it is also possible to obtain dirty nitrogen from top of the tower (KRC,2011).

2.3.9 Reboiler:

Structure: plate-type finned heat exchanger, which is installed at the bottom of the cylinder body of the oxygen rectification tower.

Function: this can allow the oxygen-enriched liquefied air to vaporize and nitrogen to condensed, in order to maintain the normal rectification process in the oxygen rectification tower.

Method of use: a portion of nitrogen is taken out from the top of the nitrogen rectification tower to serve as a heat source for the rectification process in the oxygen tower, these heat source will reboil the tower bottom at the bottom of the oxygen tower to allow the operation to obtain 99.6% pure oxygen from the bottom of the oxygen tower, while the nitrogen will be condensed into liquefied nitrogen that will enter into the top of the nitrogen rectification tower (KRC,2011).

2.3.10 Sub cooler:

Structure: plate type finned heat exchanger.

Function: utilizing the cold volume of the dirty nitrogen coming from the top of the oxygen tower to super cool the oxygen-enriched liquefied air coming from the bottom of the nitrogen rectification tower, so as to reduce the rate of the gasification of the oxygen-enriched liquefied air that enters into the cylinder body of the condensing evaporator (KRC,2011).

2.3.11 Turbo expander set:

Structure: horizontal single-stage gas bearing expander.

Function: the oxygen-enriched air is expanded in the turbo expander to provide cold volume to the distillation tower.

Method of use: after the backflow oxygen - enriched air is reheated in the bottom heat exchanger to the needed temperature, it enters into turbo expander to be expanded and sent into the heat exchanger, and the work of the expander will be absorbed by its coaxial oppositional braking blower fan. The bearing air and sealing air used by the turbo expander will be supplied by the bearing gas device (KRC, 2011).

2.3.12 Utilities equipment:-

- Instrument air and bearing air system.

- Liquefied nitrogen pump.

- Oxygen compressor.

- Liquefied nitrogen storage tank.

- Oxygen filler.

2.4 Simulation:

Process simulations enable us to find stable operation point with optimal condition for the process examined.

We use hysys for simulation instate of Aspen because hysys is available, and we use three heat exchanger instate of the multiple heat exchanger.

HYSYS programmed used to simulate the process to know the effect of change process parameters on the performance and analysis the results obtained.

The programme ASPEN HYSYS version 3.1 used for simulation the process described above with the following data tabulated .also the component of air with volume fraction in table 2.1 used and the selected fluid package is peng robinson.

Table 2.3 model air specifications (KRC,2011)

Stream name	Pressure (Mpa)	Temperature (C)	Flow rate Nm^3/hr
F1	0.1014	35	2000

Table 2.4 Design and operation condition f distillation column (KRC,2011)

Property	First column (N2)	Second column(O2)
Condenser Pressure	90 kpa	77 kpa
Reboiler Pressure	100 kpa	90 kpa
Condenser Delta P	0	0
Reboiler delta P	0	0
Number of stage	40	20
Feed inlet stage	20	10
Reflux ratio	4	3
Distillate rate	1.011kgmole/hr	-
Overheat vap rate	26.77kgmole/hr	-
Bottom product rate	-	7 kgmole /hr

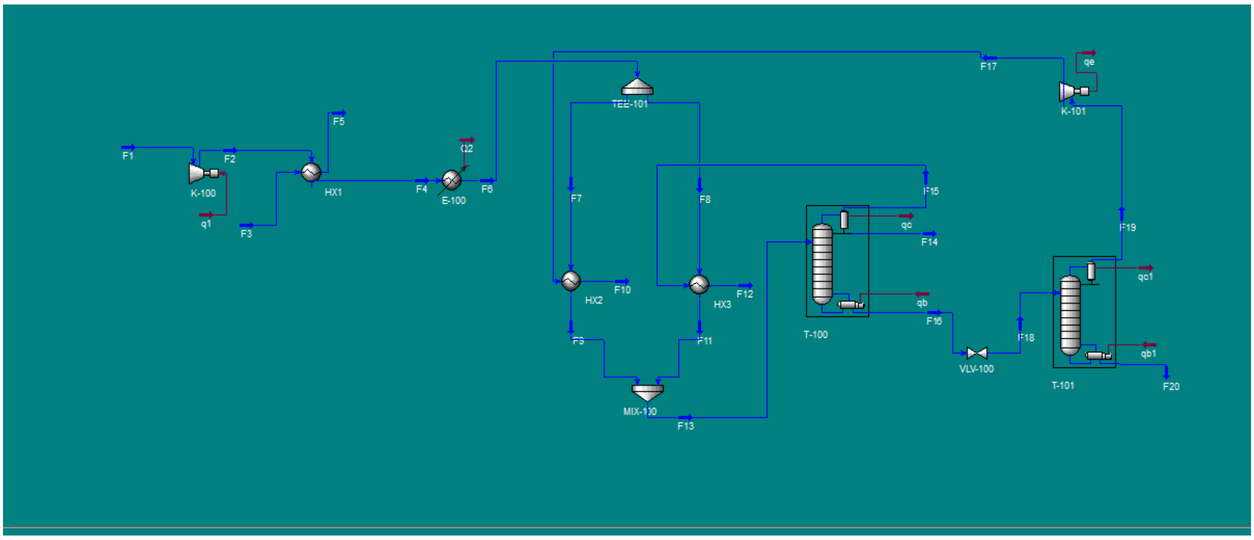


Figure 2.2 Cryogenic air separation processes

Chapter 3

Material and Energy balance

3. Material and energy balance

Material and energy balance are very important in an industry. Material balance describes the material quantity as they pass through processing operation. Such balances are statements on the conservation of mass. Similarly energy balance describes energy quantities, which are statement on the conservation of energy.

For the process there are two types of material and energy balance, overall balance and individual balance around each equipment. Calculated by following general law for material and energy balance.

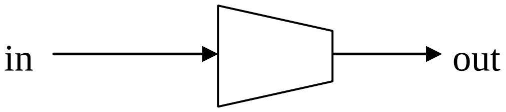
In+ generation- consumption = out

3.1 Individual material and energy balance:

Around compressor:

Material balance:

F = feed (kg.mol/hr) H = enthalpy (kj/kg.mole)



F1in= 89.23kg.mol/hr F2out = 89.23kg.mol/hr

In-out =accumulation

89.23-89.23 = accumulation → accumulation = 0 kgmol/hr

Energy balance:

Hin = 283.5kj/kgmole Hout= 7491kj/kgmole consumption = 0kj/kgmole

In+ generation- consumption = out

283.5+ gen = 7491 → generation = 7207.5kj/kgmol

Around heat exchanger1:

Material balance:

F2in =89.23kgmol/hr F3in = 45.36kgmol/hr

F4out = 89.23kg.mol/hr F5out = 45.36kgmol/hr

In-out =accumulation

(89.23+45.36) – (45.36+89.23) = accumulation → Accumulation = 0 kg.mol/hr

Energy balance:

H2in = 7491 kj/kgmole H3in = -0.0003 kj/kgmole generation = 0 kj/kgmole

H4out = -0.00003 kj/kg.mole H5out = 411.3 kj/kg.mole

In + generation – out = consumption

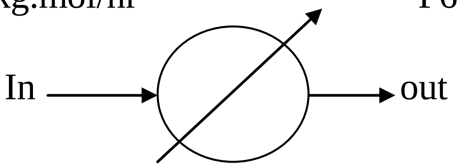
(7491 -0.00003)– (411.3+0.00003)= consumption

Consumption = 7079.7 kj/kgmole

Around cooler:

Material balance:

F4in = 89.23 kg.mol/hr F6out = 89.23 kg.mol/hr



In-out =accumulation

(89.23 – 89.23) = accumulation → accumulation = 0 kg.mol/hr

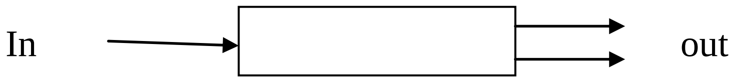
Energy balance:

Hin = 411.3 kj/kgmole generation = 0 kj/kgmole Hout = -562.4 kj/kgmole

In + generation – out = consumption

411.3+ 0 – (-562.4) = consumption → consumption = 973.7 kj/kgmole

Around T-junction:



F6in = 89.23 kg.mol/hr F7out = 31.23 kg.mol/hr

F8out = 58 kg.mol/hr

Material balance:

In-out =accumulation

89.23 – (13.23+58) = accumulation → accumulation = 0 kg.mol/hr

Energy balance:

Hin = -562.4 kj/kgmole Hout = -562.4 kj/kgmole generation = 0 kj/kgmole

In + generation – out = consumption

(-562.4) + 0 – (-562.4) = consumption → consumption = 0 kj/kgmole

Around heat exchanger 2:

Material balance:

F7in = 58kg.mol/hr F20in = 54.45kg.mol/hr

F9out = 58 kg.mol/hr F10out = 54.45kg.mol/hr

In-out =accumulation

(58 + 54.45) –(58 +54.45) = accumulation → accumulation = 0kg.mol/hr

Energy balance:

Hin = -7397.4 kj/kgmole generation = 0 kj/kgmole Hout = -7051 kj/kgmole

In + generation – out = consumption

-7397.4 + 0 – (-7051) = consumption → consumption = -346.4 kj/kgmole

Around heat exchanger 3:

Material balance:

F15in = 26.77kg.mol/hr F8in = 31.23kg.mol/hr

F11out = 31.23kg.mol/hr F12out = 26.77kg.mol/hr

In-out+ generation =accumulation

(26.77 + 31.23) – (31.23 +26.77)+0 =accumulation

→ accumulation=0kg.mol/hr

Energy balance:

H15in = -6389 kj/kgmole H8in=-562.4 kj/kgmole

H12out = -5555 kj/kgmole H11out=-564.7 kj/kgmole

consumption =0 kj/kgmole

In + generation – out = consumption

(-6389-562.4)+ generation –(-5555-564.7) = 0 →generation = -746.4 kj/kgmole

Around mixture:

F11in = 31.23kg.mol/hr F9in = 58kg.mol/hr F13out = 89.23kg.mol/hr

Material balance:

In-out =accumulation

(31.23 + 58) – 89.23 = accumulation → accumulation = 0kg.mol/hr

Energy balance:

H11in = -5555kj/kgmole H9in=-5485 kj/kgmole H13out = -5578 kj/kgmole generation = 0 kj/kgmole

In + generation – out = consumption

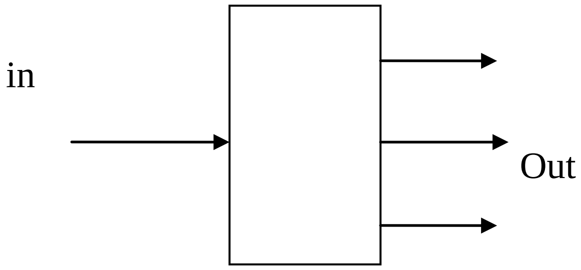
(-5555-5485)+ generation – (-5578) = 0→ generation = -5462 kj/kgmole

Around distillation:

Material balance:

F13in = 89.23kg.mol/hr F15out= 26.77kg.mol/hr

F14out = 1.011kg.mol/hr F16out = 61.45kg.mol/hr



In-out =accumulation

(89.23) –(1.011 + 61.45+26.77) = accumulation

Accumulation =0 kg.mol/hr

Energy balance:

H13in = -5578 kj/kgmole H14out=-0.0001198 kj/kgmole

H15out = -6389 kj/kgmole H16out= -0.0001226 kj/kgmole

generation = 0 kj/kgmole

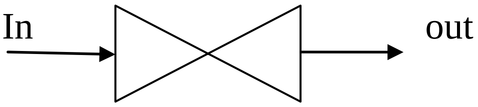
In + generation – out = consumption

(-5578)+ 0 – (-6389-.0001198-.0001226) = consumption → consumption = 811 kj/kgmole

Around throttling valve:

Material balance:

F16in = 89.23kg.mol/hr F17out = 89.23kg.mol/hr



In-out =accumulation

(89.23) –(89.23) =accumulation → accumulation = 0kg.mol/hr

Energy balance:

H16in = -0.0001kj/kg.mole H17out = -0.0001kj/kgmole generation=0 kj/kg.mole

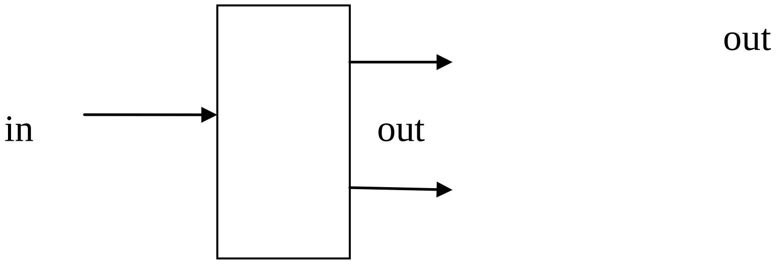
In + generation – out = consumption

(-0.0001) + 0 – (- 0.0001) = consumption → consumption = 0 kj/kgmole

Around O2 column:

Material balance:

F17in =89.23kg.mol/hr F18out = 7kg.mol/hr F19out = 54.45kg.mol/hr



(89.23)- (7 + 54.45) = accumulation → accumulation = 27.78kg.mol/hr

Energy balance:

H17in = -0.0001kj/kgmole   H18out = 0.0001292 kj/kgmole

H19out=-- 6351 kj/kgmole   generation = 0 kj/kgmole

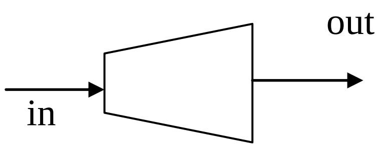
In + generation – out = consumption

(-0.0001)+ 0 –(-6351-0.0001292) = consumption   → consumption = 6352.0 kj/kgmole

Around expander:

Material balance:

F19in = 54.45kg.mol/hr   F20out = 54.45kg.mol/hr



In-out =accumulation

54.45 -54.45 = accumulation   → accumulation = 0kg.mol/hr

Energy balance:

H19in = -6290 kj/kgmole   H20out = -6835 kj/kgmole   generation = 0 kj/kgmole

In + generation – out = consumption

(-6290) + 0 – (-6835) = consumption   →   consumption = 545 kj/kgmole

3.2 overall material and energy balance:

Material balance:

F1in = 89.23kg.mol/hr   F3in=45.36 kg.mol/hr   F5out = 45.36kg.mol/hr

F12out = 26.77kg.mol/hr   F10out = 54.45kg.mol/hr F14out=1.011 kg.mol/hr

F18out=7 kg.mol/hr

In + accumulation = out

Accumulation = (45.3+26.77+54.45+1.011+7) – (89.23+45.36) = 0 kg.mol/hr

In =out= 134.59 kg.mol/hr

Energy balance:

H1in = 283.5 kj/kgmole   H3in=-0.0000874 kj/kgmole

H5out = -0.0002734kj/kgmole   H12out=-564.7 kj/kgmole

H10out=-539.2 kj/kgmole   H14out=-0.0001198 kj/kgmole

H18out=-0.0001292 kj/kgmole

generation = 0kj/kgmole

In + generation – out = consumption

consumption= (283.5-.0000874) + 0 – (-0.000273-564.7-539.2-0.0001198-0.0001292) = 1387.4004 kj/kgmole

Chapter 4

Design of Shell and Tube Heat Exchanger

4. Design of shell and tube heat exchanger

Construction:-

A shell and tube exchanger consists of a bundle of tubes enclosed in a cylindrical shell.

The ends of the tubes are fitted into tube plates.

Baffles are provided in the shell to direct the fluid flow and support the tubes.

The assembly of baffles and tubes is held together by support rods and spacers (Coulson Richard, 2003).

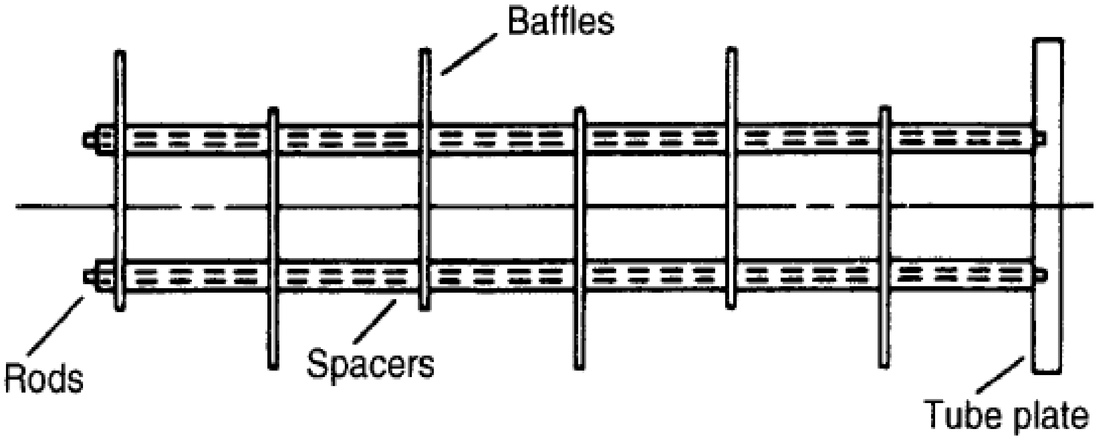


Figure 4.1 shell and tube heat exchanger (KRC, 2011)

4.1. Fluid Location:-

-Air located in the shell side and nitrogen located in the tube side to avoid losses of heat.

4.2. Selection of Flow path:-

In selecting the flow path for two fluids through an exchanger, several general approaches are used:-

4.2.1- The tube side

Fluid is more corrosive or dirtier or at a higher pressure.

4.2. 2- The shell-side

Fluid is a liquid of high viscosity or a gas.

4.2. 3- When alloy construction for one of the two fluids is required, a carbon Steel shell combined with alloy tube-side parts is less expensive than alloy in contact with the shell-side fluid combined with carbon steel headers.

4.2.4- Cleaning of the inside of tubes is more readily done than cleaning of exterior surfaces.

4.2.5- For gauge pressures in excess of 2068Kpa (300 lbf/in2) for one of the fluids, the less expensive construction has the high-pressure fluid in the tubes.

4.2.6- For a given pressure drop, higher heat-transfer coefficients are obtained on the shell side than on the tube side.

4.2.7-Heat-exchanger shutdowns are most often caused by fouling, corrosion, and erosion (Coulson Richard, 2003).

4.3. Heat-exchanger standards and codes:-

The mechanical design features, fabrication, materials of construction, and testing of shell and tube exchangers is covered by British Standard, BS 3274. The standards of the American Tubular Heat Exchanger Manufacturers Association, the TEMA standards, are also universally used. The TEMA standards cover three classes of exchanger: class Recovers exchangers for the generally severe duties of petroleum and related industries; class C covers exchangers for moderate duties in commercial and general process applications; and class B covers exchangers for use in the chemical process industries (Coulson Richard,2003).

The British and American standards should be consulted for full details of the mechanical design features of shell and tube exchangers; only brief details will be given in this chapter.

The shell of an exchange is a pressure vessel and will be designed in accordance with the appropriate national pressure vessel code or standard; the specification of standard flanges for use with heat exchangers is given in the table below.

Table 4.1 Data sheet (Coulson Richard, 2003).

Component	Air	Nitrogen
Tin	8^0cC	-198^0C
Tout	-15^0C	-30^0C
Flow rate	1691.11kg/s	1014.11kg/s
Viscosity	$0.01827 * 10^{-3}\, {}^{mNs}/_{m^2}$	$0.0178 * 10^{-3}\, {}^{mNs}/_{m^2}$
Density	$9.95kg/m^3$	$\frac{8181.5kg}{m^3}$
Thermal conductivity	0.0257w/m.c	0.0077w/m.c
Specific heat	$\frac{29.257j}{kg.C^0}$	$\frac{56.68j}{Kg.{}^\circ C}$

4.4. Design Steps:-

4. 4.1 Assume material of construction for the tubes (thermal conductivity).

Table 4.2 conductivity of metal (coulson Richard, 2003).

Metal	Temperature (C)	Kw (W/m C)
Aluminum	0 100	202 206
Brass	0	97
(70 cu, 30 Zn)	100 400	104 116
Copper	0 100	388 378
Nickel	0 212	62
Cupro-nickel (10% Ni)	0 −100	45
Monel	0 −100	30
Stainless steel (18/8)	0 −100	16
Steel	0 100 600	45 45 36
Titanium	0 −100	16

4.4.2- Heat duty equation q = [mCp (Tout–Tin)] air (1)

Where: q = heat duty m = mass cp = specific heat

4.4.3- Calculate Log Mean Temperature Difference, LMTD

$$\Delta Tlm = \frac{(\Delta T1-\Delta T2)}{Ln(\Delta T1/\Delta T2)} \qquad (2)$$

4.4.4- Calculate Temperature correction factor ft from below figure.

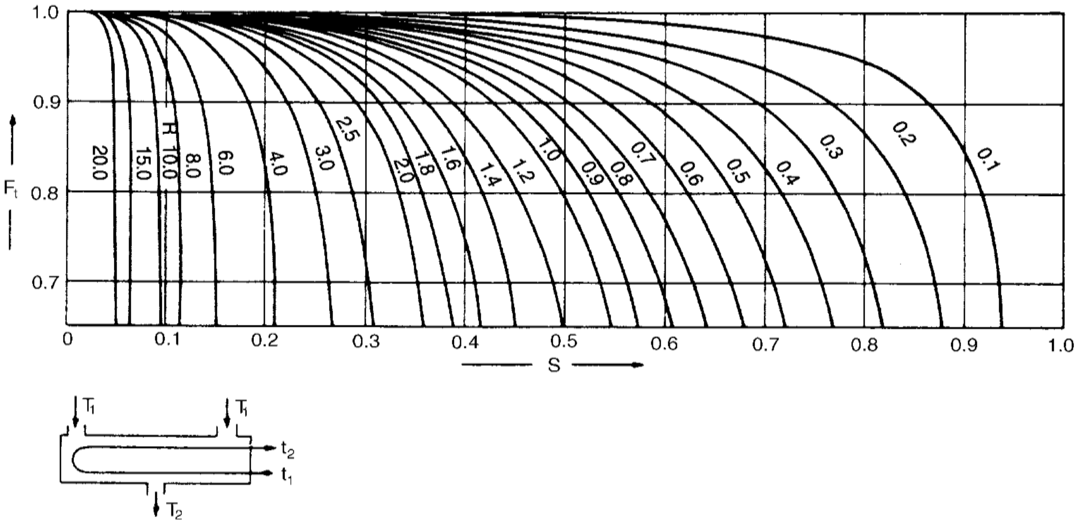


Figure 4.2 Temperature correction factors (Coulson Richard, 2003).

4.4.5- Calculate the mean temperature difference using $\Delta Tlm = Ft * LMTD$ (3)

Ft = temperature correction factor.

4. 4.6-Assume overall heat transfer coefficient as initial guess from the table below:

Table 4.3 Typical overall coefficient shell and tube exchangers (Coulson Richard, 2003).

Hot fluid	Cold fluid	U (W/m2 C)
Heat exchangers		
Water	Water	800−1500
Organic solvent	Organic solvent	100 – 300
Light oils	Light oils	100 – 400
Heavy oils	Heavy oils	50 – 300
Gases	Gases	10 – 50
Coolers		
Organic solvent	Water	250 – 750
Light oils	Water	350 – 900
Heavy oils	Water	60 – 300
Gases	Water	20 – 300
Organic solvent	Brine	150 – 500
Water	Brine	600 – 1200
Gases	Brine	15 – 250
Heaters		
Steam	Water	1500 – 4000
Steam	Organic solvent	500 – 1000
Steam	Light oils	300 – 900
Steam	Heavy oils	50 – 300
Steam	Gases	20 – 200
Dowthem	Heavy oils	50 – 300
Dowthem	Gases	20 – 200
Flue gas	Steam	30 – 100
Flue	Hydrocarbon vapors	30 – 100
Condensers		
Aqueous vapor	Water	1000 – 1500
Organic vapor	Water	700 – 1000
Organics(some non-condensable)	Water	500 – 700
Vaporizers		
Steam	Aqueous solution	1000 – 1500
Steam	Light organics	900 – 1200
Steam	Heavy organic	6000 −9000
Air- condenser exchanger		
Process fluid		
Water		300 – 450
Light organics		300 – 700
Heavy organics		50 – 150
Gases, 5-10 bar 10-30 bar		50 – 100 100 – 300
Condensing hydrocarbons		300 – 600

4.4.7- calculate the area of heat transfer A=$q/U\Delta Tlm$ (4)

U = over all heat transfer coefficient

4.4.8- Based on the assumed tube diameter and tube Length, L, calculate number of tubes:

Nt =$\frac{\boldsymbol{A}}{\boldsymbol{\pi.do.l}}$ (5)

Where Nt = number of tubes.

d◦ = tube outside diameter, mm.

4.9- Calculate the bundle diameter

Pt = 1.25d◦ (6)

Db=d◦ $(\frac{Nt}{K1})$ ^ (1/n1) (7)

Db = bundle diameter, mm.

Where 1 K and 1 n are obtained from the table bellow based on the type of tube Arrangement (Triangular or square pitch)

Table 4.4 constant for use in equation 7 (Coulson Richard, 2003).

Triangular pitch, Pt = 1.25d◦					
No. passes	1	2	4	6	8
K1	0.319	0.249	0.175	0.0743	0.0365
N1	2.142	2.207	2.285	2.499	2.675
Square pitch, Pt = 1.25d◦					
No. passes	1	2	4	6	8
K1	0.215	0.156	0.158	0.0402	0.0331
N1	2.207	2.291	2.263	2.617	2.643

4.4.10- Provide/Assume the type of floating head of the exchanger and obtain bundle Diameter clearance, BDC. Use the chart below:

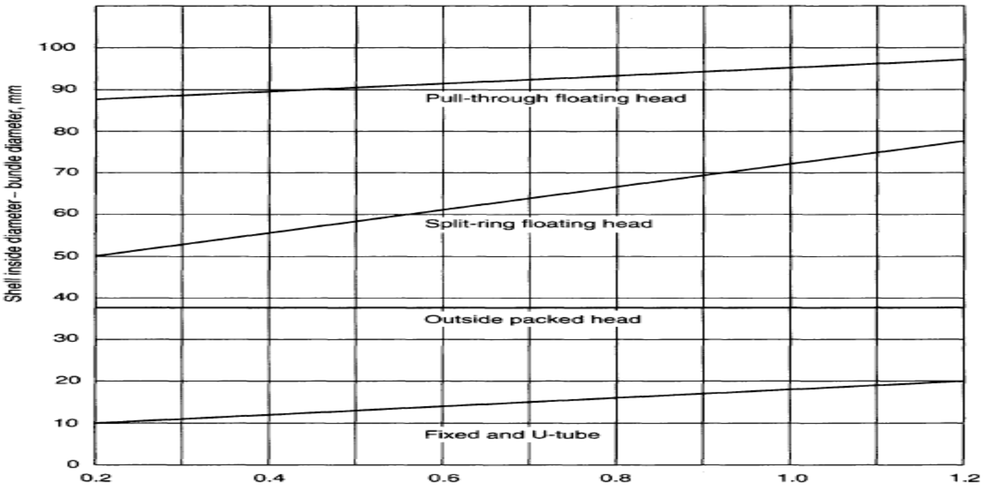


Figure 4.3 type of floating head of the exchanger (Coulson Richard, 2003).

4.4.11- Calculate the shell diameter Ds=Db + BDC (8)

Ds = shell diameter BDC = bundle Diameter clearance

4.4.12- Calculate the baffle spacing Bs=0.4Ds (9)

4.4.13- Calculate the area for cross-flow as=$(Pt - d0)Ds.\frac{Bs}{Pt}$ (10)

4.4.14- Calculate the shell equivalent diameter

4.4.15- Calculate the shell-side Reynolds number

$$\text{Re} = \frac{\boldsymbol{Gsde}}{\boldsymbol{\mu}} = \frac{\boldsymbol{us.de.\rho}}{\boldsymbol{\mu}} \quad (11)$$

Gs = flow rate de = equivalent diameter

ρ = fluid density = fluid viscosity

4.4.16- Calculate Prandtle number

Pr= $\mu cp/k$ (12)

4.4.17-Obtain the shell-side heat transfer coefficient

$$\textbf{Nu} = \frac{\boldsymbol{hs.de}}{\boldsymbol{Kf}} = \textbf{jh.Re.}\boldsymbol{Pr^{1/3}}\left(\frac{\boldsymbol{\mu}}{\boldsymbol{\mu w}}\right)^{\mathbf{0.14}} \quad (13)$$

Where jh is obtained from the chart bellow

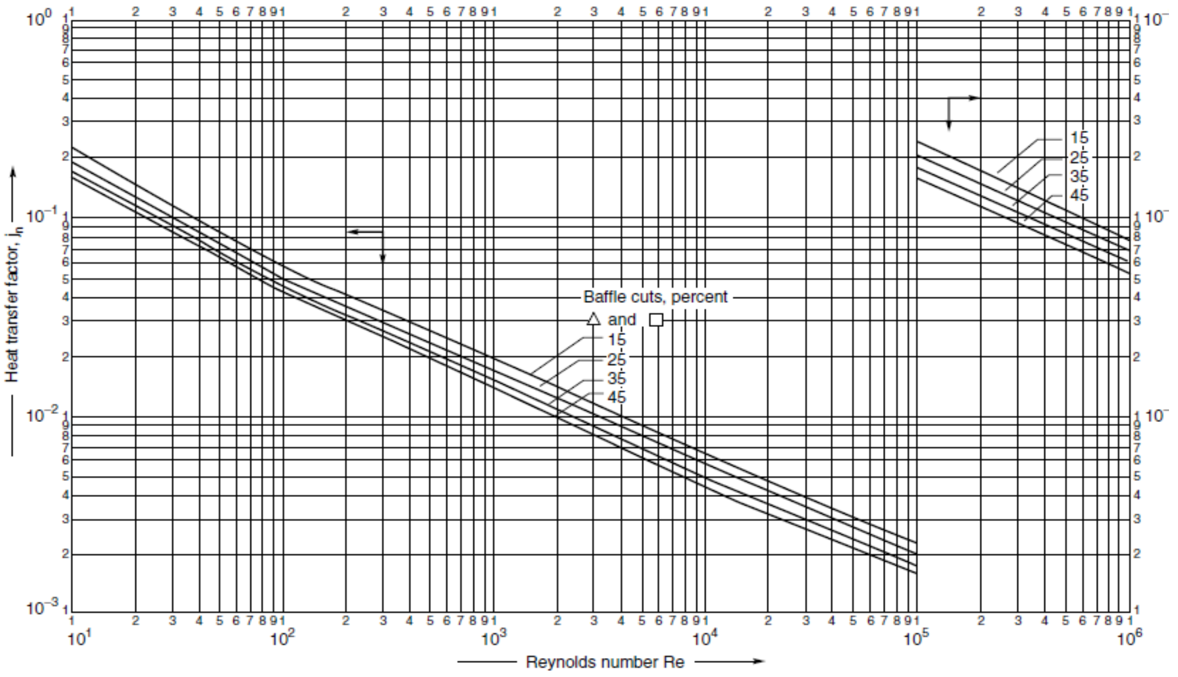


Figure 4.4 shell-side-transfer factors (Coulson Richard, 2003).

4.4.18-Calculate the pressure drop in the shell

$$\Delta PS = 8jf\left(\frac{Ds}{de}\right)\left(\frac{L}{lB}\right).\frac{\rho us^2}{2}\left(\frac{\mu}{\mu w}\right)^{-0.14} \qquad (14)$$

Where L = tube length. lB = baffle spacing

4.4.19- Calculate the number of tubes per passes, Ntp =Nt/number of passes.

4.4.20- Calculate tube-side velocity v=$\mu cp/\rho i$

ρi = the density of fluid inside tubes.

4.4.21- Calculate Prandtle and Reynolds numbers for fluids inside tubes.

4.4.22- Calculate heat transfer coefficient hi by using the following relation.

4.4.23- Calculate the overall heat transfer factor.

Based on “inside tubes flow” Ui $= \dfrac{1}{\frac{1}{h1}+\frac{1}{hdi}+\frac{di\ ln\,(d°/di)}{2Kw}+\frac{di}{d°\,hd°}+\frac{d°}{d°\,h°}}$ (15)

Or based on “outside” = Uo $= \dfrac{1}{\frac{1}{h°}+\frac{1}{hd°}+\frac{d°\ ln\,(d°/di)}{2Kw}+\frac{d°}{d1\,h°}+\frac{d°}{di\ hdi}}$ (16)

Where di h and do h are the heat transfer coefficients for the scales (dirt) inside an outside tubes, respectively.

4.4.24-Compare the calculated overall heat transfer coefficient you obtained from the previous step with that you assumed in step 8. If it is close to what you assumed, then you had a valid assumption, and then tabulate your results such as total surface area of tubes, number of tubes, exchanger length and diameter, heat duty and other design specification. Otherwise, use the calculated value in step 8 and do loop until the difference between the calculated U between two consecutive iterations is small.

5. Design calculations:-

Heat load (Q) = [mCp (Tin–Tout)] air=(1691.11kg/s)(29.257J/kg·°C)(8-(-15))°C = 1137966.52W

ΔT1 = the, in - Tc out = (8-(-30)) =38ºc

ΔT2 = The, out - Tc in = (-15-(-198)) =183

$$\Delta Tlm = \frac{(\Delta T1 - \Delta T2)}{Ln\left(\frac{\Delta T1}{\Delta T2}\right)} = \frac{38 - 183}{ln\left(\frac{38}{183}\right)} = 92.25$$

$$R = \frac{8-(-30))}{(-15-(-198))} = 0.2$$

S = $\frac{(-15-(-198))}{(8-(-198))} = 0.89$

From Figure 12.19

Ft = 0.96

$\Delta Tlm =$ 0.96*92=88.5

Assume U=30

A=Q /U$\Delta T = \frac{1137966.52}{30*88.5} = 428.6m2$

Choose 20 mm do, 16 mm di, 4.88-m-long tubes, cupro-nickel.

Allowing for tube-sheet thickness, take

L = 4.83 m, and area of one tube = 4.83×20×10-3π =0.303m2

Number of tubes =428.6 / 0.303 = 1414

As the shell-side fluid is relatively clean use 1.25 triangular pitches

Bundle diameter Db = $20\left(\frac{1414}{0.249}\right)^{\frac{1}{2.207}} = 1005mm$

From Figure 12.10, bundle diametrical clearance = 68 mm

Shell diameter, Ds = 1005 + 68 = 1073 mm.

Tube-side coefficient:-

Mean N2 temperature =$\frac{-198-30}{2} = -114^0\ C$

Tube cross-sectional area =$\frac{\pi}{4} * 16^2 = 201m^2$

Tubes per pass=$\frac{1414}{2}$ =707

Total flow area = $707 * 201 * 10^{-6} = 0.142m^2$

Nitrogen mass velocity =$\frac{1014.17}{0.14} = \frac{7242kg}{s.m^2}$

Nitrogen density = 818.5 kg/m3

Nitrogen linear velocity=$\frac{7242}{818.5} = \frac{8.85m}{s}$

$$hi = \frac{4200(1.35 + 0.02 * -114)}{16^{0.2}} * 8.85^{0.8} = -\frac{12836w}{m^2.c^0}$$

hi = tube side coeggicient

Shell-side coefficient:-

Choose baffle spacing = $\frac{Ds}{5} = \frac{1073}{5} = 214.6mm$

Tube pitch = 1.25*20= 25

Cross-flow area as=$\frac{(25-20)}{25} * 1073 * 214.6 * 10^{-6} = 0.046m^2$

Mass velocity, GS =$\frac{1691.11}{0.046} = \frac{36.8kg}{m^2s}$

Equivalent diameter de= $\frac{1.1}{20}(25^2 - 0.917 * 20^2) = \ 14.4\ mm$

Mean shell side temperature =$\frac{8-15}{2} \ = -3.5^0c$

Re = $\frac{Gs.de}{\mu} \ = \frac{36.8*14.4}{0.01827*10^{-3}} = \ 29005$

$$pr = \frac{cp.\mu}{kf} = \frac{29.257 * 0.01827 * 10^{-3}}{0.19} = 2.81$$

Choose 25 per cent baffle cut, from Figure 4.4

$jh = 4.2 * 10^{-3}$

Without the viscosity correction term

$hs = \frac{0.19}{14.4*10^{-3}} * 4.2 * 10^{-3} * 29005 * 2.81^{\frac{1}{3}} = 2268 \frac{w}{m^2c}$

hs = shell side coefficient

Overall coefficient:-

Thermal conductivity of cupro-nickel alloys = 50 W/m C

Take the fouling coefficients from Table 12.2; nitrogen 50W/m.C
And air 100 W/m.C

$$\frac{1}{U0} = \frac{1}{-12836} + \frac{1}{100} + 20 * \frac{10^{-3}\ ln\left(\frac{20}{16}\right)}{2 * 50} + \frac{20}{16} * \frac{1}{50} + \frac{20}{16} * \frac{1}{2268} = \frac{38w}{m^2.c^0}$$

Table 4.5 Summary of design result

1. Heat load (q) = 1137966.52 w
2. $\Delta T\ logarithmeic\ mean($ $\Delta Tlm)$ = 88.5
3. Area of heat transfer (A) = $428.6m2$
4. Shell diameter(Ds) = 1073mm
5. Tube-side coefficient(hi) = -12836w/$m^2.c^0$
6. Shell side coefficient(hs) = 2268w/$m^2.c^0$
7. Overall coefficient(U◦) = 38w/$m^2.c^0$
8. Number of tubes = 1414
9. The tube diameter = 0.02m
10. Tubes per passes = 707
11. Tubes and shell thickness = 0.004m
12. Tube length = 4.88m
13. Bundle diameter = 1.005m

Chapter 5

Costing and Project Evaluation

5. Costing and Project Evaluation

Cost estimation is a specialized subject and a profession in its own right. The design Engineer, however, needs to be able to make quick, rough, cost estimates to decide between alternative designs and for project evaluation. Chemical plants are built to make a profit, and an estimate of the investment required and the cost of production are needed before the profitability of a project can be assessed (Coulson Richard, 2003).

5.1 fixed capital:

Fixed capital is the total cost of the plant ready for start-up. It is the cost paid to the Contractors.

It includes the cost of:

5.1.1. Design, and other engineering and construction supervision.
5.1.2. All items of equipment and their installation.
5.1.3. All piping, instrumentation and control systems.
5.1.4. Buildings and structures.
5.1.5. Auxiliary facilities, such as utilities, land and civil engineering work.

Fixed capital = PPC ＊ (1+f10+f11+f12)

PPC= physical plant cost

f10= Design and Engineering.

f11 = Contractor’s fee.

f12= Contingency.

PPC = PCE * (1+f1+…+f9).

PCE = purchase cost of equipment.

f1 = Equipment erection
f2 = Piping.
f3= Instrumentation.
f4 = Electrical.
f5 = Buildings.
f6 = Utilities.
f7= Storages.
f8= Site development.
f9= Ancillary buildings.

5.2 Working capital:

Working capital is the additional investment needed, over and above the fixed capital, to start the plant up and operate it to the point when income is earned (coulson richard,2003).

It includes the cost of:

5.2.1. Start-up.
5.2.2. Initial catalyst charges.
5.2.3. Raw materials and intermediates in the process.
5.2.4. Finished product inventories.
5.2.5. Funds to cover outstanding accounts from customers.

5.3Operatioo cost:

An estimate of the operating costs, the cost of producing the product, is needed to judge The viability of a project, and to make choices between possible alternative processing Schemes (coulson Richard, 2003).

They are divided into two groups:

5.3.1. Fixed operating costs: costs that do not vary with production rate. These are the Bills that have to be paid whatever the quantity produced.
5.3.2. Variable operating costs: costs that are dependent on the amount of product produced.

5.3.1 Fixed costs:

- Maintenance (labor and materials).
- Operating labor.
- Laboratory costs.
- Supervision.
- Plant overheads.
- Capital charges.
- Rates (and any other local taxes).
- Insurance.
- License fees and royalty payments.

5.3.2Variable costs:

- Raw materials.
- Miscellaneous operating materials.

- Utilities (Services).
- Shipping and packaging.

Table 5.1 typical factor for estimation of fixed capital cost (Coulson Richard, 2003).

Item	Fluid	Fluid solids	Solids
f1 Equipment erection	0.4	0.45	0.50
f2 Piping	0.70	0.45	0.20
f3 Instrumentation	0.20	0.15	0.10
f4 Electrical	0.10	0.10	0.10
f5 Buildings, process	0.15	0.10	0.05
f6 Utilities	0.50	0.45	0.25
f7 Storages	0.15	0.20	0.25
f8 Site development	0.05	0.05	0.05
f9 Ancillary buildings	0.15	0.20	0.30
f10 Design and Engineering	0.30	0.25	0.20
f11 Contractor’s fee	0.05	0.05	0.05
f12 Contingency	0.10	0.10	0.10

Table 5.2 Summary of production costs (Coulson Richard, 2003).

Variable costs	Typical values
Miscellaneous materials	10 percent of maintains cost
Maintenance	5- 10 per cent of fixed capital
Laboratory costs	20-23 percent of operation labor
Supervision	20 percent of operation labor
Plant overheads	50 percent of operation labor
Capital charges	10 percent of the fixed capital
insurance	1 per cent of the fixed capital
Local taxes	2 percent of the fixed capital
Royalties	1 percent of the fixed capital

5.4 fixed capital cost:-

Table 5.3 equipment purchase price (KRC, 2011)

Equipment	Price $
Compressor	88000 $
Cooler	10000 $
Purifier	9900 $
Heat exchanger	90000 $
Distillation column	110000 $
Total purchase cost of equipment(PCE)	307900 $

f1 Equipment erection = 0.40

f2 Piping = 0.70

f3 Instrumentation = 0.20

f4 Electrical = 0.10

f5 Buildings = none required

f6 Utilities = not applicable

f7 Storages = provided in PCE

f8 Site development = not applicable

f9 Ancillary buildings = none required

Total physical plant cost (PPC) = PCE * (1+ f1+f2 +f3 +f4).

PPC = 307900 * (1 + 0.40 + 0.70 + 0.20 + 0.10) =738960 $

f10 Design and Engineering = 0.30

f11 Contractor's Fee none (unlikely to be used for a small, plant project)

f12 Contingencies = 0.10

Fixed capital = PPC * (1+ f10 + f12).

Fixed capital = 738960 * (1+ 0.30 + 0.10) = 1034544 $

5.5 Working capital:-

Take Working capital allow 5% of fixed capital to cover the cost of the initial catalyst charge = 1034544 * 0.05 = 51727.2 $

Total investment required for project = fixed capital + working capital

Total investment required for project = 1034544 + 51727.2 = 1086271.2 $

5.6 Annual operating costs:

5.6.1 Fixed cost:

- Maintenance, take as 5% of fixed capital = 1034544*0.05 = 51727.2 $
- Operating labor, we have two shifts every shift with 8 persons.
- Four engineers their cost say 28800$ per year = 7200*4 = 28800 $
- Twelve technical's their cost say 43200$ per year = 3600*12 = 43200 $
- Manager, say 12000$ per year = 12000 $
- Plant overheads, take as 50% of operating labor = 72000*0.5 = 36000 $
- Capital charges,6% of fixed capital (bank rate 4%) =1034544*0.06= 62072.64$
- Insurance, 1% of fixed capital = 1034544*0.01 = 10345.44 $
- Local taxes neglect
- Royalties not applicable

Fixed cost = 244145.28 $

5.6.2. Variable costs:

- Raw materials, air from atmosphere = 0 $
- Miscellaneous materials, 10% of maintenance cost =51727.2*0.10 = 5172.72 $
- Utilities, power 1KW= 0.03$.Consummation 814 KW per day.

Power (electricity) = 814*330*0.03 = 8058.6 $

Variable costs = 13231.32 $

Direct production costs = fixed cost + variable cost

Direct production costs = 244145.28 + 13231.32 = 257376.6 $ say = 257377 $**.**

Chapter 6

Process analysis and result

6.Process analysis and results

6.1 Process analysis

After executing and run the programme the following profiles obtained.

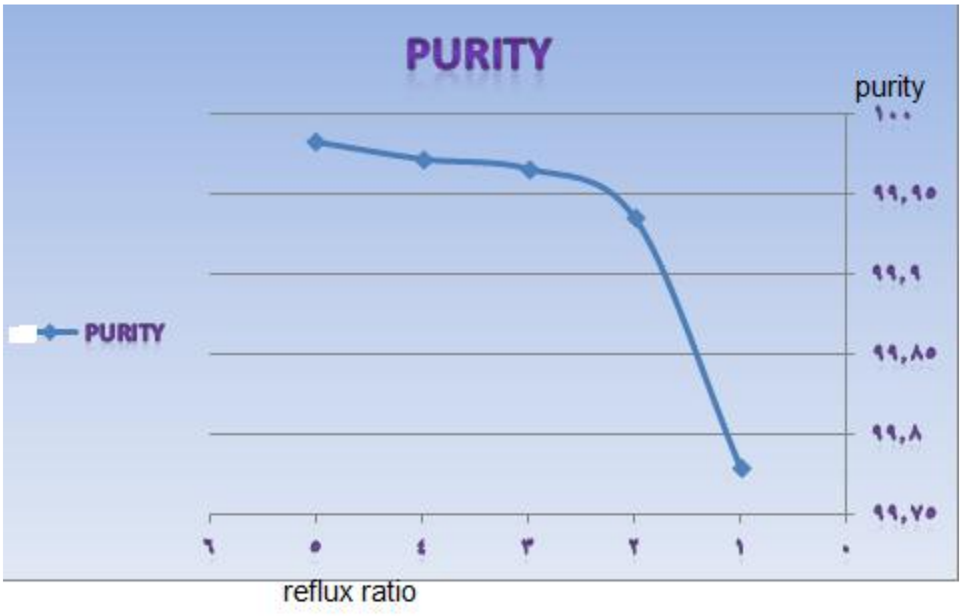


Figure 6.1 reflux ratio VS. Purity

The above figure represents the relationship between the reflux ratio and the purity, as the reflux increase the purity will increase.

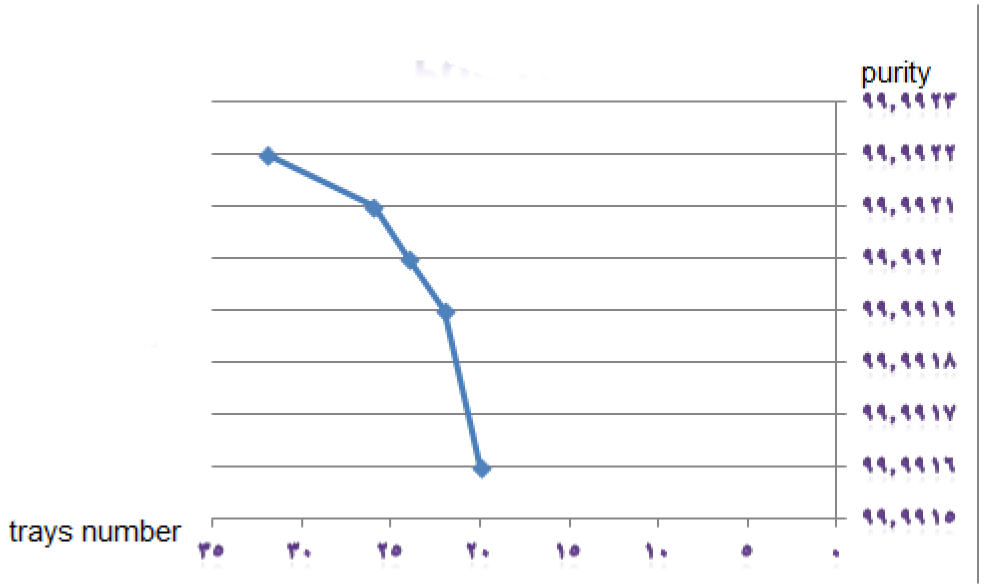


Figure 6.2 number of trays VS. Purity

The figure above represents the relationship between the purity and number of trays. Increasing the number of trays will increase the product purity.

6.2 results:

1) Cryogenic air separation method is most methods use to produce high purity nitrogen.

2) the only method of producing liquid nitrogen is cryogenic air separation.

Table 6.1 flow and purity of products resulted

Product	Flow rate(Ibmole/hr)	Purity
N2 gas	58.99	99.9922
N2 liquid	2.227	99.9999
O2 liquid	36.1	99.9963

Chapter 7

Conclusion and recommendations

7 Conclusion and recommendations

7.1 **conclusions**:

Cryogenic air separation method is most common route due to high nitrogen purity can be produced, it can produce electricity as a by-product when process work by high flow rate or capacity. The final waste out of heat exchanger can de used in regeneration process in the purifier unit. Cryogenic air separation is regarded as the most energy intensive method.

Increasing the reflux ratio in the distillation column will increase the product purity, also increasing the number of trays give high purity products. So that to produce high purity oxygen and nitrogen we have to increase the reflux ratio or the number of trays or both. But economical consideration should be taken in to account.

7.2 Recommendations:

1) Calculate the second low efficiency for further analysis.
2) Air separation is utility unit and use low scale, but if it designed for commercial unit with high scale it can generate electricity from the waste gas.
3) Use ASPEN PLUS programe for the simulation of air separation unit instead of HYSYS because it provide more equipment (e.g. multi heat exchanger).

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