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THESIS

TECHNO-ECONOMIC DESIGN OF INTEGRATED ENEGRY RECOVERY PROCESS FROM LNG RECEIVING TERMINAL

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A Thesis Submitted in Partial Fulfillment of
the Requirements for the Degree of
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The study purposes the integrated system design of waste energy recovery from regasification unit in LNG receiving terminal to utilize this energy for the emission reduction; especially VOCs, in MTP and surrounding areas. The scope also covers the engineering design of pipeline and the route for waste gas, conceptual design of gas terminal and raw drinking water production by hydrocyclone and multi-stage filter.

The result shows that the proposed integrated system can reduce VOCs, sulphur compound and CO_2 at removal efficiency of 100.0, 96.5 and 55.9%; respectively. In addition, the VOCs content in treated gas is lower than the minimum international air pollution standard for acute, intermediate and chronic levels. The treated gas and by-product from the gas treatment system can be utilized as fuel gas for power generation and low quality fuel feedstock. The by-product from LNG regasification can be the raw drinking water with the combined technique of freeze desalination and reverse osmosis.

The economic evaluation shows that this integrated system is feasible in wide range of feed conditions. The system can operate with positive economic whether the waste gas feed available or not. The payback period of normal scenario is 1.45 years comparing to 1.50 years if there is no waste gas feed. In addition, the sensitivity analysis demonstrates that payback period is feasible in 10 years although under the extreme condition of price factors.

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Student's signature Thesis Advisor's signature

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Taradon Piromchart

March 2013

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LIST OF ABBREVIATIONS

ρ_I	=	Density of ice (kg/m^3)
r	=	Radius inside hydrocyclone (m)
F_C	=	Centrifugal force ($kg\ m/s^2$)
R_f	=	Tangential velocity (m/s)
ε	=	Efficiency of hydrocyclone
ρ	=	Density (kg/m^3)
¶	=	Molecular flux of momentum
g	=	Gravitational acceleration (m/s)
$\bar{u}$	=	Time smoothed velocity (m/s)
ư	=	Instantaneous deviation of velocity (m/s)
ν	=	Overall velocity vector (m/s)
F	=	External body force (N)
μ_T	=	Turbulent viscosity (kg/m-s)
k	=	Turbulent kinetic energy (m^2/s^2)
α_s	=	Solid volume fraction
ρ_s	=	Density of solid phase (kg/m^3)
$\bar{u}_s$	=	Velocity of solid phase (m/s)
m_{ls}	=	Un-directional mass transfer between fluid and solid phase
N	=	Total number of phases
$F_{lift,s}$	=	Lift force (N)
$F_{vm,s}$	=	Virtual mass force (N)
K_{sl}	=	Fluid-solid exchange coefficient
μ_l	=	Viscosity of liquid (kg/m-s)
τ_s	=	Particulate relaxation time
e_{ls}	=	Coefficient of restitution
$C_{fr,ls}$	=	Coefficient of friction between the 1^{st} ans s^{th} solid phase particles
d_l	=	Diameter of the particles of solid (m)

LIST OF ABBREVIATIONS (Continued)

P_s	=	Solid pressure ($kg/m\text{-}s^2$)
λ_s	=	Solid bulk viscosity (kg/m-s)
μ_s	=	Solid shear viscosity (kg/m-s)
$G_{0,ss}$	=	Radial distribution
S	=	Distance between grains (m)
$Ø_{ls}$	=	Energy exchange between the 1st fluid or solid phase and the sth solid phase ($kg/m\text{-}s^3$)
$\gamma_{\theta s}$	=	Collisional dissipation of energy ($kg/m\text{-}s^3$)
$k_{\theta s}$	=	Diffusion coefficient (m^2/s)
$\mu_{s,col}$	=	Collisional viscosity (kg/m-s)
$\mu_{s,kin}$	=	Kinetic viscosity (kg/m-s)
$\mu_{s,fr}$	=	Frictional viscosity (kg/m-s)
Ø	=	Angle of internal friction
D_o	=	Overflow diameter (m)
D_u	=	Underflow diameter (m)
AUD	=	Australian dollar
BTEX	=	Benzene, Toluene, Ethyl benzene and Xylene
CAPEX	=	Capital Expenditure
CFD	=	Computational Fluid Dynamics
COS	=	Carbonyl Sulfide
DNS	=	Direct Numeric Simulation
GOT	=	Gulf of Thailand
HHV	=	Higher Heating Value
HSE	=	Health, Safety and Environment
KOD	=	Knock Out Drum
LES	=	Large Eddy Simulation
LCO_2	=	Liquid Carbon Dioxide
LHV	=	Lower Heating Value
LNG	=	Liquid Natural Gas

LIST OF ABBREVIATIONS (Continued)

LPG	=	Liquefied Petroleum Gas
LTS	=	Low Temperature Separator
MMBTU	=	Million British Thermal Unit
MMSCFD	=	Million Standard Cubic Foot per Day
MTP	=	Map Ta Pud Industrial Estate
MTPA	=	Metric Ton Per Annual
NG	=	Natural Gas
OPEX	=	Operating Expenditure
ORV	=	Open Rack Vaporizer
PEA	=	Provincial Electricity Authority
RANs	=	Reynold Average Navier-Stroke
RSM	=	Reynold Stress Model
RNG	=	Re-Normalization Group
SD	=	Staging Drum
STD	=	Standard condition
THB	=	Thai Baht
VOCs	=	Volatile Organic Compounds

TECHNO-ECONOMIC DESIGN OF INTEGRATED ENERGY RECOVERY PROCESS FROM LNG RECEIVING TERMINAL

INTRODUCTION

Currently, many problems occur due to the growth of civilization and industrial on all over the world. Thailand gets the impact result from these reasons both direct and indirect impact. Some problem is incoming to Thailand but in some problem has been happened for a long time without any mitigation plan. However, the solution of one problem may be an indirect solution for other problems such as issue below:

1. Gas emission problem in Thailand

Map Ta Pud (MTP) is the main industrial estate in Thailand where comprise of many diversities of process plant e.g. Petrochemical plant, Refinery Plant, and Power generation plant, etc. There are more than 100 plants in the area (Included refinery) plus one port terminal industry, and there are over 200 combustion stacks for excess gas combustion before emit to atmosphere. For last three decades, many issues impact to Health, Safety and Environment to villagers who live in the surrounding area of Map Ta Pud industrial estate area.

In 2005, Campaign for Alternative Industrial Network, Greenpeace Southeast Asia (2005) performed bucket air sampling by applied the popularization of simplified “Grab sampling” for analysis. Grab sampling is to simulate the lung’s breathing of foul chemical odors and allows for full testing of their chemical makeup. This popular method has been known as the “Bucket Brigade”. The result of analysis in 5 month monitoring found that:

1. Benzene (Carcinogen) detected in four of the five samples exceeded the US EPA Annual Ambient Air Screening Level by as much as 60 times.

2. Vinyl Chloride (Carcinogen) detected in two samples exceeded the EPA Annual Ambient Air Screening Level by as much as 86 times.

3. 1, 2-Dichloroethane (EDC) (Carcinogen) detected in two samples exceeded the EPA Annual Ambient Air Screening Level by as much as 3,378 times.

4. Chloroform (Carcinogen) detected in a sample was exceeded of the EPA Annual Ambient Air Screening Level by 119 times.

5. 20 different kinds of toxic chemicals were identified in the 5 air samples.

6. At least 6 up to 12 Volatile Organic Compounds (VOCs) and sulfur compounds were detected in each sample. At least 2 of the toxic chemicals in each sample were over one or more health protective standards or screening levels, giving proof the toxic cocktail inhaled in MTP.

2. Water scarcity

Water is the primary substance of life on earth, and it is increasingly in short supply. Water shortages affect 88 developing countries that are home to half of the world's population (Miller, 2003). In these places, 80-90% of all diseases and 30% of all deaths result from poor water quality. Furthermore, over the next 25 years, the number of people affected by severe water shortages is expected to increase 4 times. Some of this problem increasing relates to population growth. Some of this problem relates to the demands of industrialization. It directly impacts to the global weather condition. Currently, weather condition has changed. Greenhouse effect is one of phenomena that scientist believes that it is the source of global ambient temperature gradually increasing. Currently, water consumption doubles every 20 years, about twice the rate of population growth.

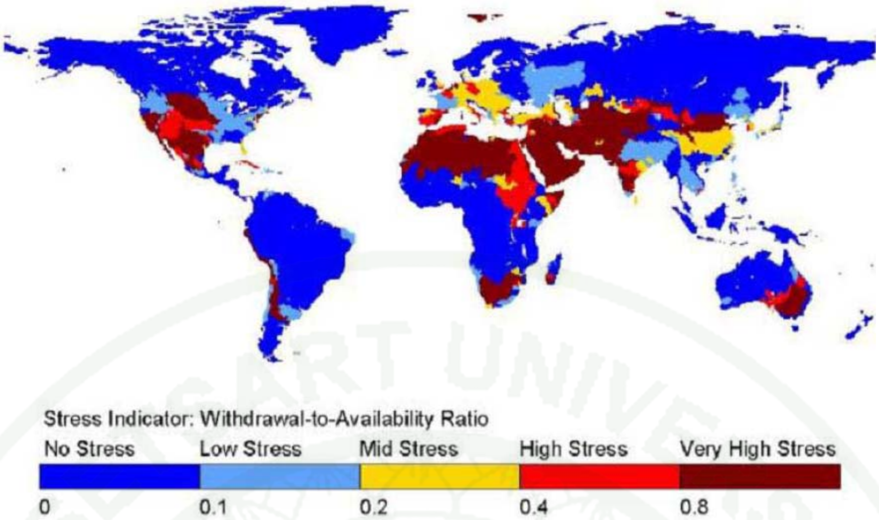


Figure 1 Predicted worldwide water stress index in 2015.

Source: Miller (2003)

Figure 1 shows the global water stress tended to occur in 2015. Therefore, water scarcity is the major world problem, which needs the solution for fresh water sustainability. Water conservation and new low cost technology of seawater desalination and fresh water purification are important on this purpose.

In Thailand, especially the Eastern region, the water scarcity or water stress is the major problem as per historical data in year 2005. There are many factors of water scarcity in Rayong and other area in this region such as the rapid growth of industrialize, the rapid growth of the local population, type of community (e.g. agricultural and residential) and geography, etc. In addition, seasonal and unpredictable of weather is also the major for water scarcity. Therefore, if there is the way to produce raw water by using waste energy of industrial plant, it will reduce the raw water demand from a natural source and sustain the raw water in Thailand, especially in this region. The example of solution method is to utilize the waste energy for raw water or raw drinking water productions. The water production is then to recycle to the industrial plant as industrial feed water or sell the raw drinking water as drinking water feedstock.

3. LNG receiving terminal

Liquefied Natural Gas (LNG) consists primarily of methane. The liquefaction process starts from removal of impurities and some components such as carbon dioxide, mercury, and sulfur. The gas is then cooled to about - 160 ^{o}C to liquid phase at near atmospheric pressure. When natural gas is liquefied, its volume can be reduced up to 600 times, and to economical transportation by ship. From the storage tank, LNG will be re-gasified back into the gas phase prior fed into the natural gas pipeline system.

In Thailand, LNG is planned to import for sustainable the country energy requirement by PTT PCL. Therefore, the LNG receiving terminal and regasification facility was constructed in MTP industrial estate and completed in Q3 2011.

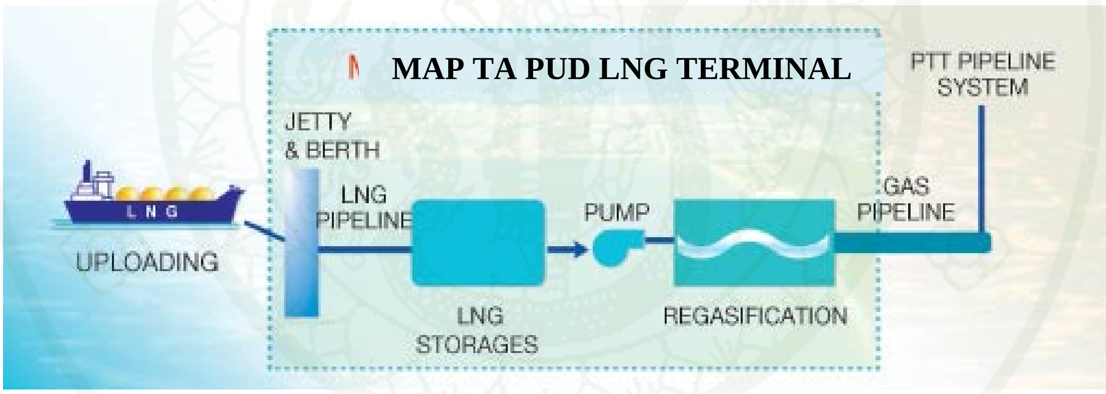


Figure 2 Overview of block diagram of LNG receiving terminal.

Source: Modified PTTLNG Ltd. (2009)

(a) (b)

Figure 3 Topology of LNG terminal project in Map Ta Pud; (a) LNG location, (b) Overview location of LNG terminal in Map Ta Pud

Source: (a) PTTLNG (2009)
(b) Google Earth (n.d.)

By the process of regasification, there is an enormous energy released from the system as per Figures 2 and 3. In general, this excess energy is discarded to the environment without any recovery or utilization. For PTT LNG, the capacity of the re-gasifier (Seawater-based system) is designed at 5 MTPA (Metric Ton per Annual) and planned to expand the plant capacity to 10 MTPA maximum in the future. Therefore, if there is the integrated process, which can utilize this, excess energy, it would give the better solution; especially for gas emission reduction and freshwater production in MTP.

4. Hydrocyclone

Seawater is a salt solution, which consists of inorganic salt and organic salt. The freezing desalination process is alternatively used to purify the seawater based on the principle that the ice crystal can form and exclude the impurities. The remaining liquid is the concentrated brine. After the crystallization process, the ice and brine slurry will be sent to the separation equipment in order to concentrate the ice slurry from the brine slurry. The ice will melt to the pure water. The freezing desalination is not commercially available now due to the difficulty of separation system. The optimal design of separation system for freezing desalination process requires a fundamental understanding of eutectic phase diagram of salt solution, the eutectic freezing crystallization, and hydrocyclone.

In the past three decades, several researchers show the great attempt for studying the separation system for freezing desalination process. The first success in separating the ice from the brine has been achieved by the use of countercurrent wash column, in which a small portion of the product water, flowing in a direction counter to that of the ice motion, to wash the ice (Howe, 1974). However, the wash column gives the high recovery of ice, when the ice crystal from the freezer is large enough to form the porous bed. The alternative separation method of the wash column is the hydrocyclone. Hydrocyclone is effectively used with particles as small as 10 μm while the wash column requires an average particle size of over 300 μm (Dalby et al., 1990). The ice particle size was studied by Margolis (Evan et al., 2008). The study was performed by continuous well-stirred ice crystallizer producing ice with direct contact refrigeration. The effluent of crystallizer was examined to determine the concentration and the characteristics of the ice particles formed. The minimum and the maximum particle diameter investigated from the experiment at -5 °C were 200 and 1,200 μm; respectively. According to the study of ice particle size, the hydrocyclone was suitable instead of a wash column.

Stepakoff (1974) studied the continuous eutectic freezing process based on the technology developed for direct contact freeze crystallization. The salt solution used in the eutectic freezing experiment was potassium chloride, because it did not form a hydrate when it crystallized at low temperature. It was easier to separate and collect the ice crystal at low temperature. The hydrocyclone was applied to separate the lighter solid or ice crystal from the salt crystal.

To design the separation system, the temperature and pressure of the ice slurry must be specified. The eutectic temperature of NaCl solution is - 23 °C. The seawater does not contain only NaCl but also other inorganic salts i.e. $CaCl_2$, $MgSO_4$, and $CaSO_4$. Barduhn and Manudhane (1979) studied eutectic temperature for various ions in natural seawater and found that the temperature of NaCl combined with other ions could lower the eutectic temperature below - 50 °C.

In this work, basic hydrocyclone module of Aspen Plus cannot apply for ice-seawater system. Therefore, another goal is to develop the CFD model of hydrocyclone for seawater desalination process based on the supplemented eutectic freezing crystallization. The work included hydrocyclone dimension adjustment to give the best performance for ice-seawater separation.

OBJECTIVE

The objective is to design a new integrated system for emission reduction in MTP and seawater desalination by utilizing of the waste energy from LNG receiving terminal.

Scopes

1. To design the process to recover waste energy from LNG receiving terminal with the integration of waste gas treatment and seawater desalination process for emission reduction in MTP.

2. To perform pipeline network simulation of feed waste gas pipeline.

3. To design the seawater desalination process for raw drinking water production.

4. To develop the model of hydrocyclone for ice-seawater separation in freezing desalination process by Computational Fluid Dynamic (CFD).

5. To perform cost estimation and basic payback period for whole integrated system.

Expected Results

1. The new integrated system can reduce the emission in MTP to qualified level of international air pollution screening level by the utilizing of excess energy from LNG regasification process.

2. Seawater desalination process is designed into the integrated system. The raw drinking water from this process is the by-product from the whole system.

3. The result of basic economic model shows the feasibility of the new integrated process implementation.

4. The CFD result of a modified hydrocyclone can be applied for ice-seawater separation.

LITERATURE REVIEW

This review section is divided into 4 parts; seawater desalination, hydrocyclone, gas treatment and LNG regasification. Each part describes the basic of these technologies related to the study:

1. Seawater desalination

1.1 Overview of desalination system

The principal of seawater desalination comprise of feed delivery system, pretreatment, desalination, post-treatment as shown in Figure 4.

1.1.1 Pretreatment of feed water to desalination plants

The water is fed to the desalination plant contains dissolved solid, other impurity such as slit, algae, bacteria, and other forms of small plant and animal life. The important group of contaminant is Transparent Ex-polymer Particle (TEP) which is formed from dissolved polymers exuded by phytoplankton and bacteria found in the sea. Therefore, the pretreatment process is required to remove these impurities by filtration.

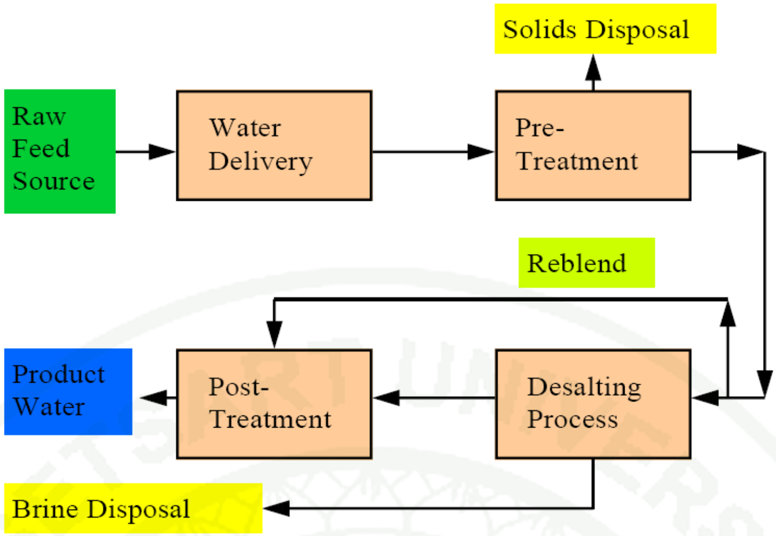


Figure 4 Overall process block diagram of desalination system

Source: Clayton (2006)

1.1.2 Desalination process

Desalination is the process, which reduces the quantity of dissolved substance in a water source. If the salty water content is reduced properly, it can be converted to fresh water suitable for domestic purpose. Natural seawater contains dissolved substance such as sodium chloride, calcium bicarbonate, magnesium sulphate and naturally occurring substance.

In seawater desalination process, there are 4 key techniques which are

a) Thermal technique: According to the thermal process, this technique can be categorized into 2 sub-systems as shown in Table 1.

b) Membrane technique (Anonymous, n.d.): Employ synthetic membrane to filter the major dissolved solid (especially salts). The process normally relates to high pressure condition.

c) Electrodialysis (ED) technique (Anonymous, 1998): Since the salts in the seawater are composed of both positive and negative ions i.e. NaCl ➔ Na^+ + Cl^-. Electrodialysis uses ion–exchange membrane, which is selective to positive and negative ions. Under the influence of direct electric current (DC), the positive sodium ions is designed to pass through a cation membrane and the negative chloride ions is designed to pass through an anion membrane. The incoming saline water is thus converted to concentrated brine and desalinated fresh water.

d) Reverse Osmosis (RO) technique (Miller, 2003): Osmosis is to transfer of water through a semi-permeable membrane from a low concentration to a high concentration solution. The process is reversed if the pressure is applied to the high-concentration side of the membrane, namely water diffuse through the semi-permeable membrane into the low concentration solution.

However, there are the advantage and disadvantage points for each process. The detail of each technique will be described in the next section.

Table 1 Category of sub-system for thermal desalination process

Thermal desalination process	
Heating system	**Freezing system**
Principle: Saline water is heated to generate steam, which is then condensed to produce non-saline water.	**Principle:** Use thermal energy to freeze and then melt the ice formed in the freezing stage, and then separate the ice and brine
Example	**Example**
- Multi-stage flash evaporation distillation (MSF)	- Vacuum freezing
- Vapor compression distillation (VCD)	- Secondary refrigerant freezing
- Solar distillation	- Clathrate or hydrate formation process

Source: Modified Anonymous (n.d.)

(1) Thermal process

a. (a) Multi-Stage Flash evaporation/distillation (MSF): The saline or seawater is heated and evaporated. The pure water is obtained by condensing the vapor. When the water is heated in a vessel, both temperature and pressure of saline will increase. The heated water passes to another chamber or flash vapor at a lower pressure, which causes vapor to be formed. The vapor is led off and condensed into pure water using the cold seawater, which feeds the first heating stage. The concentrated brine is passed to the second flash vapor at a lower pressure to evaporate more water. The process is repeated through a series of the vessel or chambers until atmospheric pressure is reached. Typically, an MSF plat can contain from 4 to 50 stages; approximately. The generic process schematic of MSF is shown in Figure 5.

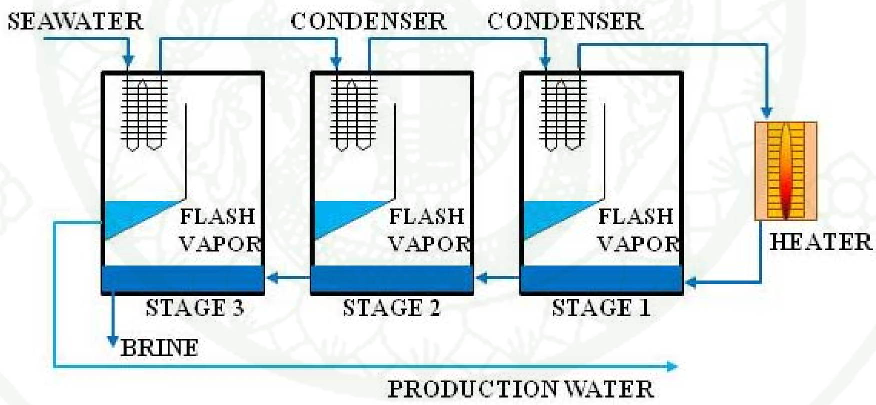


Figure 5 Process schematic of the Multi-Stage Flash evaporation

Source: Modified from Clayton (2006)

Advantage

- Multi-stage flash evaporation is considered to be most reliable and the most widely used because the extensive design and operational experience gained over the last three decades.

Disadvantage

- A typical MSF plant may have 20 to 50 stages. On the other hand, many stages increase the overall efficiency of heat recovery in the plant and decrease MSF operating costs but increase the capital cost of the plant.

- In most recently built MSF plants, 50 to 75 percent of the waste concentrate from the last stage is mixed with the incoming feed water to increase the heat recovery and decrease the amount of water needing pretreatment. However, this also increases the corrosion and scaling (i.e., precipitation of inorganic minerals) in the plant due to the increased salt concentration in the circulating brine.

(b) Vapor Compression Distillation (VCD): Steam is generated from the seawater using a source of heat, and the vapor is then compressed using a compressor. As a result of this compression, the temperature and the pressure of the steam are increased. The incoming seawater is used to cool the compresses steam, which then condenses into distilled fresh water and at the same time the seawater is heated further producing steam. The process schematic of VCD is shown in Figure 6.

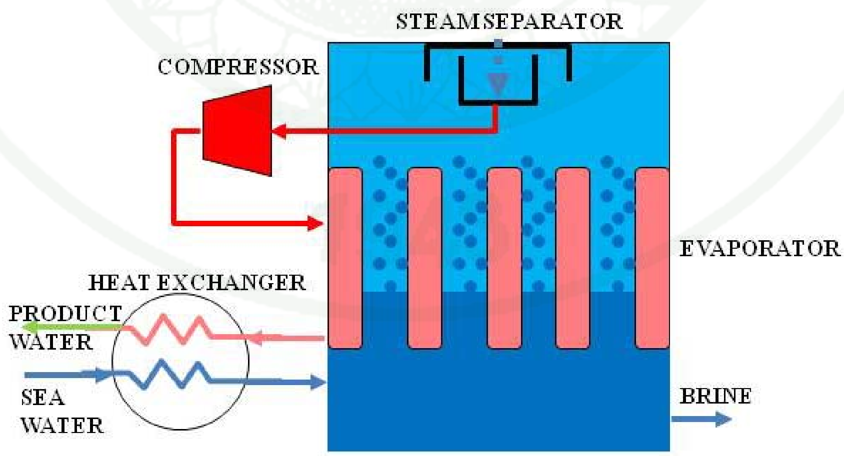


Figure 6 Process schematic of Vapor Compression Distillation (VCD)

Source: Modified from Clayton (2006)

Advantage

- VCD is suitable to apply wherever the requirement for desalinated fresh water is relatively small such as small communities, ships or holiday resorts.

- Since the process is to reduce the pressure in distillation chamber, the boiling of saline water will be reduced, and the energy requirement for distillation is also decreased.

Disadvantage

- Since the main equipment of VCD is vapor compressor, which is a dynamic mechanical equipment, therefore, the possible of equipment broken down is higher than other methods which use the static equipment such as a heater or a boiler.

(c) Solar distillation: Heat from the sun warms the seawater in a glass-covered tank causing some to evaporate. The vapor is condensed on a glass cover, and the resultant fresh water is collected as shown in the diagram in, but is not suitable for the large-scale production water. Problems can arise from the growth of algae. The process schematic of solar distillation is shown in Figure 7.

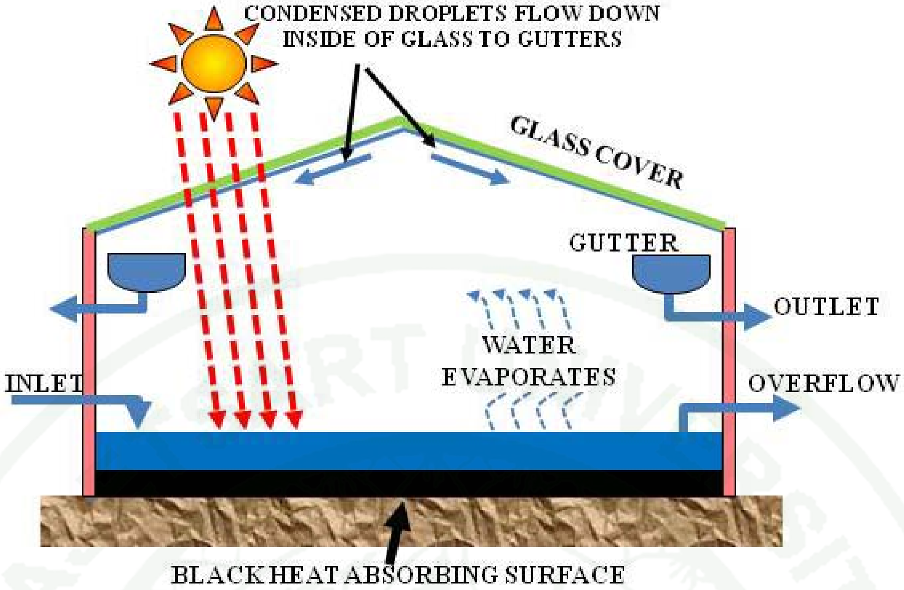


Figure 7 Process schematic of Solar distillation

Source: Modified from Clayton (2006)

Advantage

- Solar distillation is a low-cost system.
- Development of this system in Australia is reliable and effective.

Disadvantage

- Solar distillation is not suitable for large-scale production water
- Growth of algae can occur on the underside of the glass cover.
- Good sealing is required otherwise; the vapor and heat can escape and limit the effectiveness of the system.

(d) Freeze desalination: When salty water freezes, the ice crystallizes from pure water leaving the dissolved organic and inorganic solids (e.g. salt) in liquid pockets of high salinity brine. Traditional freezing processes involve five steps: precooking of the feed water, crystallization of ice into slurry, separation of ice from the brine, washing the ice, and melting the ice. Although freshwater can be obtained quite easily from ice where seawater freezes naturally, the engineering involved in constructing and operating a freeze desalination plant is quite complicated. Freeze desalination has the potential to concentrate a wider variety of wastes streams to higher concentrations with less energy than any of the distillation process. In fact, the energy requirements for freezing and reverse osmosis are comparable. Pretreatment of incoming feed water is not necessary, and not a critical corrosion problem with this freezing approach. The process schematic of freeze desalination is shown in Figure 8.

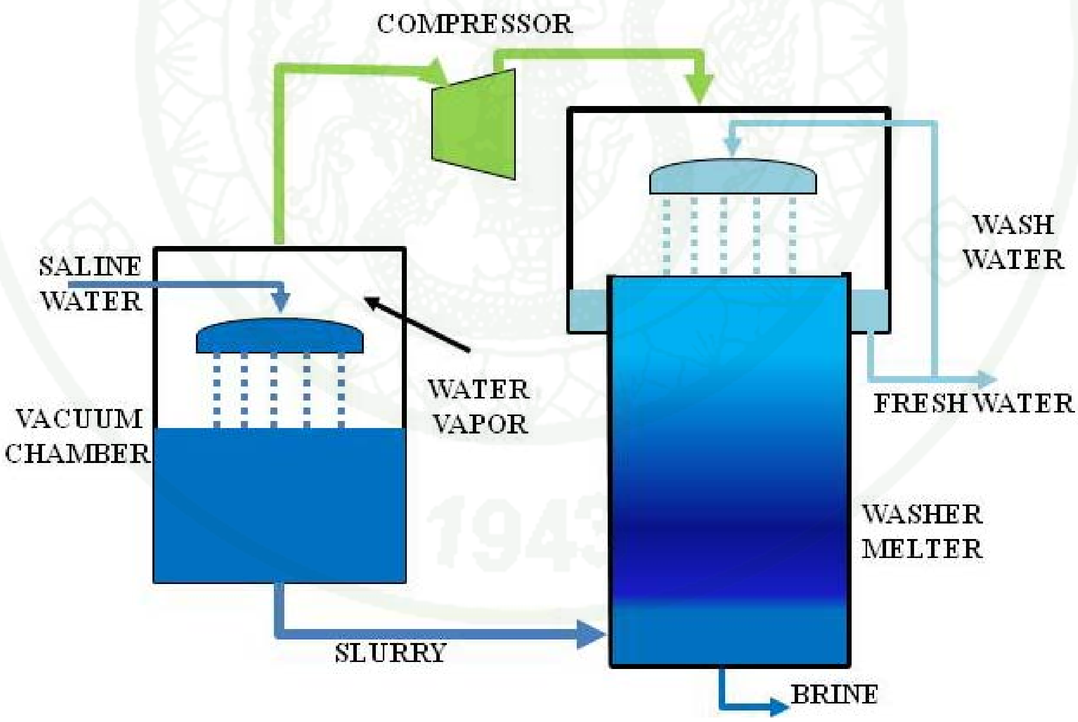


Figure 8 Process schematic of Freeze desalination

Source: Modified from Clayton (2006)

Advantage

- Concentrate a wider variety of wastes streams to higher concentrations with less energy than any of the distillation process.

- Pretreatment of incoming feed water is not necessary, and corrosion is much less of a problem with freezing due to the low operating temperatures.

- Some equipment development required for freeze desalination has occurred over the last 30 years.

Disadvantage

- Some potential involves spraying seawater, or contaminated freshwater, into the air when winter temperatures fall below 29 °F for significant periods of time. This variation on freeze desalination can only be used in colder winter climates.

- Costs are likely to run about \$1.50 to \$3.00 per 1,000 gallons, even for small scale applications (i.e. less than 1 mgd).

1.1.3 Post treatment of the desalinated water

Thermal desalination processes produce water with a very low dissolved content; in effect, they produce distilled water. The water from the desalinated process taste insipid, so if the water is used for drinking purpose, so if the water is used for drinking purpose, some dissolved solids and air must be added back to the desalinated water. Disinfection is required before the water can be used. The additional form of post treatments might be necessary such as re-aeration, corrosion control, and disinfection.

1.2 Eutectic curve of water/brine

The phase diagram of the binary system; NaCl-H_2O is shown in Figure 9. The concentrations of salt have been maintained below 23.3 %wt, marked by the horizontal line, in order to avoid a saturated solution. Salt water can be considered a different material from pure water.

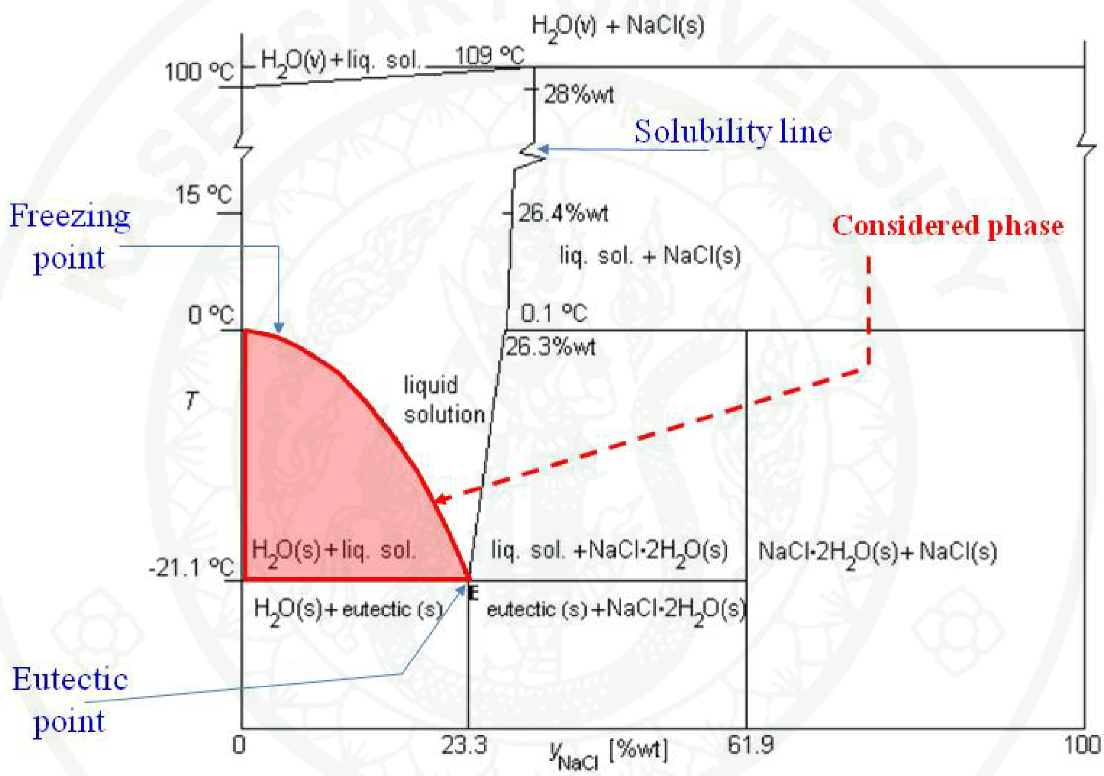


Figure 9 Eutectic curve of NaCl-H_2O system at atmospheric pressure

Source: Martinez (n.d.)

The concentrations and conductivities of the solutions have been plotted in graph shown below in Figure 10.

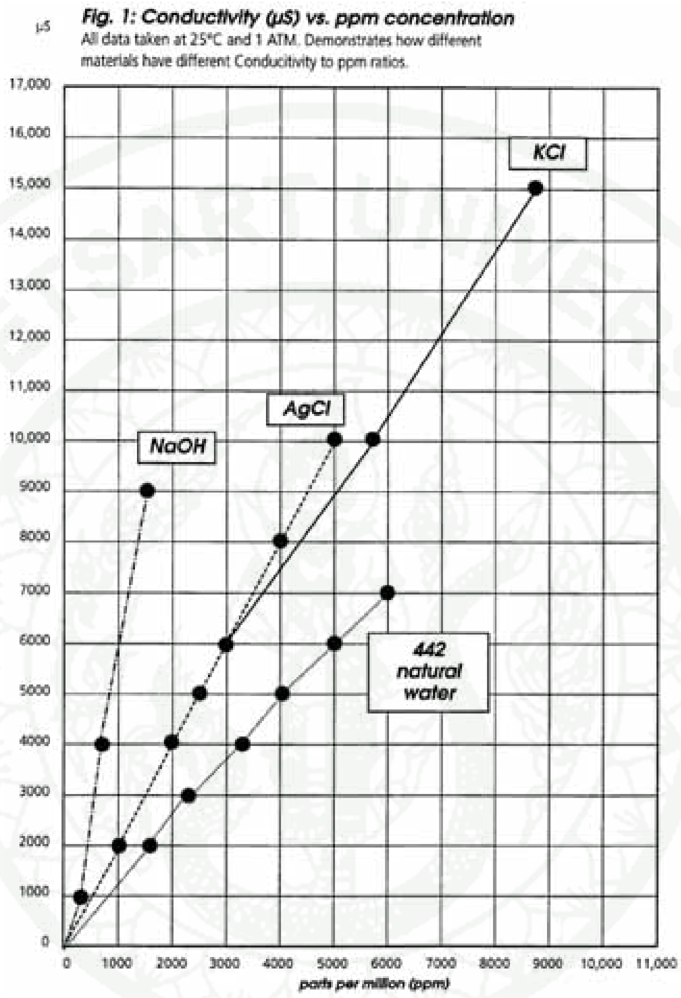


Figure 10 Conductivity versus concentration relation of NaCl

Source: Eutech instruments (n.d.)

2. Hydrocyclone

2.1 Overview of Hydrocyclone

Hydrocyclones are used for solid-liquid separations; as well as for solids classification, and liquid-liquid separation. It is a centrifugal device with a stationary wall, the centrifugal force being generated by the liquid motion. Hydrocyclone is simple, robust, separating devices, which can be used over the particle size range from 4 - 500 µm.

2.2 Principle of hydrocylcone

Liquid cyclones can be used for the classification of solid particles over a size range from 5 to 100 µm. Commercial units are available in a wide range of materials of construction and sizes; from as small as 10 mm up to 30 m diameter. The hydrocyclone separates solid and liquid or liquid and liquid by the difference in density between the fluid and material. Due to the fluid acquires a spiraling motion caused by the tangent feeding, the material of larger density is thrown against the wall of the hydrocyclone and dragged to the underflow while the one of smaller density proceeds for the overflow, forming a free vortex (outer vortex) and a forced vortex (inner vortex) in agreement with Figure 11.

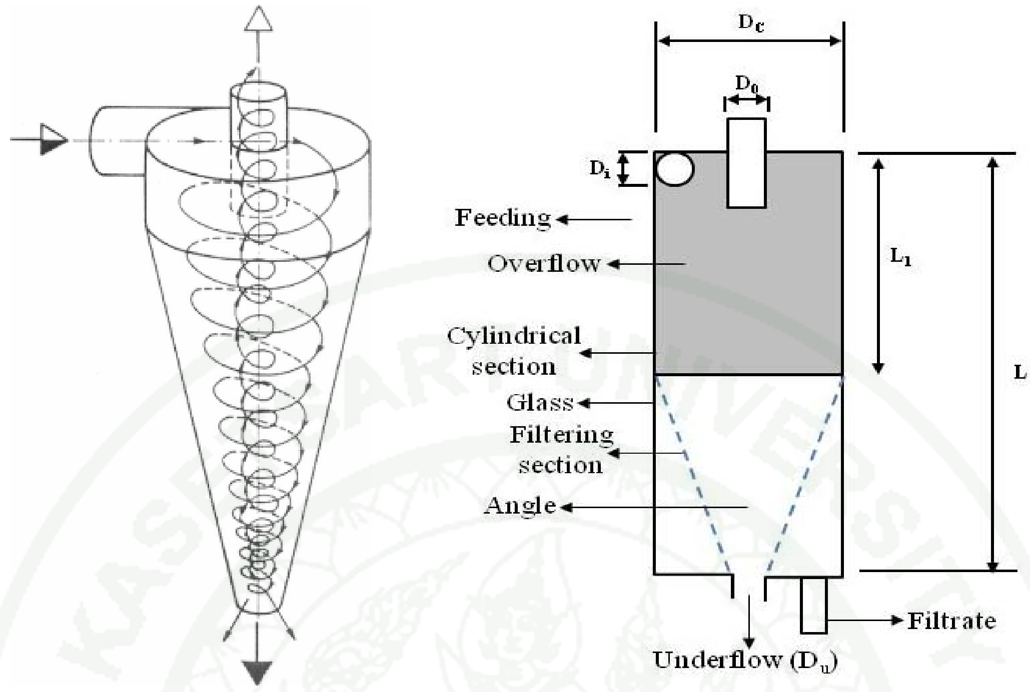


Figure 11 Flow pattern and scheme diagram of hydrocyclone

Source: Modified Viera et al. (2005)

2.3 Hydrocyclone design

2.3.1 Inlet design

Hydrocyclones designed prior to 1950 featured outer wall tangential feed entry and 12-15 mm thick rubber liners. This design was not adequate for fine separations or for abrasive slurry applications. Most hydrocyclone manufacturers have redesigned their inlets to include some form of involutes, ramped or scrolled feed style. They found that all of these designs provide a measured advantage in hydrocyclone performance compared to the earlier tangential designs. Figure 12 illustrates the various types of hydrocyclone feed entries. The inlet opening or cross-sectional area of the orifice feeding into the cylindrical section of the inlet has an effect on capacity as well as D_{50}, and most hydrocyclone models have several options to increase or decrease this area based on the desired flow rates and cut point.

In general, the larger inlet area, the higher the hydrocyclone capacity and the larger the predicted D_{50}.

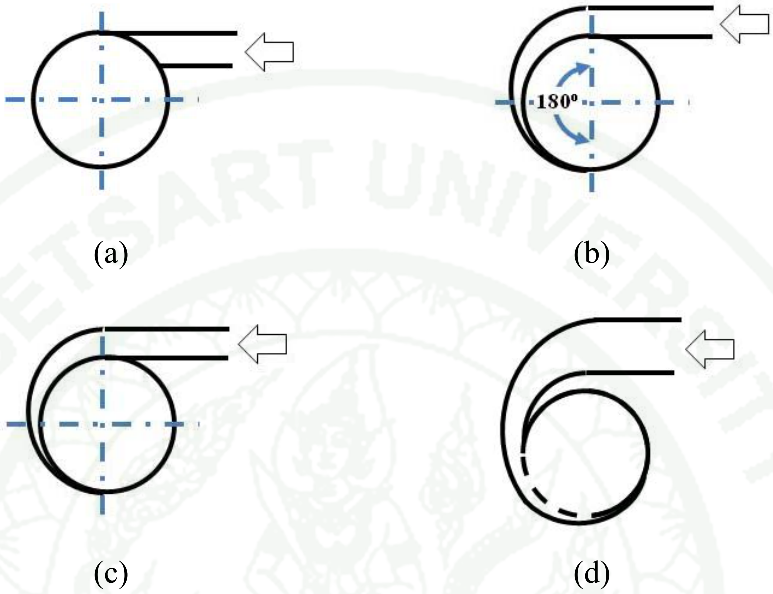


Figure 12 Hydrocyclone inlet pattern; (a) Outer wall tangential, (b) Involute, (c) Ramped entry or Scrolled envolute, (d) Involute ramp

Source: Modified Arterburn (n.d.)

2.3.2 Cylinder Section

Typically, hydrocyclone has a cylinder section length equal to the hydrocyclone diameter. This can be a separate section or integral to the inlet head. While the longer cylinder section provided greater residence time and thus more capacity, it also reduces the tangential velocity. The added cylinder length results in the minimal improvement in hydrocyclone separation and will increase hydrocyclone capacity at the same pressure by 8-10%. Larger 660-840 mm diameter hydrocyclone typically have shorter cylinder sections.

2.3.3 Mounting Angle

Hydrocyclones can be mounted at angles ranging from vertical to nearly horizontal. The effect of lowering the mounting angle will increase the expected D_{50} for a given hydrocyclone by 20-40% depending on the angle. This has been a popular tool to increase concentrator tonnage by effectively producing a coarser product. However, the configuration of mounting at angles less than 45^o (based horizontal) give a result in some maintenance problems, most notably with inlet head wear life (Figure 13).

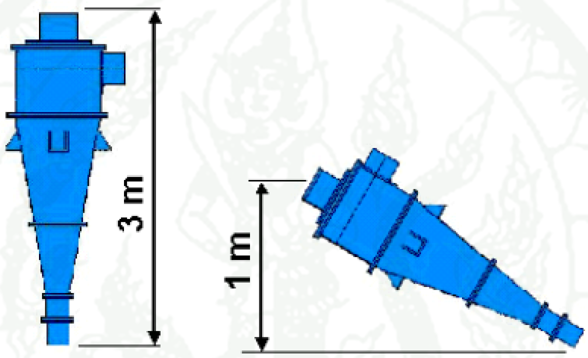


Figure 13 Example of mounting angle configuration vertical and angled

Source: Modified from Arterburn (n.d.)

Most large mill circuits have 660 – 840 mm diameter hydrocyclone. These large diameter hydrocyclone are normally 2,500 to 3,000 mm tall. These tall hydrocyclone provide a substantial head on the underflow discharge of the hydrocyclone. In order to maintain a high underflow density, the apex size must be closely monitored and maintained.

Hydrocyclone installation configuration at 45^o greatly reduces the head on the underflow discharge. A consistently high underflow density can be achieved because apex diameter is not as critical as in vertical installations. In addition, the lower head results in reduced velocity of the slurry spraying out of the hydrocyclone. This increases the component life of the apex by about 100% compared to vertically mounted hydrocyclone.

2.4 Design criteria for hydrocyclone

The separating efficiency of liquid cyclones depends on the particle size and density, and the density and viscosity of the liquid medium.

2.4.1 Modeling theory: Residence Time Theory

There are several theories for the describing the separation that takes place in hydrocyclone. One of the best known is the residence time theory. In this approach, a particle of a given size will be collected if the time it remains inside the cyclone is equal to or greater than the time necessary for it to reach the wall. The monographs by Zanker can be used to make a preliminary estimate of the size of cyclone needed.

$$\frac{d_{50}}{D_c} = K\left[\frac{\mu D_c}{Q(\rho_s - \rho)}\right]^{0.5} \tag{1}$$

Where

d_{50} =	Particle cut size
D_C =	Diameter of the cylindrical section of the cyclone
μ =	Liquid viscosity
Q =	Volumetric feed flow rate
K =	A parameter characteristic of each cyclone design

It is important to highlight that the cut size obtained during operation of the hydrocyclone results not only from the action of the centrifugal field but also from the solid material carried by the downward stream fluid. In order to allow estimation of the reduced cut size, many authors suggested additional factors to account for the underflow-to-throughput ratio effect.

$$\frac{d'_{50}}{D_c}=K\left[\frac{\mu D_c}{Q(\rho_s-\rho)}\right]F(R_L)G(C_v) \quad (2)$$

Where

$$F(R_L)=\frac{1}{1+1.73R_L}$$

$$G(C_V)=e^{(4.5C_V)}$$

R_L is the underflow – to - throughput ratio and C_v is the volumetric feed concentration. Equation (3) is suitable for prediction of the underflow-to-throughput ratio:

$$R_L=B\left(\frac{D_U}{D_C}\right)^C \quad (3)$$

Where

D_U the underflow diameter and B and C are constants for a given cyclone design.

2.4.2 Multiphase flow modeling

A large number of flows encountered in nature and technology are a mixture of phases. Physical phases of matter are gas, liquid, and solid, but the concept of phase in a multiphase flow system is applied in a broader sense. Multiphase flow regimes can be grouped into four categories: gas-liquid or liquid-liquid flows; gas-solid flows; liquid-solid flows; and three-phase flows.

a) Liquid-Solid flow regime

The following regimes are liquid-solid flows:

(1) Slurry flow: This flow is the transport of particles in liquids. The fundamental behavior of liquid-solid flows varies with the properties of the solid particles relative to those of the liquid. In slurry flows, the Stokes number is normally less than 1. When the Stokes number is larger than 1, the characteristic of the flow is liquid-solid fluidization.

(2) Hydrotransport: This describes the distribution of solid particles in a continuous liquid.

(3) Sedimentation: This describes a tall column initially containing a uniform dispersed mixture of particles. At the bottom, the particles will slow down and form a sludge layer. At the top, a clear interface will appear and in the middle a constant settling zone will exist.

2.4.3 Fluid motion

The fluid motion is very significant in the modeling and simulation impact factor. For fluid motion, it is reasonable to make a calculation based on locally averaged quantities.

Equation of continuity;

$$\frac{\partial \varepsilon}{\partial t}+\nabla\cdot(\varepsilon u)=0 \qquad (4)$$

Equation of momentum transfer in x, y and z direction;

$$
\begin{aligned}
&\frac{\partial}{\partial t}(\varepsilon u)+\frac{\partial}{\partial x}(\varepsilon uu)+\frac{\partial}{\partial y}(\varepsilon vu)+\frac{\partial}{\partial z}(\varepsilon wu)=-\frac{\varepsilon}{\rho}\frac{\partial p}{\partial x}+\frac{K_x}{\rho}\\
&\frac{\partial}{\partial t}(\varepsilon v)+\frac{\partial}{\partial x}(\varepsilon uv)+\frac{\partial}{\partial y}(\varepsilon vv)+\frac{\partial}{\partial z}(\varepsilon wv)=-\frac{\varepsilon}{\rho}\frac{\partial p}{\partial y}+\frac{K_y}{\rho}\\
&\frac{\partial}{\partial t}(\varepsilon w)+\frac{\partial}{\partial x}(\varepsilon uw)+\frac{\partial}{\partial y}(\varepsilon vw)+\frac{\partial}{\partial z}(\varepsilon ww)=-\frac{\varepsilon}{\rho}\frac{\partial p}{\partial z}+\frac{K_z}{\rho}
\end{aligned}
\tag{5}
$$

Equation of motion;

$$\frac{\partial(\rho_f \varepsilon u)}{\partial t}+\nabla\cdot(\rho_f \varepsilon u u)=-\varepsilon\nabla p-F \tag{6}$$

$$F=\frac{\sum_{i=1}^{n} f_{pi}}{\Delta V} \tag{7}$$

Where

ρ_f = Fluid density

P = Fluid pressure

u = Fluid velocity

F = Volumetric fluid-particle interaction force

ε = Porosity or void fraction (the ratio of void volume to the volume of the cell

ΔV = Volume of a computational cell

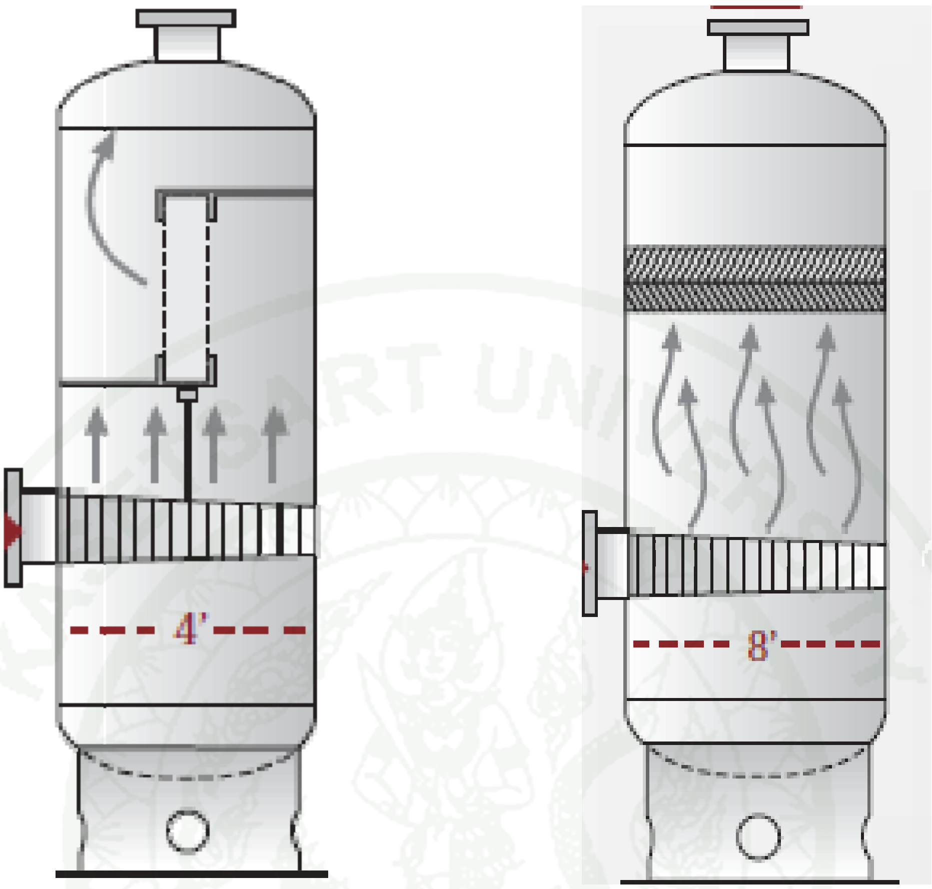
n = Number of particles in the cell

3. Gas treatment

In general, there are several gas treatment techniques such as solid bed adsorption, liquid absorption and reaction, etc., however, in this study, the low temperature approach is considered. There are 2 main approaches, low temperature gas-liquid separator and cryogenic distillation.

Low temperature gas-liquid separation is applied from the concept of general gas-liquid separator but it uses the advantage of low energy source; such as refrigeration system, to reduce the temperature of gas. At low temperature, the high boiling point gas would change from vapor to liquid phase. Therefore, it increases the liquid phase in feed condition to the separator. The inlet and outlet device of the Low Temperature Separator LTS is likely to G-L separator. Since the function is similar. The different between the general and LTS is the material. In general, if gas is sweetening gas. The material of the separator is carbon steel; however, the minimum design temperature of carbon steel is -29 ^{o}C. Hence, if the operating temperature of LTS is lower than -29 ^{o}C or have a potential to operate under -29 ^{o}C. The type of material would be changed to others such as aluminum. Figure 14 shows the diagram of gas-liquid separator with inlet and outlet device.

Cryogenic Distillation (Anonymous, n.d.) is similar to the conventional distillation but it operates at extremely low temperature to separate components according to their different boiling temperatures. Cryogenic separation is widely used for purification e.g. Ethane and CO_2. This method is advantageous for pure component stream in high pressure, which would be liquefied. Water has to be removed before the cryogenic cooling step to avoid blockage from freezing.



(a) (b)

Figure 14 Flow diagram of G-L separator with inlet and outlet device (a) Small vessel using double pocket vane technology (b) Vessel with standard mesh-vane combination

Source: AMISTCO (n.d)

4. LNG regasification

In LNG regasification, the major part is the LNG vaporizer. The purpose of the vaporizer is to heat up the LNG until changing cold NG. There are many kinds of heating media such as air, seawater and hot gas. It depends on the technology. The descriptions of commonly used-type of vaporizer are following (Yang et al., 2004):

4.1 Open Rack Vaporizers (ORV)

ORV use ambient seawater as their source of heat in an open, falling film type arrangement to vaporize LNG passing through the tubes. After heat exchanged, cold seawater is sent back to the sea. In general, for using ORVs the preferred seawater temperature is always above 8 °C. This type of vaporizer is widely used and well proven in baseload LNG terminal service. In Thailand, ORV is applied for LNG receiving terminal in MTP (PTTLNG Ltd., 2009). The schematic of ORV is shown in Figure 15.

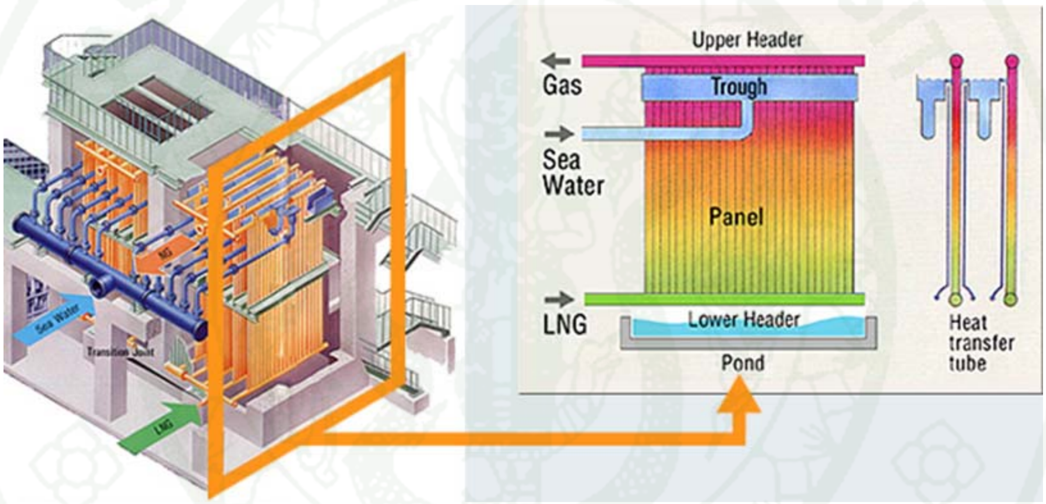
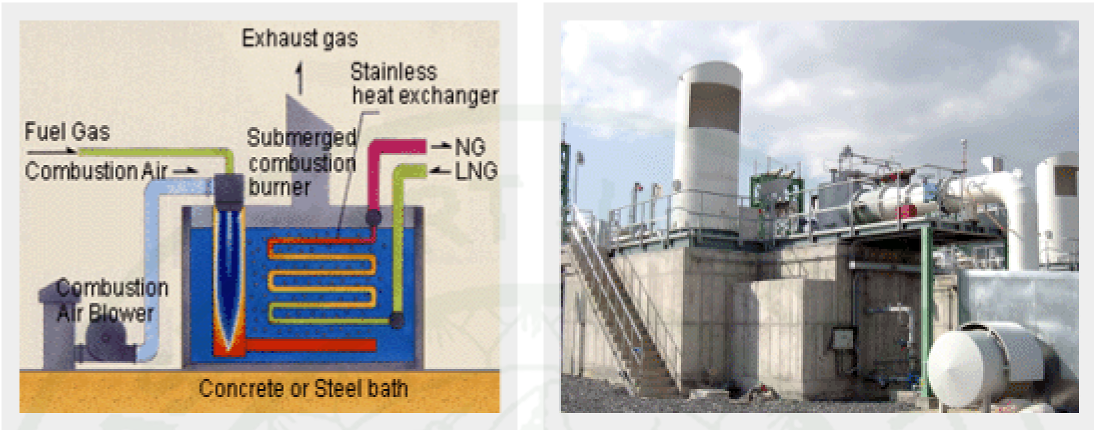


Figure 15 Internal view and fluid flow direction of Open Rack Vaporizer (ORV)

Source: Tokyo gas company (n.d.)

4.2 Submerged Combustion Vaporizers (SCV)

The SCV vaporizes LNG contained inside stainless steel tubes in a submerged water bath with a combustion burner. In the baseload terminal SCV, the fuel gas is burned in a large single burner rather than multiple smaller burners because it is more economical and it achieves low NO_x and CO levels. The hot flue gases are sparged into a bath of water where the LNG vaporization coils are located. However, the large amount of flue gas and the combustion gas contented such as NO_x and CO_2 is a concern issue in operation. The schematic of SCV is shown in Figure 16.



(a) (b)

Figure 16 The overview picture of Submerged Combustion Vaporizer (SCV); (a) Schematic diagram of SCV, (b) the picture of SCV

Source: Wonil T&I Co. Ltd. (n.d.)

4.3 Shell and Tube type Vaporizers (STV)

In general, the STV is smaller in size and cost competitive compared to an ORV or SCV system. Heat is usually supplied to the LNG vaporizer by a closed circuit with a suitable heat transfer medium. They are mainly used when a suitable heat source is available. Design of these types of vaporizer systems requires a stable LNG flow at design and turndown conditions with provisions to prevent the potential for freeze-up within the vaporizer. These have had only limited application to date. The schematic of STV is shown in Figure 17.

Figure 17 The overview of Shell and Tube Vaporizer (STV)

Source: Anonymous (n.d.)

4.4 Ambient Air-heated Vaporizers (AAV)

In general, the principle of AAV is similar to conventional cryogenic vaporizer. Ambient air; either in a natural draft mode or a forced draft mode, is fed to vaporize. The heat from air is exchanged to LNG via external of finned tube. In general, the material of AAV is aluminum. However, this technique is suitable for the warm location because the ambient temperature is high and it is convenient to utilize the current heat energy to vaporize the LNG. The example picture of AAV is shown in Figure 18.

Figure 18 The picture of Ambient Air Vaporizer (AAV)

Source: Anonymous (n.d.)

MATERIALS AND METHODS

Materials

1. Software for simulation

a. ASPEN Plus v.7.2 (Licensed to Chemical engineering department)

b. ASPEN Flare System Analyzer v.7.2 (Licensed to Chemical engineering department)

c. ANSYS Fluent v.13 (Licensed to Faculty of Engineering)

d. ANSYS ICEM CFD v.13 (Licensed to Faculty of Engineering)

2. Support digital database

a. Google Earth v.6.1.0.5001

b. Estimate equipment cost from local vendor

3. Map and plan of industrial plant location in Map-Ta-Pud

4. High performance computer, Dell Precision T5500, as detail below:

a. Dual Intel® Xeon® E5620 2.4 GHz, 12M cache, 5.86 GT/s QPI, Turbo, Hyper-Tread, 4 Cores

b. 12GB (6x2GB) DDR3 RDIMM, 1,333 MHz, ECC

c. 512 MB PCIe x16 NVIDIA Quadro FX580

d. 1TB 7200 rpm 3.5” 6.0 Gb/s SATA hard drive with SATA controller information.

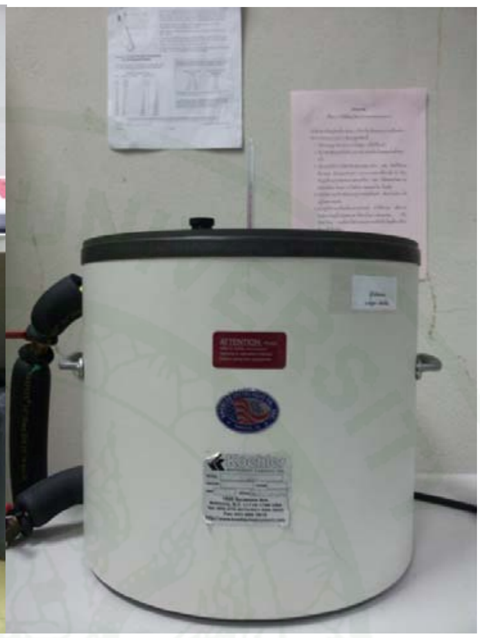
5. Freezing point test equipment (Figure 19)

a. Cooling chamber tester: Koehler, K46001

b. Refrigerated and heating circulator: Julabo, F25-ED

c. Test jar (Total height = 150 mm, ID = 31 mm, Thickness = 2 mm, filling level = 50 mm from bottom)

(a)



(b)

(c)

(d)

Figure 19 Freezing point test equipment: (a) Overview system between cooling chamber and cooling circulator, (b) Cooling chamber, (c) Top view of the cooling chamber with 4 holding ports, (d) Test Jar.

Methods

In the study, the research method is divided in 5 parts with total 19 steps as following:

Main parts of this study are:

1. Literature survey and data gathering including relevance laboratory data.
2. Laboratory testing.
3. Conceptual process design and equipment sizing.
4. Process and equipment design simulation.
5. Preliminary cost estimation and economic evaluation

According to these main parts, they can be described step-by-step as in the following:

1. To define the problem statement.

2. To perform literature survey for technology involving to LNG regasification.

3. To survey the possible and potential technology of utilize energy from LNG receiving terminal process.

4. To survey the process of waste treatment in disposal water from industrial plant.

5. To select the possible technology for utilization of LNG waste energy e.g. Freeze desalination process and low temperature reaction treatment.

6. To study and conclude the gas emission in Map-Ta-Pud industrial estate area.

7. To analyze the impact of emission to Health, Safety and Environment (HSE) of community in the surrounding area.

8. To acquire the information of LNG receiving terminal process in Map-Ta-Pud industrial estate.

9. To conclude the information relate to all properties, characteristic of contaminant in waste gas and related component for integrated system.

10. To perform laboratory test on freezing point of actual seawater from Gulf of Thailand.

11. To design and propose the overview process block diagram of integrated process for waste energy utilization.

12. To generate the simulation models for a new integrated process for waste energy utilization.

13. To generate simulation models for gas emission reduction process with waste energy from LNG receiving terminal.

14. To validate the proposed model under various scenario such as:

14.1 LNG operation capacity to LNG receiving terminal

14.2 The toxic level in waste gas stream

14.3 Waste gas flow rate base on normal condition and emergency condition

14.4 Operating temperature of freeze desalination unit

15. To size the equipment and pipeline in waste gas pipeline network and integrated system.

16. To perform CFD simulation of hydrocyclone for ice-seawater separation

16.1 Geometry generation

16.2 Mesh independent test

16.3 Transport equation model selection and set up.

16.4 Sensitivity analysis.

16.5 Geometry adjustment and validation for separation efficiency improvement.

17. To validate the process base on energy consumption of an integrated process by process designs adjustment.

18. To perform preliminary costing and economic evaluation of the overall process.

19. Analysis and conclusion.

The explanations in detail of research methodology in this study are shown in the following sections.

Literature survey and data gathering

According to the purpose of the research study, excess energy from LNG regasification process and the emission gas from flare stack of each plant in MTP area are focused. The information related to the process design is

1. The flow rate and gas composition of waste gas (excess gas before flaring).

2. The distance of waste gas source to waste energy source.

3. The number of waste gas source and waste energy source.

4. The toxic level in air pollution in the atmosphere surrounding MTP area.

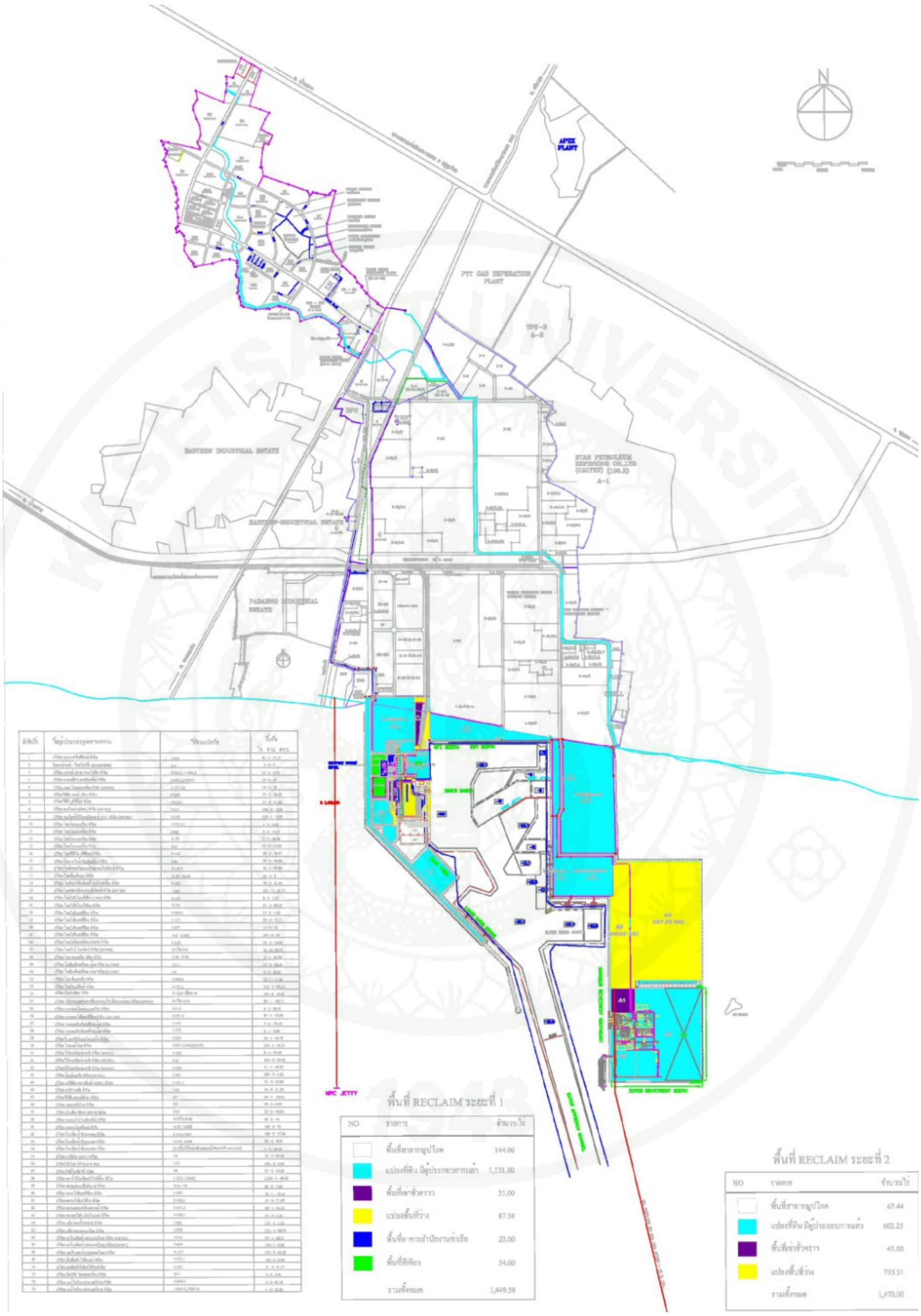


Figure 20 Map Ta Pud industrial Estate Map

Source: Map Ta Pud Industrial Estate (2011)

Waste gas composition is retrieved from Greenpeace (Greenpeace, 2005). It showed that there were toxic components, cancer predominate components and volatile hydrocarbon components, in the sampling of air pollution measurement as per the area of MTP in Figure 20. The air analysis results were compared to the international air pollution screening level standard as shown in Table 2. The value of each type of standard is different. In some composition such as 1,2 Dichloroethane, benzene, etc., EPA region 6 is the most stringent value but in some composition is not such as carbonyl sulfide. The result also shows in negative side. The measurement value of many compositions exceed than the screening level, however, it depends on the standard, which is applied. In the study, the target of gas treatment performance is justified by the amount of these compositions in the treated gas. These compositions in treated gas are expected to get less than the least value from all standards in Table 2.

Table 2 Standard of air pollution screening for VOC and other substance.

Composition	Screening Levels and Standard (ug/m3)(a)							Measurement value (ug/m3)
	(1)	(2.1)	(2.2)	(3.1)	(3.2)	(3.3)	(4)	
1,2-Dichloroethane	0.074	160	4	-	-	718	3.8	250
MEK	5,200	3,900	390	-	-	-	-	8.3
MIK	3,100	2,050	205	-	-	-	-	20
Acetone	370	5,900	590	31,100	15,500	15,500	-	37
Benzene	0.25	75	3	59.8	4.8	-	12	15
Carbon Disulfide	730	30	3	-	-	359	-	14
Carbonyl Sulfide	-	8	0.8	-	-	-	-	21.2
Chloroethane	2.3	500	50	17,900	-	-	-	9.8
Chloroform	0.084	100	10	119	59.8	23.9	4.30	10
Ethylbenzene	1,100	2,000	200	-	1,190	-	-	21
Hydrgen Sulfide	2.1	1	-	239	23.9	-	-	15.5
MTBE	3.7	450	45	2,390	838	838	-	18
Ethylene Chloride	4.1	260	26	718	359	359	212.8	8.5
P&M-xylene	-	2,080	208	-	-	-	-	66
O-xylene	730	-	-	-	-	-	-	24
Styrene	1,100	110	11	-	-	71.8	-	11
Tetrachloroethene	0.33	340	34	239	-	47.9	105.3	6.8
Toluene	400	1,880	188	1,190	-	95.8	-	190
Trichloroethene	0.017	1,350	135	2,390	119	-	58.8	11
Vinyl Chloride	0.22	130	13	599	35.9	-	1.2	19

Table 2 (Continued)

Note: [(a)]

(1) means EPA Region 6 Screening Level
(2.1) means Short-term Texas Effects Screening Levels
(2.2) means Long-term Texas Effects Screening Levels
(3.1) means Acute-ATSDR Minimal Risk Levels
(3.2) means Intermediate-ATSDR Minimal Risk Levels
(3.3) means Chronic-ATSDR Minimal Risk Levels
(4) means Louisiana Toxic Air Pollution Ambient Air Standards

Source: Greenpeace (2005)

It is assumed that the waste gas is not completed combustion at flare and leads to have toxic component remained in disposal stream from flare and it also emit directly to MTP atmosphere. The components and their content level in waste gas (Feed local flare stack) have not been reported because this public information did not reveal any specific components in waste gas. Therefore, in the study, this information is assumed as components contented in waste gas of each plant.

However, the toxic content level in the report based on ground level sampling point, their contents are lower than originated level in waste gas stream and the content of most hydrocarbons (such as methane) and H_2, etc are not shown because they are combusted at flare already. Therefore, two scenarios for the study; High toxic case (HiT) and Low toxic case (LoT), are required with additional assumptions following:

1. The toxic content level of LoT case bases on the value in the publication report. For HiT case, the value of toxic content is assumed to be 1% for all toxic gases and also more than the low case approximately 1-10 times (Depending on the composition).

2. The real waste gas composition should be located in between of the range of HiT and LoT.

3. The other basic component in waste gas (except toxic gas) such as hydrocarbon (methane, ethane, etc), H_2 and inert gas are assumed as components in waste gas stream

With assumption above, the composition of each case is summarized in Tables 3 and 4.

Table 3 Waste gas composition (Dry condition), the quantity level and their chemical property in high toxic case (HiT) and low toxic case (LoT).

Component name	Formula	Molecular Weight	Mole fraction		Scenario Differential ratio
			HiT	LoT	
METHANE	CH_4	16.04	0.25	0.2529	0.988
ETHANE	C_2H_6	30.07	0.15	0.1518	0.988
PROPANE	C_3H_8	44.10	0.06	0.0607	0.988
ISOBUTANE	C_4H_{10}	58.12	0.02	0.0202	0.988
N-BUTANE	C_4H_{10}	58.12	0.01	0.0101	0.988
2-METHYL-BUTANE (i-C_5)	C_5H_{12}	72.15	0.01	0.0101	0.988
N-PENTANE	C_5H_{12}	72.15	0.01	0.0101	0.988
N-HEXANE	C_6H_{14}	86.18	0.01	0.0101	0.988
HYDROGEN	H_2	2.02	0.10	0.1012	0.988
HYDROGEN-SULFIDE	H_2S	34.08	0.01	0.0103	0.968
NITROGEN	N_2	28.01	0.02	0.0202	0.988
CARBON-DIOXIDE	CO_2	44.01	0.10	0.1012	0.988
CARBON-MONOXIDE	CO	28.01	0.05	0.0506	0.988
BENZENE	C_6H_6	78.11	0.01	0.0044	2.293
TOLUENE	C_7H_8	92.14	0.01	0.0468	0.214
P-XYLENE	C_8H_{10}	106.17	0.01	0.0071	1.417
M-XYLENE	C_8H_{10}	106.17	0.01	0.0071	1.417
O-XYLENE	C_8H_{10}	106.17	0.01	0.0051	1.948
VINYL-CHLORIDE	C_2H_3Cl	62.49	0.01	0.0069	1.448
1,2-DICHLOROETHANE	$C_2H_4Cl_2$	98.96	0.01	0.0574	0.174
CHLOROFORM	$CHCl_3$	119.38	0.01	0.0019	5.256
CARBONYL-SULFIDE	COS	60.07	0.01	0.0080	1.248
METHYL-TERT-BUTYL-ETHER (MTBE)	$C_5H_{12}O$	88.15	0.01	0.0046	2.156
METHYL-ETHYL-KETONE (MEK)	C_4H_8O	72.11	0.01	0.0026	3.825
METHYL-ISOBUTYL-KETONE (MIK)	$C_6H_{12}O$	100.16	0.01	0.0045	2.205
ACETONE	C_3H_6O	58.08	0.01	0.0145	0.691
CARBON-DISULFIDE	CS_2	76.13	0.01	0.0042	2.394
ETHYL-CHLORIDE	C_2H_5Cl	64.51	0.01	0.0030	3.342

Table 3 (Continued)

Component name	Formula	Molecular Weight	Mole fraction		Scenario Differential ratio
			HiT	LoT	
ETHYLBENZENE	C_8H_{10}	106.17	0.01	0.0045	2.226
DICHLOROMETHANE	CH_2Cl_2	84.93	0.01	0.0026	3.816
STYRENE	C_8H_8	104.15	0.01	0.0024	4.169
TETRACHLORO-ETHYLENE	C_2Cl_4	165.83	0.01	0.0009	10.738
TRICHLOROETHYLENE	C_2HCl_3	131.39	0.01	0.0019	5.259
WATER	H_2O	18.02	0.00	0.0000	0.000

Table 4 Boiling point of each gas component in waste gas at different condition.

Component	Formula	Component type	Boiling point (^oC)		
			NBP	13 barg	14 barg
STYRENE	C_8H_8	HC	145.16	283.20	287.40
O-XYLENE	C_8H_{10}	HC	144.42	280.10	284.60
M-XYLENE	C_8H_{10}	HC	139.11	272.80	277.50
P-XYLENE	C_8H_{10}	HC	138.36	272.80	277.40
ETHYLBENZENE	C_8H_{10}	HC	136.2	270.30	275.20
TETRACHLORO-ETHYLENE	C_2Cl_4	Non-HC	121.25	252.70	255.90
METHYL-ISOBUTYL-KETONE (MIK)	$C_6H_{12}O$	Non-HC	116.45	241.80	243.50
TOLUENE	C_7H_8	HC	110.649	237.60	242.40
WATER	H_2O	H_2O	100.00	195.30	198.60
TRICHLOROETHYLENE	C_2HCl_3	Non-HC	87.25	206.30	209.40
1,2-DICHLOROETHANE	$C_2H_4Cl_2$	Non-HC	83.55	198.80	204.20
BENZENE	C_6H_6	HC	80.09	198.40	202.70
METHYL-ETHYL-KETONE (MEK)	C_4H_8O	Non-HC	79.55	191.20	195.40
N-HEXANE	C_6H_{14}	HC	68.73	184.60	188.40
CHLOROFORM	$CHCl_3$	Non-HC	61.15	171.90	176.10
ACETONE	C_3H_6O	Non-HC	56.05	160.20	163.70
METHYL-TERT-BUTYL-ETHER (MTBE)	$C_5H_{12}O$	HC	55.15	166.10	170.80
CARBON-DISULFIDE	CS_2	Sulphur	46.21	155.30	158.50
DICHLOROMETHANE	CH_2Cl_2	Non-HC	39.85	142.90	146.20
N-PENTANE	C_5H_{12}	HC	36.06	142.40	146.30
2-METHYL-BUTANE (i-C_5)	C_5H_{12}	HC	27.88	133.10	137.70
ETHYL-CHLORIDE	C_2H_5Cl	Non-HC	12.35	108.20	111.10
N-BUTANE	C_4H_{10}	HC	-0.50	95.30	98.97
ISOBUTANE	C_4H_{10}	HC	-11.73	81.73	85.42
VINYL-CHLORIDE	C_2H_3Cl	Non-HC	-13.35	76.20	80.21
PROPANE	C_3H_8	HC	-42.10	40.71	43.89
CARBONYL-SULFIDE	COS	Sulphur	-50.15	28.56	33.11

Table 4 (Continued)

Component	Formula	Component type	Boiling point (°C)		
			NBP	13 barg	14 barg
HYDROGEN-SULFIDE	H_2S	Sulphur	-59.65	10.81	12.62
CARBON-DIOXIDE	CO_2	LCO_2	-78.55	-30.59	-27.68
ETHANE	C_2H_6	Gas	-88.60	-20.28	-18.17
METHANE	CH_4	Gas	-161.52	-116.30	-114.90
CARBON-MONOXIDE	CO	Gas	-191.45	-158.60	-156.70
NITROGEN	N_2	Gas	-195.8	-164.00	-162.80
HYDROGEN	H_2	Gas	-252.59	-239.50	-239.00

The number of waste gas source depends on the selection criteria as below:

1. The possible gas source must be located in the circle area closed to air pollution measurement sampling point.

2. The location must be located in the boundary of MTP Industrial Estate. The gas source, which located in neighbor industrial estate such Pa-Dang Industrial Estate, Hemmaraj East Industrial Estate, Eastern Industrial Estate and RIL Industrial Estate are excluded because the distance of pipeline network is quite long and the route cannot be specified clearly compared to MTP area. In addition, this study is to demonstrate the possibility of the integrated system. If the result is worth to do in the further step, source selection must be re-considered for the detail engineering stage.

3. The identification and coordination of the gas source location is implemented from Google Earth software.

4. The location bases on the existing flare stacks located in that industrial plant.

Figure 21 showed that six gas sources qualified under the selection criteria. By this information, the route of waste gas pipeline network is generated followed to the existing gas pipeline network showed in Google Earth software.

Figure 21 The location of waste gas source (Red circle) and air pollution sampling point (Light blue flag).

For the waste gas flow rate, it is calculated with assumptions:

1. The target gas sources operate continuously without excess waste gas for flaring, therefore, only purge gas is burnt at flare stack.

2. The flow rate of waste gas to flare for all plant bases on the following equations:

$$\text{Purge gas flow} = 0.5K \times D^3 \times MW^{-0.565} \quad (8)$$

$$\text{Purge gas flow} = K \times D^3 \times MW^{-0.565} \quad (9)$$

Where

Purge gas flow rate in Sm^3/h

K = Flow factor (-)

D = Stack diameter (in)

MW = Purge gas molecular weight

Source: PTTEP PLC (2010)

3. K factor is the flow factor depending on the stack configuration. In this study, the stack configuration is assumed the similar configuration for all plants. In addition, the K value is calibrated to an actual operating plant value. This K factor will be used in related calculation.

The calculation of waste gas flow rate from each plant is deployed from the simple equations. In these equations, the purge gas is a function of stack diameter, gas molecular weight and flow factor. If there is air ingress protector equipped; e.g. “Molecular seal”, with flare stack (8), the requirement of purge gas would be less than the “without air ingress protector” case (9). In this study, Equation (9) is used, although the most of flare design of MTP plants is equipped with air ingress protector because the flow rate of this equation is higher than Equation (8). It also leads to get the result in case of flow limitation.

Table 5 Waste gas flow rate estimation options (a) Molecular seal is not equipped in the local flare design (b) Molecular seal is not equipped in the local flare design

Gas source	MW (kg/kmol)	Flare tip diameter (in)	Estimated Purge rate (MMSCFD)			
			With Molecular seal		Without Molecular seal	
			Normal	Emergency	Normal	Emergency
Flare#1	33.1615	32	0.1433	0.61	0.2867	1.22
Flare#2	33.1615	36	0.1612	0.87	0.3225	1.74
Flare#3	33.1615	38	0.1702	1.02	0.3404	2.04
Flare#4	33.1615	48	0.2150	2.06	0.4300	4.12
Flare#5	33.1615	54	0.2419	2.93	0.4837	5.87
Flare#6	33.1615	60	0.2687	4.02	0.5375	8.05

In view of process design, the technique is limited to the low temperature energy from LNG regasification; therefore, the most of process design must be specified with the low temperature or cryogenic technology. In the design Open Rack Vaporizer (ORV) is used as regasification facility as shown in Figure 15. Moreover, the process of LNG receiving terminal is the one of important issues for design the integrated process for energy recovery. Therefore, the information of LNG process design should include:

1. LNG receiving terminal process flow diagram (Figure 22).

2. LNG regasification technique and their related equipments; especially for existing LNG regasification in MTP.

3. Design capacity of the MTP LNG receiving terminal.

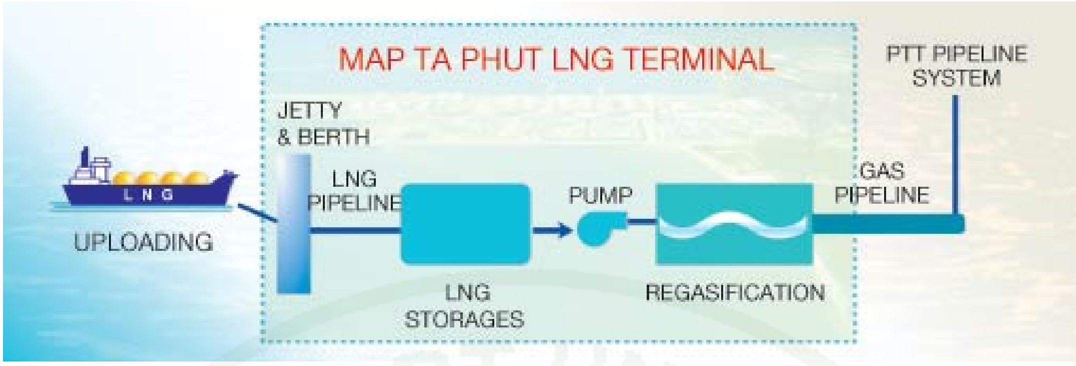


Figure 22 Process block diagram of LNG receiving terminal in MTP.

Source: PTTLNG Ltd. (2009)

After gathering all preliminary information as described above, the conceptual design processes are drafted. The process can be designed in several ways such as for emission gas treatment, wastewater treatment, gas utilization, etc. However, in any design option, the practical design is formulated the simulation from the results.

Simultaneous process design and validation of relevant technology play a major role to accomplish the final design. The result must be qualified under design criteria as following:

1. Compile to the objective of emission reduction and waste energy utilization issue.
2. Minimization of impact on LNG receiving terminal.
3. Possibility of modification on current design or constructibility.
4. Availability of current technology (Including pilot scale technology) applied to the design.
5. Preliminary process condition and related equipment material.
6. Economic evaluation.

Meanwhile, another critical part is the waste gas transportation from each source to waste energy treatment. Since one of design criteria is the minimization of the impact to LNG receiving terminal, the best recovery design must be located inside or nearby LNG receiving terminal to avoid the impact of LNG receiving terminal modification. By this reason, the gas pipeline network is designed to be at transportation method in this study and information of gas pipeline is required as below:

1. The location of waste gas source in MTP area.

2. The distance between each waste gas source to LNG receiving terminal included the possible pipeline route direction.

3. The minimum estimation of pipeline connection for the pipeline route.

Consequently, after the frame of the process design has been performed, the detailed step is including input data and simulation is performed. In this study, seawater is used as a feed for freeze desalination technique, therefore, the important parameters; salt composition, salt content and freezing point, of seawater must be acquired.

Table 6 The radical composition of seawater

Radical	**Seawater**	
	mg/l	**%**
Na^+	11,208	30.37
K^+	394	1.07
Ca^+	416	1.13
Mg^+	1,310	3.56
Cl^-	20,565	55.90
SO_4^-	2,764	7.51
CO_3^{2-}	14	0.04
HCO_3^-	117	0.32
Total	36,788	99.90

Source: AWWA Journal (1988)

The salt composition is retrieved from particular seawater analysis of the sample in the South China between peninsular and Eastern Malaysia. This data is also used for the design of the water maker system in oil and gas production platform in Gulf of Thailand. After the formulation of the salt radical to salt content, the result is shown in Table 7.

Table 7 The composition of seawater and its quantity in simulated seawater source for simulation.

Case	**1**	**2**	**3**	**4**
Reference case	**7[(a)]**	**8**	**10**	**12**
Component	**Composition (Mass fraction)**			
NaCl	0.0277	0.0078	0.0783	0.1565
KCl	0.0006	0.0002	0.0016	0.0033
$KHCO_3$	0.0002	0.0001	0.0005	0.0011
$CaCl_2$	0.0012	0.0003	0.0033	0.0065
$CaCO_3$	0.0000	0.0000	0.0001	0.0001
$MgCl_2$	0.0022	0.0006	0.0063	0.0125
$MgSO_4$	0.0035	0.0010	0.0100	0.0199
H_2O	0.9647	0.9900	0.9000	0.8000

Note: (a) mean this run is a base-case sample in the study.

In the study, there are 4 data sets for using as eutectic testing. The data set#1 is the reference data for GOT. The water content is 96.47 % wt. Data set#2 - #4 are calculated data based on salt- water ratio from data set#1 and the salt compositions are calculated based on a percentage of each salt component in total salt. The objective of water ratio variation is to find the freezing point of seawater and also to confirm the eutectic temperature. These parameters are the simulation results by Aspen Plus based on standard library electrolyte database in the software. In addition, seawater samples from GOT are also tested for freezing point measurement to confirm the result from simulation.

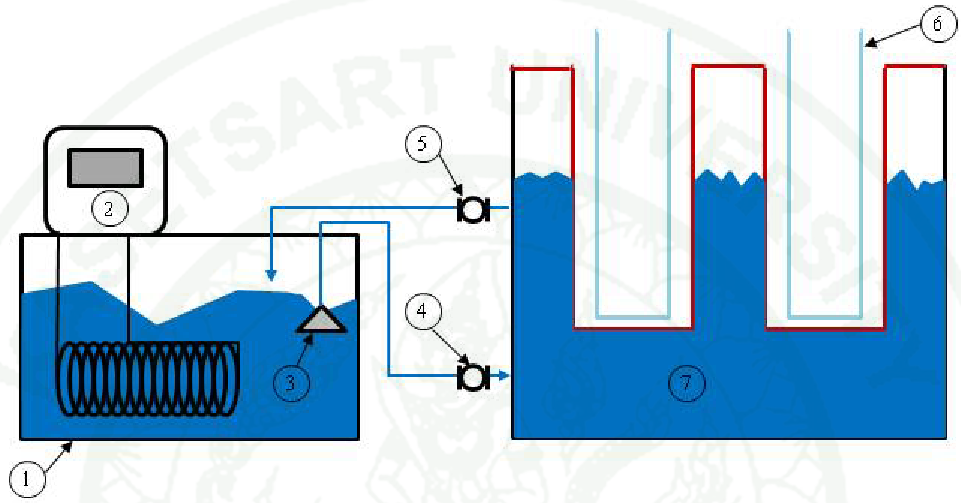
In addition, the effect of seawater supply requirement due the change of salinity is included in the scope of study because the salinity or salt concentration in feed seawater is likely to impact on the heat removal of LNG vaporizer. In the study, 3 cases from Table 7 (Case#1, #2 and #3) are selected and considered for the seawater requirement in freeze desalination process. The compositions of seawater are shown in Table 8.

Table 8 The composition of salt content at each seawater case.

SWT composition	**Mass fraction**		
	Low salt	**Base case**	**High salt**
NaCl	0.0078274	0.0276595	0.0782740
KCl	0.0001645	0.0005814	0.0016452
$KHCO_3$	0.0000529	0.0001869	0.0005290
$CaCl_2$	0.0003269	0.0011552	0.0032691
$CaCO_3$	0.0000067	0.0000238	0.0000672
$MgCl_2$	0.0006258	0.0022114	0.0062581
$MgSO_4$	0.0009957	0.0035186	0.0099574
H_2O	0.9900000	0.9646633	0.9000000

Laboratory test

This part is to find the preliminary information of freezing point of the seawater for the simulation. The flow diagram of test apparatus is shown in Figure 23.



No.	Description
1	Cooling bath
2	Temperature control
3	Circulation pump
4	Inlet valve
5	Outlet valve
6	Sample tube
7	Cooling chamber

Figure 23 Flow diagram of freezing point test equipment.

The petroleum cloud & pour point test apparatus is applied for the freezing point test as shown in Figure 23. There are two main parts; cooling chamber and cooling circulator. The cooling chamber is a compact four-place portable chamber for the immersion of copper test jackets in suitable freezing mixtures. It is designed to allow inlet and outlet connect together for circulating refrigerated coolant from an external source. The chamber consists of steel exterior housing with polyurethane enamel finished, and there is also copper interior for corrosion resistance. Removable composition top plate and 1/2" cork insulation around interior aid in cold retention.

Cooling circulator is design to cooling the media, which is ethanol down to -28 ^{o}C. The refrigerant system based on R134a and there is the heating coil to control the temperature as per set point.

Three samples in GOT are tested. The seawater sample is filled in the test jar until reach the design filling level. Before putting filled test jar in the holding port of cooling chamber, the temperature in the cooling chamber must be set at 1 ^{o}C and stabilize temperature 30 minute. After putting the test jar in the holding port, set point will change to -20 by 1 ^{o}C step. The freezing point is detected at the point that seawater begins to change phase from liquid to solid. The test conducts for 3 times for result confirmation.

Conceptual process design and equipment sizing

In the process design, overview of process block diagram must be generated and consider with the design criteria in the process. However, it is not possible to complete the final process in a single step because there are many unknown factors and results related to the designed such as technologies and/or facilities availability and energy consumption, etc. Therefore, the iterations of the design must be performed as per the working step flow diagram below:

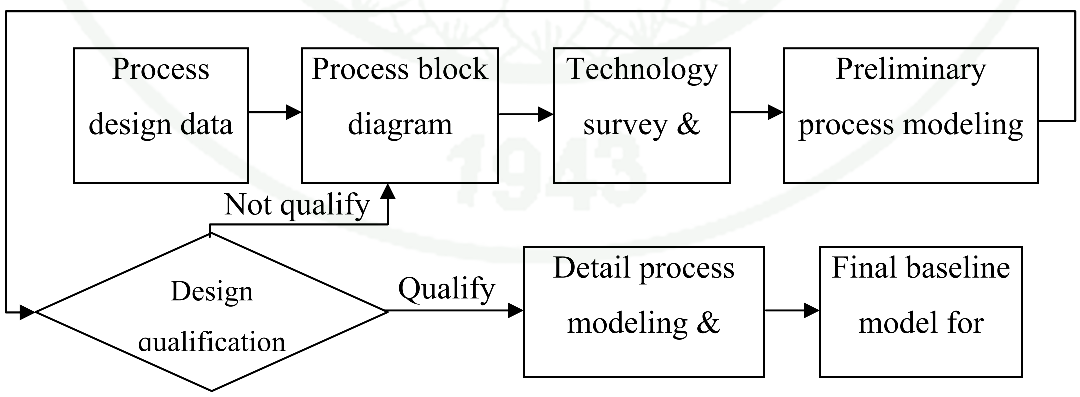


Figure 24 General flow diagram for process and equipment design

Part of process design data acquisition included the operating condition of the process both LNG regasification process and integrated process. The operating condition is classified by the processing plant as below:

1. Waste gas treatment plant

1.1 Waste gas supply pressure

In general, the pressure of flare system of each plant depends on the facility design and many others factor. However, the range of operation of flare header is 0.5-3.5 barg; approximately. In some case, this flare gas has to be routed to waste gas network and delivered to gas treatment plant, the pressure supply at 0.5 barg should be the minimum value for the waste gas treatment plant. If there is the source plant operated at the higher pressure, it would not impact to the design because the inlet of gas treatment plant pressure is designed to handle the waste gas in the range of 0.01 – 0.5 barg. In addition, it also has own flare system to prevent over pressure of gas inlet.

1.2 Waste gas flow rate scenario

Refer to the design of purge gas to flare, 2 scenarios has to be considered; normal operation scenario and emergency operation scenario. Since, in the study, only purge gas is focused as the main supply waste gas source, therefore, the maximum of the purge must be concerned; carefully. The flow rate of purging gas for both scenarios can be calculated by equations (9).

2. LNG Receiving terminal

2.1 LNG gas feed capacity

Refer to the information in publication website (PTTLNG Ltd., 2009), the design capacity of LNG terminal plant are divided in 2 phases of construction, 5 MTPA and additional 5 MTPA (Total capacity of phase 1 and 2 is 10 MTPA). The plant capacity limitation is LNG vaporizer. In the study, the process design considers only the first phase because the firmed development plan for additional phase has not been announced, officially. However, the study also considers the scenario of turndown capacity of the LNG vaporizer. The vaporizer is assumed to have turndown capacity at 60% or 3 MTPA. Therefore, the result of the study applies 3 MTPA LNG as a minimum scenario for LNG terminal plant capacity.

2.2 LNG purity

In general, the purity of LNG product is more than 99% of methane content; however, there are many source of commercial supplier in the world or even for the contracted supplier to MTP LNG terminal. Therefore, in the study, the purity of LNG is set at 100% of methane content. In addition, the deviation of boiling point of natural gas between 99 and 100% is less than 1 ^{o}C, it is reasonable to assume the purity of product to be 100% methane content.

3. Seawater desalination plant

3.1 Operating temperature

Refer to the freeze desalination of eutectic solution; such as seawater, the range of operating temperature is quite wide because it depends on the contaminant content level in the desalination product. In this study, the objective of freeze desalination is to produce the desalinated water as per raw drinking water quality. Therefore, the limitation of the design is to balance between raw drinking water product salinity and operating temperature by using the crystallization temperature of salt contained in seawater as key criteria.

3.2 Disposal water temperature

With the process of seawater desalination, the rejected water stream as salt concentrated stream is generally produced and disposed to the seawater source. However, for freeze desalination, the temperature of this rejected water stream is extremely low compared to other type of seawater desalinations e.g. membrane and multi-effect distillation. Therefore, the impact of different temperature between disposal sink (sea) and the rejected water stream to marine life is considered in the study.

3.3 Disposal water salinity

The salinity of the rejected water is another issue as well as disposal water. However, there is no official report on this issue, in the study, the salinity of rejected water was considered, but it will not be the constraint for the process design. Moreover, the design of rejected water disposal is designed to locate at depth 50 m on the seabed, which is the layer lower than the surface layer and thermocline layer in Gulf of Thailand (Anonymous, n.d) as shown in Figure 25. By this approach, it would reduce the impact of the salinity different to marine life because mostly marine life is located at the surface layer or mixing zone approximated 0 – 30 m. depth. At this layer, the salinity of water is about 35.8 – 37.0 ppt. Thermocline layer is located in the range of 30 – 40 m. depth. The salinity of thermocline layer is 36.0 - 36.8 ppt, approximately. Therefore, the deepening of disposal point (depth more than 40 m) would reduce the salinity different in the sea.

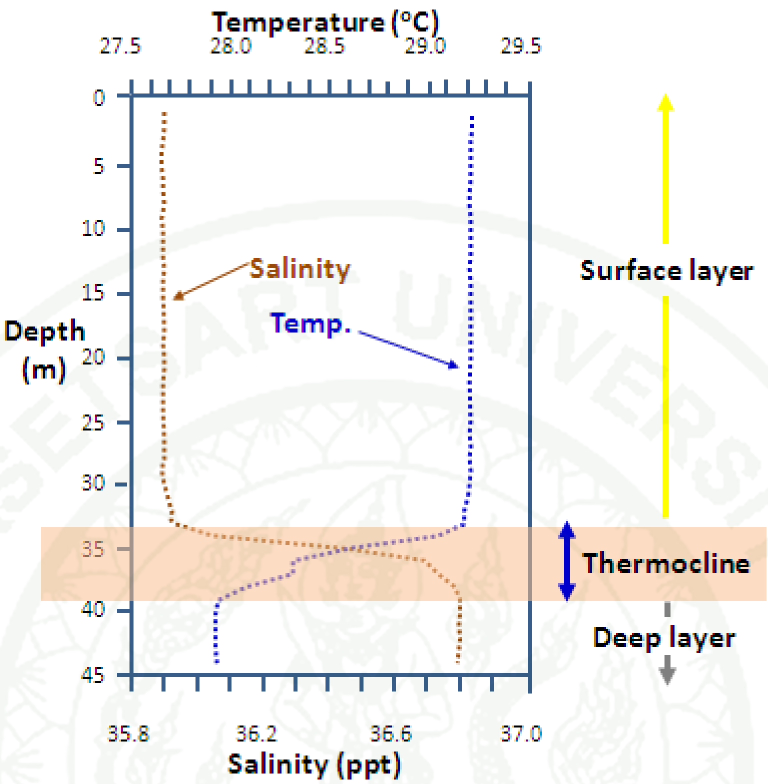


Figure 25 The salinity and temperature profile in vertical depth in Gulf of Thailand.

Source: Modified Anonymous (n.d.)

Process and equipment design simulation

There are three design parts including:

1. Waste gas pipeline network simulation
2. Waste gas treatment process and seawater desalination process simulation
3. CFD simulation of hydrocyclone

The detail of parameter setting up and the assumption in the simulations may be identical or different depended on the criteria of simulation.

1. Waste gas pipeline network simulation

Aspen Flare Analyzer is applied with the assumption that the pressure of supply gas is under flare system condition and also the user receiving pressure is designed at atmospheric pressure. Therefore, the utilization of this software is allowable, and this software gives more detail compared to the process simulation software. By these reason above, Aspen Flare Analyzer is selected to use for pipeline sizing in the study.

Gas compositions and gas flow rate are deployed from Aspen Plus. The condition of the simulation is considered on the saturated gas condition (water saturated condition) at supply pressure = 0.5 barg and 35 ^{o}C (Normal operating temperature). Although the actual supply pressure is in the range of 0.5-3.5 barg, only 0.5 barg is selected because it would help to be the margin in the design of equipment and pipeline size and the economic evaluation of the project.

The pipeline configuration bases on the source of gas and existing pipe rack or possible area for pipeline installation between supply source and destination process (LNG receiving terminal plant). There are 6 scenarios with consideration of range of supply temperature and the flow rate capacity of waste gas stream.

The thermodynamic equation package is designed to consistent with process simulation (Peng-Robinson). For the pressure drop, the equation of “Isothermal Gas” is applied because this method is designed for a compressible gas that assumes the isothermal expansion of the gas as it passes along the pipe. Flare System Analyzer uses averaged properties of the fluid over the length of the pipe. The outlet temperature from the pipe is calculated by adiabatic heat balance either with or without heat transfer. Pressure losses due to change in elevation are ignored.

The justification criteria of the design are the pressure drop across the network, pipeline erosion factor and noise generation in pipeline.

2. Process simulation

Aspen Plus is selected to design the process model and its simulation for both waste gas treatment system and freeze desalination system. It is able to simulate the hydrocarbon related system such as for petrochemical plant, and it can be simulated the electrolyte system by using built-in library of software.

2.1 Waste gas treatment system

The input of gas composition is the standard component in the library included thermodynamic database; Peng-Robinson, which is the same method to pipeline network simulation. The pressure drop across of all equipments is neglected. The ambient temperature is 35 ^oC, and it is the same to the gas supply temperature at saturated condition. The variation of the operating temperature of the supply is ignored because of the design stage. Operating temperature of the main equipment depends on the limitation. For the gas blower and gas booster compressor, the maximum outlet temperature is limited at 60 ^oC because of the limitation of seal part in the compressor itself.

2.2 Seawater desalination system

Standard NRTL of thermodynamic equation package (ELECNRTL) was selected as base modeling method in electrolyte system. For the calculation option, salt formation and ice formation were selected. However, water dissociation reaction is excluded because there is no reaction involved the H_3O^+ (or H^+) and the OH^- ions. In other words, if there is a possibility of water being generated (from combining H_3O^+ with OH^-), the water dissociation reaction is required and will be generated.

Pressure drop across all equipments is ignored as considering for waste gas treatment system. Supply seawater temperature = 35 ^oC, which is equal to normal ambient temperature. The operating temperature of desalination system is justified from the simulation result. The eutectic point of reference information from literature survey as Table 9 is used for the data calibration.

Table 9 Reference eutectic temperature of salt solution with a different kind of solution system

Solution system	**Eutectic Temperature, ^oC**	**Eutectic solution Composition, %wt.**
NaCl	-21.2	23.3% NaCl
NaCl + KCl	-22.9	20.2% NaCl + 5.8 % KCl
NaCl + Na_2SO_4	-21.7	22.8 % NaCl +0.3% Na_2SO_4
$CaCl_2$	-49.8	30.22% $CaCl_2$
$MgCl_2$	-33.6	21.0% $MgCl_2$
$CaCl_2$+ NaCl	-55	26% $CaCl_2$ +5% $MgCl_2$
$MgCl_2$ +NaCl	-35	22.7% $MgCl_2$ +1.56 % NaCl

Source: Barduhn and Manudhane (1979)

This reference information is one source of data for the simulation calibration test. However, the direct reference information is the laboratory freezing point test of seawater. The model of each system is generated and checked the eutectic point according to detail in the system. The eutectic is assumed to the maximum operating temperature in the design, however, in this process design, the freezing point is considered as the operating temperature criteria. Therefore, the freezing point from laboratory testing is required. Another selection criterion is the salinity of the product. If the product salinity is higher than drinking water specification, the operating temperature must be higher to reduce the salt contented in the product.

In addition, in some configuration, the salinity of fresh water product from freeze desalination may not reach the specification or the cost of operation (OPEX) may lead the project lost on profit. Therefore, the 2nd stage of desalination may be required. In the study, Reverse Osmosis (RO) technique is selected to apply to the purpose. Table 10 shows the basic specification of RO performance.

Table 10 The comparison table of desalination technique

Parameter	Unit	Seawater RO	Brackish RO	Multi-Effect Distillation	Electrodialysis Reversal
Feed Water Salinity	mg/L TDS	>32,000	<32,000	>35,000	3,000-12,000
Product Water Salinity	mg/L TDS	<500	<200	<10	<10
Minimum Product	m^3/Day	0.5	0.5	120	90
% Recovery	%	≤30	≥80	40-65	>90
Energy Required	-	Electrical Energy	Electrical Energy	Electrical Energy or Waste Heat energy	Electrical Energy or Waste Heat energy
Capital Cost	AUD/(m^3/day)	1.6-2.5	0.6-1.8	2.5-3.9	0.57-3.25
Operating cost	AUD/m^3	1.89-2.20	0.65-1.50	With Waste Heat: 0.55-0.95 Without Waste Heat: 1.80-2.80	1.00-2.80

Source: Modified Agriculture, Fisheries & Foresty-Australia. (2002)

In the study, seawater RO is selected for applied because TDS in the product is higher than the water specification for hydrocyclone configuration.

3. CFD simulation of hydrocyclone

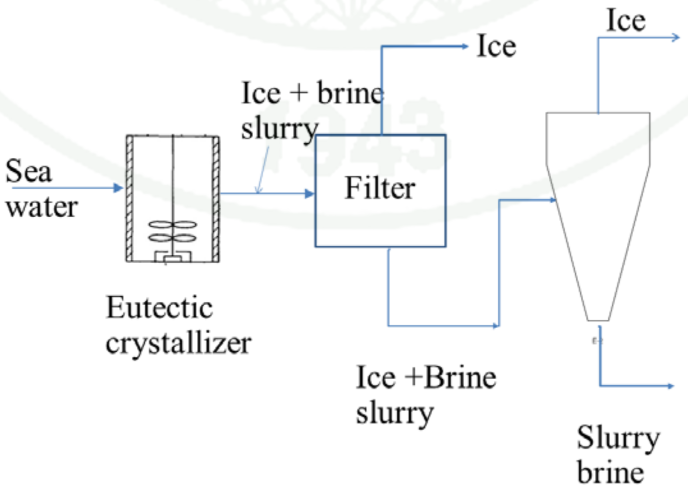


Figure 26 Overall crystallization process.

In this study, hydrocyclone is preliminary designed by CFD simulation. It is expected to apply for ice-seawater separation because the hydrocyclone generally has a small footprint with large capacity in other application. In addition, it would reduce the investment cost of the ice-seawater separation system. Therefore, in the study, the scope of hydrocyclone improvement for freeze desalination application is included in the scope.

The parameters of CFD simulation of hydrocyclone are the density and viscosity, determined by Aspen Plus simulation of eutectic crystallizer as shown in Figure 26. The selected condition of the seawater exiting the eutectic crystallizer is -16 ^{o}C and 5 barg, and then sent to filtration equipment to remove some of ice before feeding to hydrocyclone. The objective is to obtain the optimum operating condition of hydrocyclone; inlet velocity and the inlet ice fraction. In addition, the investigation of the effect of overflow and underflow diameter on the flow discharge pattern of ice-seawater slurry inside hydrocyclone is included as shown in Figure 27. The physical properties obtained from Aspen Plus are presented in Table 11.

Table 11 Parameters for CFD simulation

Material	**Density (kg/m^3)**	**Viscosity (kg/ms)**
Seawater	1,221	0.0039
Ice	917	-

The process of the designation of hydrocyclone is shown as the diagram below:

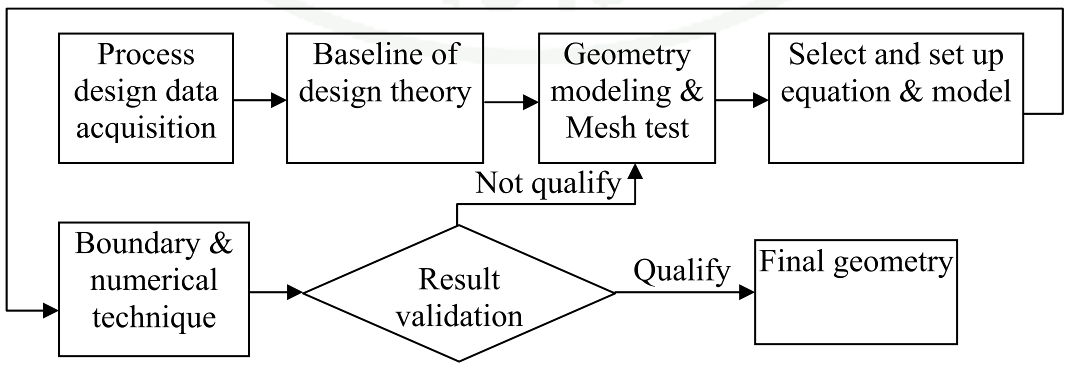


Figure 27 Flow diagram for process and equipment design

The validation criteria in Figure 27 is the convergence of the model and the result of the CFD. The expected result of the hydrocyclone design is the separation efficiency. It would not less than 80%.

3.1 Geometry and mesh test

The important task for success in separating ice slurry from the brine slurry is to select the geometry of hydrocyclone, which can separate small density materials. Rietema design geometry is used to separate the solid from the liquid (Svarovsky, 1977). Ruiyun's work reveals that Rietema hydrocyclone was not applicable for concentrating ice slurry (Ruiyun et al., 2011). The geometry of hydrocyclone in this research is retrieved from the de-oiling hydrocyclone of HY1 proposed by Yuan (Elsayed, 2005) because the de-oiling hydrocyclone can separate the oil and water, which have small density difference.

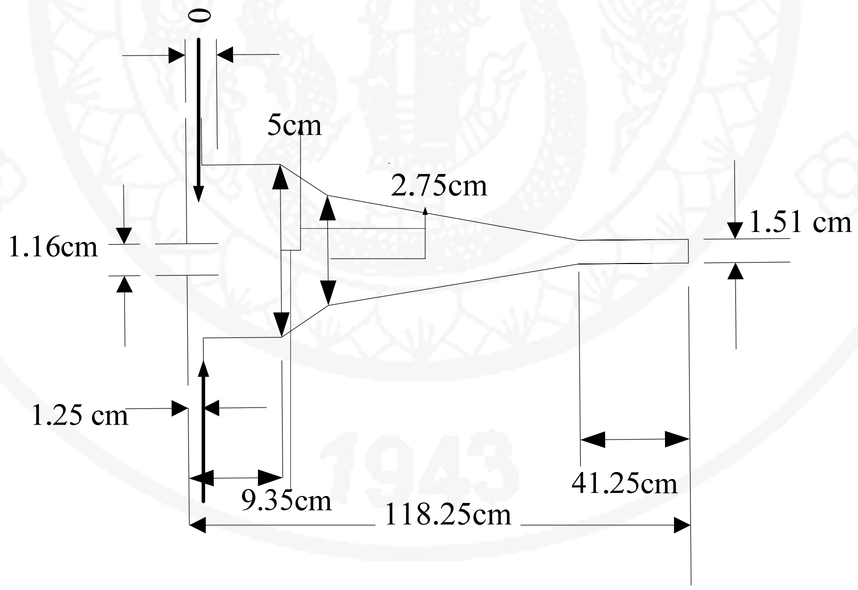


Figure 28 Hydrocyclone geometry

Finer element of hydrocyclone provides the better result of CFD simulation, but it also takes longer computational time. Five different mesh sizes between 120,000 - 400,000 nodes are examined to find the optimum node number. The optimum number of nodes is selected by the criterion that the relative difference of the mass flow rate at overflow and underflow between nodes must be less than 5×10^{-8}. The optimal number of 5 cm chamber diameter of hydrocyclone is 330,000 nodes.

3.2 Governing equation

The key to successfully modeling the high swirling flow inside hydrocyclone is a selection of turbulence model. Turbulence models have been categorized into three categories: Direct Numerical Simulation (DNS), Large Eddy Simulation (LES), and Reynold Averaging Navier-Stroke (RANs). Direct numerical simulation requires no additional modeling, but it is impractical due to highly computationally expensive. The flow inside hydrocyclone is anisotropic turbulence and has rapid strain change. LES falls between DNS and RANs. LES requires transient simulation modeling; consequently, LES takes longer computational time to obtain the stable flow. Therefore, RANs equations must be considered. Bhaskar et.al (2007) carried out the study of difference turbulence model such as standard k-ε, Re-Normalized Group (RNG) k-ε, and Reynold Stress Model (RSM) by comparing the simulation results and the experimental results of the water exiting the spigot opening. The results indicate that RSM model give a better agreement with the experimental results than other type of RANs model because it considers the marginal error of 4-8%. RSM model is the most suitable turbulence modeling approach because it accounts for the effect of streamline curvature, swirl, rotation, and rapid changes in strain rate.

$$\frac{\partial}{\partial t}(\rho u_i)+\nabla\cdot(\rho u_i u_j)=-\nabla p-\rho\overline{u_i' u_j'}+\rho g+F \tag{10}$$

RSM abandons the assumption of isotropic turbulence; thus, it requires additional modeling of Reynold stress terms term as shown in equation (10).

3.3 Multiphase model

There are two approaches for multiphase flow modeling: Eulerian-Eulerian approach and Euler-Lagrangian approach. The difference between Eulerian-Eulerian approach and Euler-Lagrangian approach is that the dispersed phase in the Eulerian-Lagrangian is not treated as a continuum, and the control volume of the dispersed phase moves with the fluid. Eulerian-Lagrangian approach is suitable for the solid volume fraction mixture below 10%. Above 10% of ice fraction, Eulerian-Eulerian approach must be applied.

The Eulerian-Granular model is the extension of Eulerian-Eulerian approach applied to calculate the fluid-solid flow behavior for the separation of ice and seawater. The momentum equation of fluid-solid is shown in following:

$$\frac{\partial}{\partial t}\left(\alpha_s \rho_s \vec{v}_s\right) + \nabla\cdot\left(\alpha_s \rho_s \vec{v}_s \vec{v}_s\right) = -\alpha_s \nabla p - \nabla p_s + \nabla\cdot\bar{\bar{\tau}}_s + \alpha_s \rho_s \vec{g} + \sum_{l=1}^{N}\left(K_{ls}(\vec{v}_l - \vec{v}_s) + \dot{m}_{ls}\vec{v}_{ls} - \dot{m}_{sl}\vec{v}_{sl}\right) + \left(\vec{F}_s + \vec{F}_{lift,s} + \vec{F}_{vm,s}\right) \qquad (11)$$

Equation (11) shows the term $\sum_{l=1}^{N}\left(K_{ls}(\vec{v}_l\ \vec{v}_s) + \dot{m}_{ls}\vec{v}_{ls}\ \dot{m}_{sl}\vec{v}_{sl}\right)$. It is the interaction term between fluid-solid phases. The fluid-solid interphase exchange coefficient models, which are available in CFD software, are Wen-Yu, Gidaspow, and Symlal-Orbien. Wen-Yu model is appropriate for dilute phase calculation, and Gidaspow is mostly applied to calculate the dense fluidized bed model. Symlal-Orbien predicts the accurate results when compared with other drag models. If the range of ice solid fraction in hydrocyclone feed is 0.1-0.4, Symlal-Orbien and Gidaspow must be considered. The validating results for drag force coefficient can be obtained from the CFD modeling of fluidized bed reactor. Almuttahar (2008) studied the CFD modeling of air and Fluid Catalytic Cracking (FCC) particles in the riser of high density. In addition, the effect of drag models on the simulation results are investigated by comparing the simulated profiles of fluid and solid velocity with the experimental data. Symlal-Orbien provides the better agreement results with the experimental data. Consequently, the drag model, Symlal-Orbien is chosen to calculate the solid phase. The Symlal-Orbien fluid-solid momentum exchange coefficient is described by

$$K_{sl} = \frac{3}{4}\frac{\alpha_s \alpha_l \rho_s}{v_{r,s}^2 d_s} \cdot C_D \cdot \left(\frac{Re_s}{v_{r,s}}\right) \cdot |\vec{v}_s - \vec{v}_l| \tag{12}$$

$$C_D = \left(0.63 + \frac{4.8}{\sqrt{Re_s / v_{r,s}}}\right)^2 \tag{13}$$

$$Re_s = \frac{\rho_l \, d_s}{\mu_l} \tag{14}$$

$$v_{r,s} = 0.5\left(A - 0.06 Re_s + \sqrt{(0.06 \cdot Re_s)^2 + 0.012 Re_s (2B - A) + A^2}\right) \tag{15}$$

With

$$A = \alpha_l^{4.14} \tag{16}$$

$$B = 0.8 \times \alpha_l^{1.28} \text{ for } \alpha_l \leq 0.85 \tag{17}$$

$$B = \alpha_s^{2.65} \text{ for } \alpha_s > 0.85 \tag{18}$$

The granular phase parameter is calculated as following:

Solid pressure (P_s)

$$P_s = 2\rho_s (1 + e_{ss}) \alpha_s^2 g_{0,ss} \theta_s \tag{19}$$

Radial distribution ($g_{0,ss}$)

$$g_{0,ss} = \left[1 - \left(\frac{\alpha_s}{\alpha_{s,max}}\right)^{\frac{1}{3}}\right]^{-1} \tag{20}$$

Collisional viscosity ($\mu_{s,col}$)

$$\mu_{s,col} = \frac{4}{5}\alpha_s \rho_s d_s g_{0,ss} (1 + e_{ss}) \left(\frac{\theta_s}{\pi}\right)^{1/2} \alpha_s \tag{21}$$

Kinetic viscosity ($\mu_{s,kin}$)

$$\mu_{s,kin} = \frac{\alpha_s d_s \rho_s \sqrt{\theta_s \pi}}{6(3 - e_{ss})}\left[1 + \frac{2}{5}(1 + e_{ss})(3e_{ss} - 1)\alpha_s g_{0,ss}\right] \tag{22}$$

Granular bulk viscosity $$\lambda_s = \frac{4}{3}\alpha_s\rho_s d_s g_{0,ss}(1+e_{ss})\left(\frac{\theta_s}{\pi}\right)^{1/2} \quad (23)$$

3.4 Boundary and Numerical technique

The volume of feed inlet can be specified by both the magnitude velocity and the pressure inlet. The overflow and the underflow were assigned as the pressure outlet set to zero. The outflow boundary cannot be applied to the overflow and the underflow because Fluent assumes that the split fraction is equal in both overflow and underflow. The Fluent solver is 3D double precision solver, Phase Coupled Simple (PC-SIMPLE), and all quantities are discretized with the first order upwind scheme. PC-SIMPLE is the extension of the SIMPLE algorithm to the multiphase flow. Fluent 14 allows to select of the discretization scheme from several upwind scheme, first-order upwind scheme, and the QUICK scheme. The assumption of the first order upwind scheme is that the scalar value stored at the cell centers equals to the face center. Higher order upwind scheme means the scalar value at the face is discretized from the cell value, so the second order upwind scheme provides more accurate results than the first order upwind scheme, but take more computational time. In this work, Eulerian-Granular model coupled with Reynold Stress Model (RSM) is highly computational expensive; consequently, the first order upwind scheme is applied instead. The convergence criteria should fall below 10^{-5}, therefore, the convergence is assigned as 10^{-6}.

Preliminary cost estimation and economic evaluation

The cost database, it is acquired from 5 sources:

1. Publication web site (Anonymous, n.d.): This information is applied to the most of equipment in the process except gas-liquid separator or gas-liquid knock out drum. This database based on cost of equipment in year 2007 and the price basis location is FOB at gulf coast of USA (ICC Publishing, 2000).

2. Aspen software: The price of gas-liquid separator or gas-liquid knock-out drum based on the software databank.

3. Local vendor database: This information is applied to all of pipeline; waste gas pipeline network and seawater transfer pipeline, in the study.

4. The cost of civil work is assumed from actual of industrial experience.

5. RO CAPEX and OPEX: the information is extrapolated from the database in Agriculture, Fisheries & Forestry-Australia report (2002) as shown in Tables 12 and 13. The assumptions in Table 12 are below:

a. Electrical power supply available

b. Basic Pre-filtration included

c. Cost quoted are for plant only i.e. no civil works, external tanks and pipe works.

The assumptions of Table 13 are

a. Electrical power cost = 4.2 THB/kWh

b. Cost to desalinate water only - does not include the cost to deliver water to and from the desalination plant.

c. Operating cost assumes basic pre-filtration sufficient for purpose

d. Operating cost assumes basic pre-filtration sufficient for purpose

With above cost analysis assumption, the currency exchange rate is assumed to 30 THB/AUD.

Table 12 CAPEX analysis of seawater RO with variation of TDS and Production flowrate

TDS (mg/L)	**RO Production flow rate (m^3/d)**			
	5	15	50	21,600
2,000	5,000	13,000	35,000	5,941,801
10,000	7,000	17,500	44,000	5,641,437
35,000	12,000	27,000	85,000	14,374,953
170,000	40,286	83,015	293,112	42,845,000

Source: Agriculture, Fisheries & Foresty-Australia. (2002)

Table 13 OPEX analysis of seawater RO with variation of TDS and Production flowrate

TDS (mg/L)	**RO Production flow rate (m^3/d)**			
	5	15	50	21,600
2,000	1.00	0.90	0.65	0.21
10,000	1.50	1.30	0.93	0.27
35,000	2.20	2.00	1.89	1.26
170,000	3.35	3.00	3.08	4.96

Source: Agriculture, Fisheries & Foresty-Australia. (2002)

RESULTS AND DISCUSSION

In the study, the result is divided in 7 parts following:

1. Process design conception selection
2. Waste gas pipe line network
3. Process design of waste gas treatment system
4. Process design of seawater desalination plant
5. CFD modeling and analysis of hydrocyclone
6. Cost Estimation
7. Preliminary economic evaluation and sensitivity analysis

The detail of each part is shown as below:

1. Process design concept selection

1.1 Waste gas treatment system

In the study, there are many options of the process design concept. It can be categorized in 3 schemes.

1. 1^{st} schematic: Waste gas treatment by indirect cooling without product purification process.

2. 2^{nd} schematic: Waste gas treatment by indirect cooling with product purification process.

3. 3^{rd} schematic: Waste gas treatment by direct cooling with product purification process.

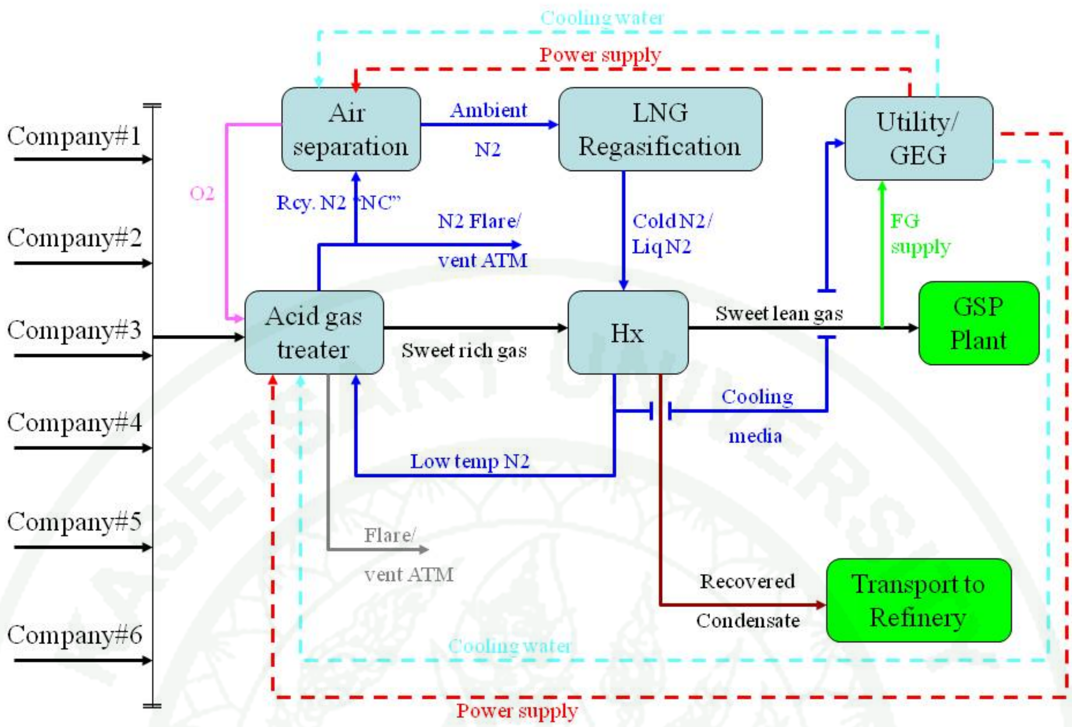


Figure 29 Process block diagram of waste gas treatment by indirect cooling without product purification. (1st schematic)

All schematics have the same design concept, that is, to centralize the waste gas from each source to treat in specific location. Referring to the assumptions, there are 6 sources of waste gas and route to waste gas network system then deliver to gas treatment. The different is the cold energy recovery from the LNG regasification system. The 1st and the 2nd schematics use N_2 as the cooling media to recover cold energy from LNG vaporizer. N_2 is designed to produce from air separation unit by cryogenic separation. O_2 from this unit is designed to feed to acid gas treatment system. Since waste gas from the network composes of many contaminants which are possible to perform oxidation reaction with O_2, therefore the advantage of this design is to utilize the natural source of air for deactivate the toxic component without loss of energy. N_2 from air separation unit is compressed and fed to the vaporizer for LNG vaporization. Low temperature N_2 feeds to heat exchanger for condensing the non-hydrocarbon components from the acid gas treater before delivered to methane purification. If the design is to send the sweet NG to Gas Separation Plant (GSP), the methane purification process is not required as shown in Figure 29. In case of methane purification requirement (Figure 30), the product from the system is the Natural Gas for Vehicle (NGV). By-product from this system is the condensed liquid from purification unit composed of non-hydrocarbon; such as VOCs and sulphur oxide derivative components, and heavy hydrocarbon; such as pentane and BTEX.

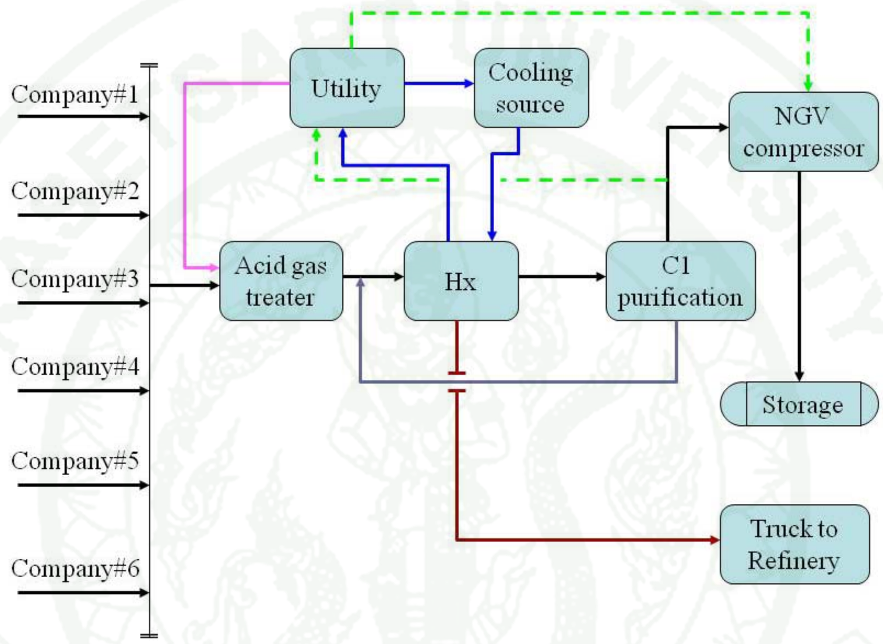


Figure 30 Process block diagram of waste gas treatment by indirect cooling with product purification. (2^{nd} schematic)

In case of the 3^{rd} schematic, the concept of design is slightly different from the previous ones. This concept is to minimize the existing facilities of LNG receiving terminal as much possible. Therefore, the heating media for the vaporizer is the same; seawater. However, the operating temperature of the vaporizer can be adjusted to get the benefit for seawater desalination function also. By this concept, the freeze desalination technique is also applied for the vaporizer and the product of the unit is fresh water as shown in Figure 31.

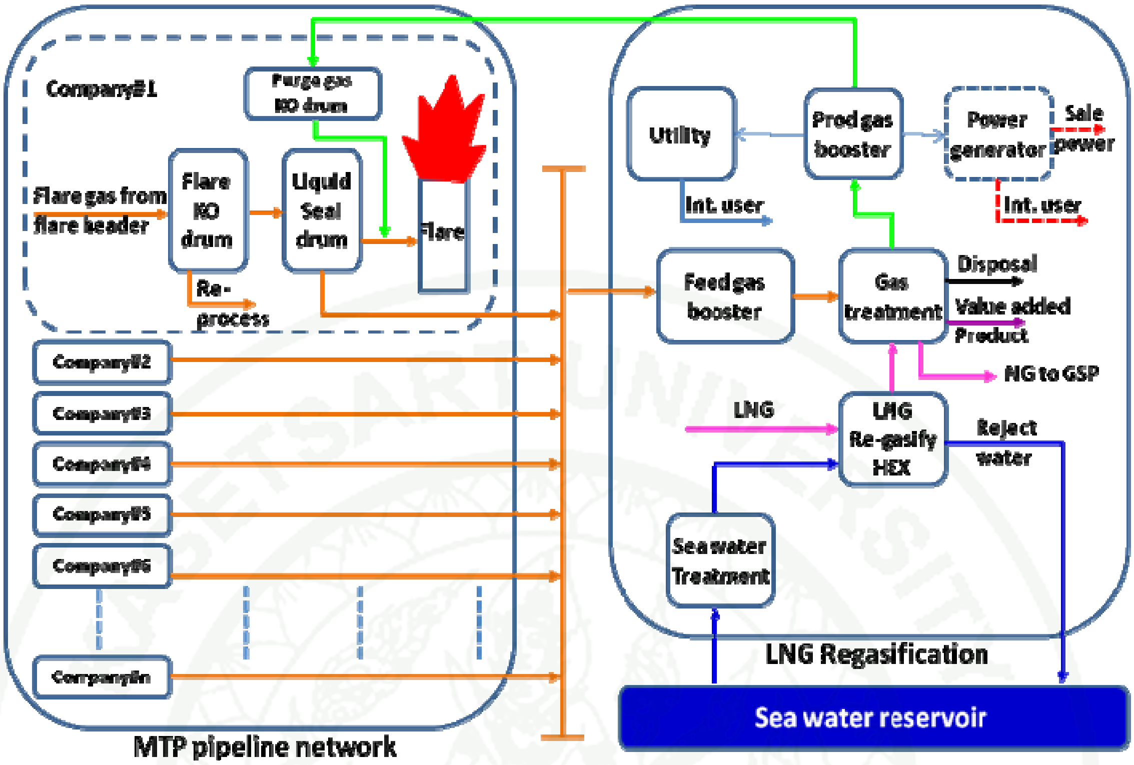


Figure 31 Process block diagram of waste gas treatment by indirect cooling with product purification.

Waste gas from pipeline network is routed to the treatment system and treated with the cryogenic technique. With the existing LNG vaporizer type, Open Rack Vaporizer (ORV) is used for LNG vaporization. Seawater is the heating media for this unit. The outlet seawater from the vaporizer is disposed back to the sea. After LNG is heated up, the low temperature NG is fed to the Low Temperature Separation (LTS) system for waste gas treatment. Liquid products from the LTS contain fuel liquids with sulfur compounds. They must be sent for disposal with specific waste treatment system, especially for sulfur and VOCs contaminants or utilizing as fuel for further proposes. Treated gas can be used for purge gas of source plant, utilizing as fuel for power generation for internal process facility. The remaining power can be sold to main power grid directly. Since the compositions of waste gas from all source compose of hydrocarbon e.g. methane, ethane, etc., it can be utilized as energy source for electricity production.

Since there are both advantages and disadvantages in each concept, therefore, the comparisons between these concepts are summarized as shown in Table 14.

Table 14 Pros and Cons of all process conceptual scheme.

	Pros			Cons		
Parameter	**Scheme #1**	**Scheme #2**	**Scheme #3**	**Scheme #1**	**Scheme #2**	**Scheme #3**
Recovery energy loss			The loss energy by direct cooling is less than indirect cooling.	Since N_2 is heating media at LNG vaporizer and transfer the energy to waste gas for condensing the toxic and heavy hydrocarbon component therefore, the loss of energy transfer via N_2 media is more than direct energy transfer.		
Additional facilities	Air separation unit, N_2 and O_2 compressor, heat exchanger, new LNG vaporizer	Air separation unit, N_2 and O_2 compressor, heat exchanger, New LNG vaporizer, methane purification	Fresh water and reject water Tank, ice filter, heat exchanger			
Additional footprint requirement			Medium area requirement if compare to scheme#1 and #2 because thermal conductivity of seawater is much higher than	Large area required for new LNG vaporizer because the concept is the heat transferring by gas, therefore, the thermal conductivity is very low if comparing to liquid such as seawater. In addition, it may also		

Table 14 (Continued)

Parameter	Pros Scheme #1	Pros Scheme #2	Pros Scheme #3	Cons Scheme #1	Cons Scheme #2	Cons Scheme #3
			N_2	requires footprint more than Ambient air-heated vaporizer by the same reason.		
Process modification				Require new design		
Energy consumption			Less than scheme# and 2 because there is no loop of utility compressor e.g. N_2	The additional loop of N_2 requires more additional energy consumption. Especially, in scheme#2, the C1 purification leads to have more energy consumption also.		
Period of modification			Able to use existing system in parallel	Long period of modification due to many facilities involvement		

This table shows that scheme#3 gives more benefit than the others in many issues such recovery energy loss and installation footprint of the additional facilities. In the 1st and 2nd schemes, N_2 is used as energy transferring media, therefore, the loss at the absorption stage and at transferring stage are more than single stage energy absorption such as seawater direct energy absorption. Moreover, N_2 is low thermal conductivity gas; the required capacity of N_2 production unit and heat exchanger is quite large in comparison to the liquid heating system and it also required the large installation area. Therefore, the scheme#3 is suitable under the concept of implementation.

The conceptual process of heating media has to be generated and the seawater heating media; the freeze desalination, is selected to apply to the integrated system.

1.2 Seawater desalination system

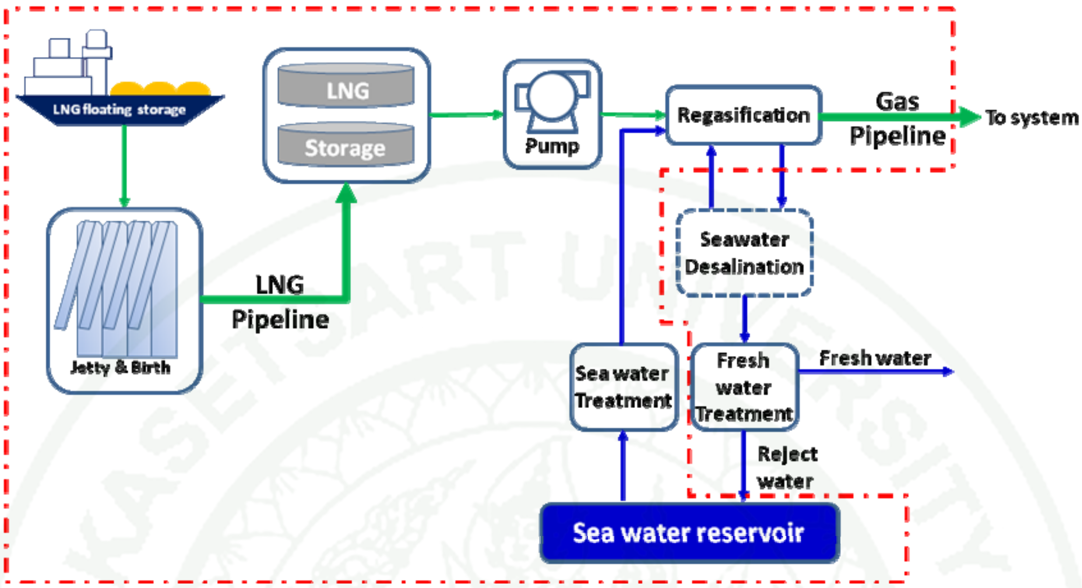


Figure 32 Block flow diagram of MTP LNG terminal and integrated seawater desalination.

The concept of this process is to integrate the existing with LNG regasification process with seawater desalination. In general, there are many types of LNG regasification/vaporizer (Yang and Huang, 2004). In case of MTP LNG terminal, ORV type is designed for regasification purpose. The principle of ORV type is to heat LNG by using seawater as heating media. In Figure 32, 5 MTPA LNG from floating storage is designed to feed to LNG storage via loading arm in Jetty unit and LNG pipeline. LNG is pumped by cryogenic pump from storage to regasification unit. In this unit, pre-treated seawater from seawater treatment is pumped to ORV as heating medium. LNG is changed to vapor phase at this unit as Natural Gas (NG) and then sent to gas metering and gas distribution system via gas pipeline. Cooled seawater from ORV is then returned to seawater. In general, ORV is designed to operate at the temperature higher than 8 ^{o}C (Yang and Huang, 2004) but for desalination, seawater must be operated at lower 0 ^{o}C. At this condition, this seawater is in the eutectic condition and leads ice (solid fresh water) and brine (liquid salt solution) to separate by phase different. Ice/brine slurry from desalination unit is separated by using filter and hydrocyclone to increase the production recovery efficiency. Rejected solution (brine solution) is sent to buffer pond for increasing temperature before charged back to seawater or GOT.

1.3 Waste gas recovery system

The common system before delivered waste gas to pipeline network is waste gas recovery system in any process scheme. Figures 33 and 34 show the conventional flare system and method for modifying the existing system to divert waste gas to pipeline network.

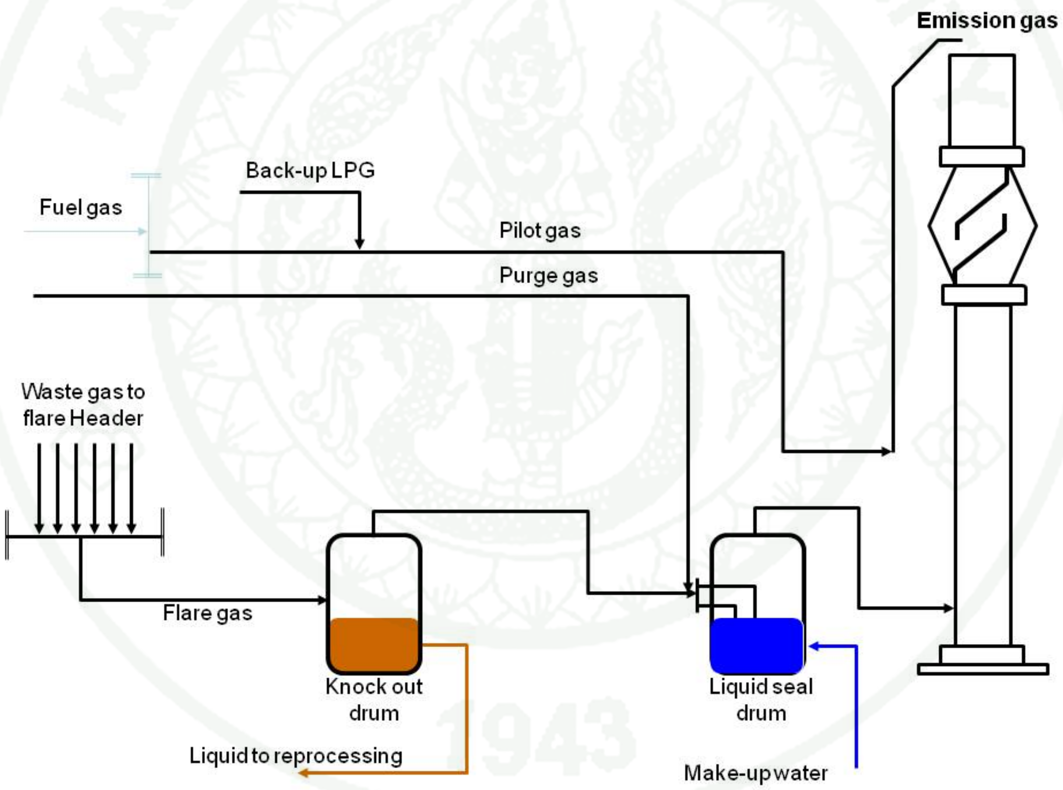


Figure 33 Conventional process flow diagram of flare system.

For conventional flare system process, there are 2 sources of waste gas; flare header, and purge gas (untreated gas). Gas from flare header delivers to flare Knock-Out Drum (KOD) for separating the liquid, which is entrained or carried with gas to flare stack. Liquid from KOD is returned to primary system for re-processing at closed drain system. After gas passing through KOD, this gas is sent to the staging drum, flare stack and then burner, respectively. At the meantime, purge gas is fed to flare stack continuously to prevent air ingress and leads to have flash back scenario in case of lack of waste gas flowing from flare header.

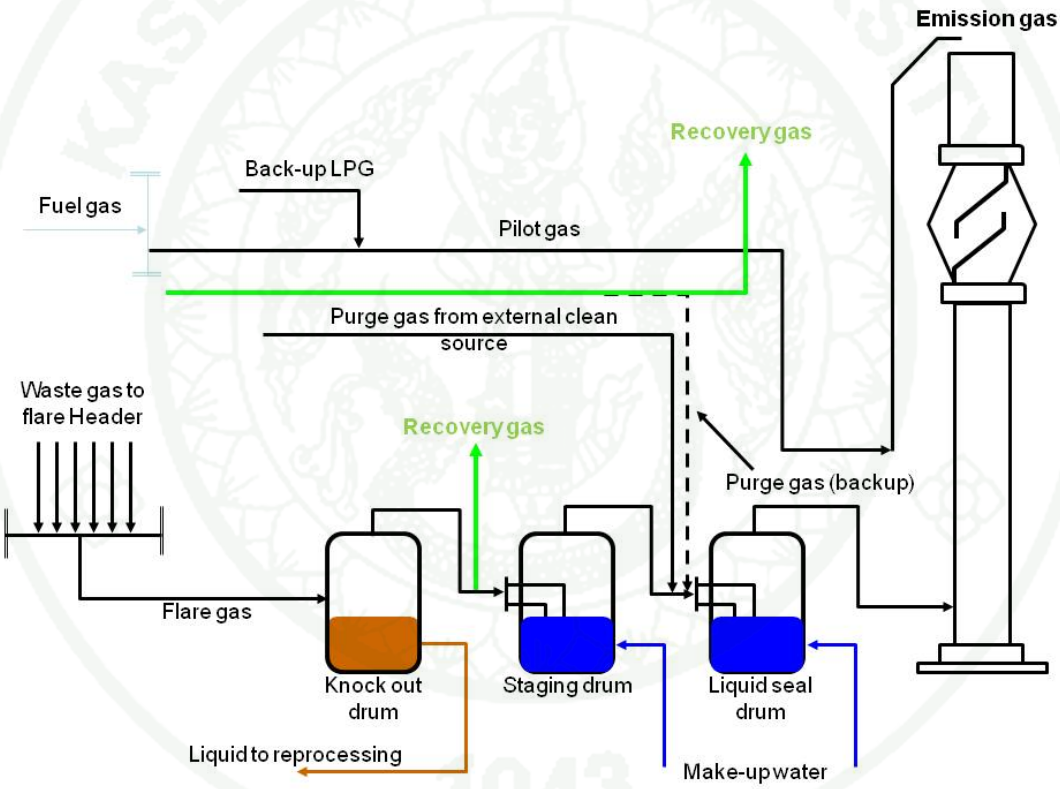


Figure 34 Modified process flow diagram of flare system for waste gas.

Although only the purge gas is considered as a waste gas in the study but in real operating condition, waste gas source composed of 2 sources; a) excess gas from flare header and b) purge gas. Therefore, in the study, the modification needs to be considered for both sources of waste gas as following:

a) Staging drum installation - This configuration is designed to recover the excess waste gas from flare header. Refer to Figure 34, gas passed through KOD to staging drum. The function of this staging drum is designed to staging or diverting the flow of flare gas to another direction. It is not designed to prevent the back flowing from flare tip as staging drum. The purpose of liquid in staging drum is to control the flow of gas by hydraulic pressure of the liquid height in the drum. During normal operation, the flare gas is diverted to the waste gas network and sent to waste gas treatment plant. In case of emergency such as blowdown case, the large flow rate of blowdown gas from whole process routes to flare gas header as primary route. However, the pressure of gas due to backpressure from network leads the hydraulic pressure of liquid in staging drum less than set point; then, this amount of gas is forced to route to local flare as original plant design.

b) Branch purge gas line - This configuration is designed to recover untreated purge gas and to deliver to the integrated system for waste gas treatment. Purge gas from existing line is branched to the waste gas pipeline network. The modified line of purge gas is designed to feed directly to existing purge gas tie-in point (Solid black line). A new source of purge gas is the clean source from external supplier such as NG from PTT distribution pipeline network in MTP. The existing line (Dotted black line) is designed to use as back-up purge gas in case the clean purge is not available. By this design, the availability and function of purge gas to flare is still the same by this configuration, although, the consumption of purge is often changed by the operating condition of the plant such as normal condition and emergency condition.

2. Waste gas pipe line network

2.1 Pipeline route selection

Each waste gas pipeline is designed to utilize the existing pipeline support in the area. The distance of pipeline and the route are shown in the Figure 35.

Figure 35 Route and distance of pipeline between waste gas source and LNG receiving terminal.

Source: Modified from Google Earth (n.d.)

The common component of the pipeline route comprises of block valve, 90° elbow, and standard tee branch. The list of the components is shown in Table 15. In the pipeline network design, it is separated in segment for size calculation and simulation. Waste gas composition and flow rate through each segment based on the result from process simulation.

Source no. 6 (Flare#6) has the longest distance about 9,750 m. and source no. 5 (Flare#5) is the nearest source. The total pipeline distance is 12,200 m. The elbow is designed to use 90^o for turning the flow. The estimated numbers of elbow also includes the bridge support at the road crossing point.

In each waste gas source, gate vale is applied instead of ball valve the shutdown valve due to simulation limitation and check valve is designed to prevent back flow from the main pipeline in the network. Including with the fitting and valve of each line, the overall calculated equivalent length is 14,166.31 m.

Table 15 Pipeline configuration detail in each segment of waste gas pipeline network.

Segment	Connection		Flow rate		Pipe	Equivalent length (m)	Valve Gate type	Fitting Qty. Total	
	From	To	Volumetric (MMSCFD)	Mass (kg/h)	Distance (m)		100% open	90^o Elbow	90^o Tee
1TP12	GSAT1	TP12	1.22	1794.27	600	624.76	2	11	2
2TP24	GSAT2	TP24	1.74	2558.34	350	440.62	2	12	2
3TP36	GSAT3	TP36	2.04	2999.15	350	454.50	2	14	2
4TP43	GSAT4	TP43	4.12	6074.09	150	253.47	2	11	2
5TP51	GSAT5	TP51	5.87	8768.96	1000	1150.00	2	12	2
6TP36	GSAT6	TP36	8.05	12129.08	2200	2405.19	2	17	2
BLWKO1G	BLWKO1	BLW	23.04	34323.88	200	453.90	2	10	2
MHDR	TP51	BLWKO1	23.04	34323.88	1650	4071.69	0	12	2
TP12-51	TP12	TP51	17.17	25554.93	3800	1903.98	2	14	2
TP24-12	TP24	TP12	15.95	23760.65	850	1013.21	0	10	2
TP36-43	TP36	TP43	10.09	15128.23	500	669.42	0	11	2
TP43-24	TP43	TP24	14.21	21202.31	550	725.57	2	10	2

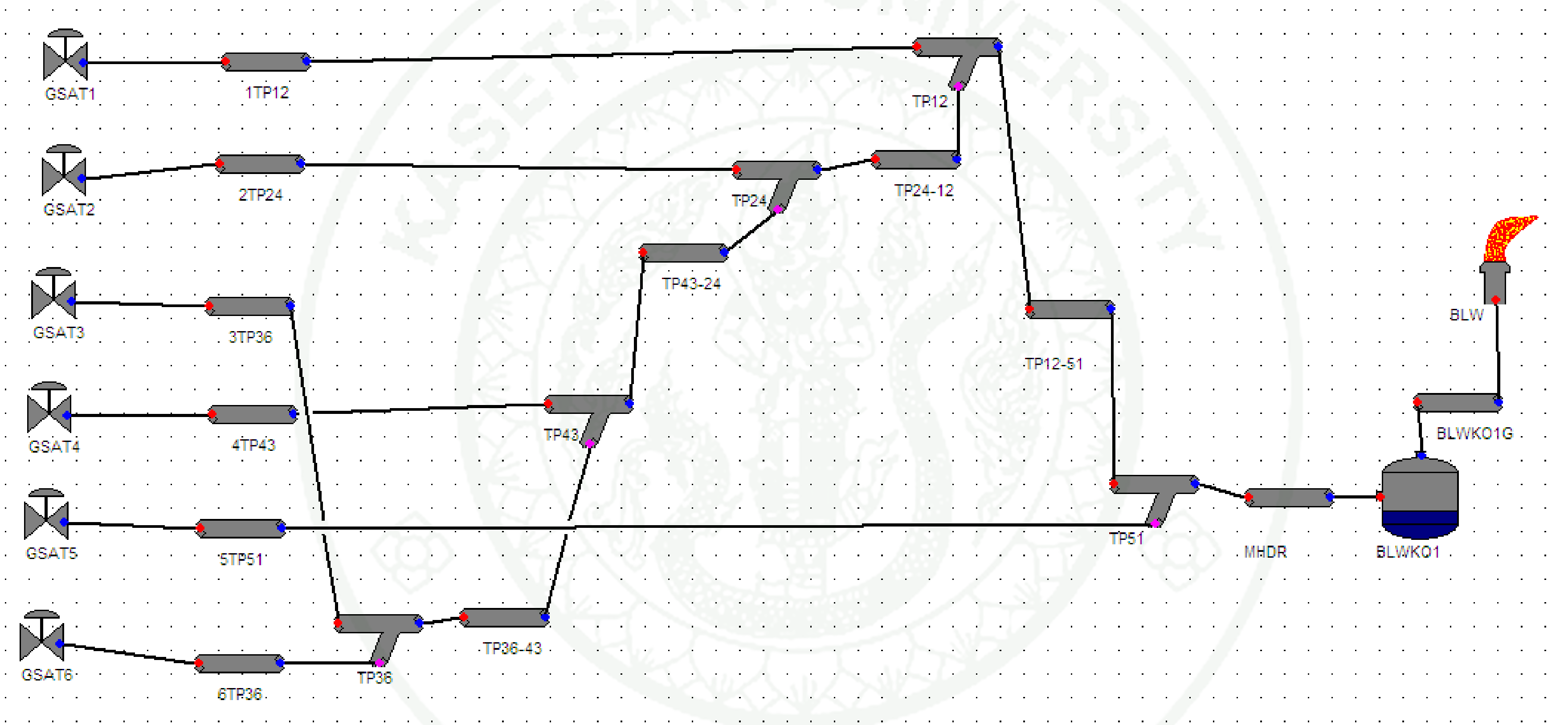


Figure 36 Process flow diagram of waste gas pipeline network in Aspen Flare System Analyzer.

Table 16 Mass flow rate and inlet pressure of waste gas in each segment of pipeline network for all scenario.

Segment	From	To	Gas Flow rate (kg/h)				Upstream pressure (Bara)			
			Normal		Emergency		Normal		Emergency	
			HiT	LoT	HiT	LoT	HiT	LoT	HiT	LoT
1TP12	GSAT1	TP12	422.78	408.37	1,794.27	1,733.95	1.0330	1.0323	1.4325	1.4201
2TP24	GSAT2	TP24	475.40	459.23	2,558.34	2,472.44	1.0217	1.0214	1.3773	1.3666
3TP36	GSAT3	TP36	501.71	484.66	2,999.15	2,898.50	1.0200	1.0198	1.4141	1.4027
4TP43	GSAT4	TP43	633.40	611.94	6,074.09	5,891.02	1.0202	1.0200	1.4254	1.4136
5TP51	GSAT5	TP51	712.32	688.21	8,768.96	8,511.81	1.0170	1.0168	1.2786	1.2708
6TP36	GSAT6	TP36	791.38	764.63	12,129.08	11,780.74	1.0203	1.0201	1.4701	1.4574
BLWKO1G	BLWKO1	BLW	3,536.99	3,417.05	34,323.88	33,288.45	1.0137	1.0137	1.0299	1.0294
MHDR	TP51	BLWKO1	3,536.99	3,417.05	34,323.88	33,288.45	1.0156	1.0155	1.1546	1.1503
TP12-51	TP12	TP51	2,824.67	2,728.84	25,554.93	24,776.64	1.0174	1.0172	1.2492	1.2421
TP24-12	TP24	TP12	2,401.89	2,320.46	23,760.65	23,042.70	1.0185	1.0183	1.3161	1.3072
TP36-43	TP36	TP43	1,293.09	1,249.29	15,128.23	14,679.24	1.0197	1.0195	1.4062	1.3950
TP43-24	TP43	TP24	1,926.49	1,861.23	21,202.31	20,570.25	1.0193	1.0191	1.3760	1.3655

Table 17 Velocity, maximum velocity following practice and maximum erosion velocity of waste gas in each segment of pipeline network for all scenario.

Segment	From	To	Max erosion velocity criteria[a] (m/s)	Design practice (m/s)				Simulated velocity (m/s)			
				Normal		Emergency		Normal		Emergency	
				HiT	LoT	HiT	LoT	HiT	LoT	HiT	LoT
1TP12	GSAT1	TP12	20	63.28	63.74	63.29	63.75	5.36	5.33	18.46	18.47
2TP24	GSAT2	TP24	20	63.28	63.74	63.29	63.75	3.44	3.42	14.26	14.28
3TP36	GSAT3	TP36	20	63.28	63.74	63.29	63.75	1.25	1.24	5.39	5.41
4TP43	GSAT4	TP43	20	63.28	63.74	63.29	63.75	2.79	2.77	19.75	19.80
5TP51	GSAT5	TP51	20	63.28	63.74	63.29	63.75	1.75	1.73	18.88	18.86
6TP36	GSAT6	TP36	20	63.28	63.74	63.30	63.75	0.90	0.89	9.94	9.97
BLWKO1G	BLWKO1	BLW	20	63.28	63.74	63.28	63.74	1.76	1.75	16.96	16.88
MHDR	TP51	BLWKO1	20	63.28	63.74	63.28	63.74	1.76	1.74	16.67	16.60
TP12-51	TP12	TP51	20	63.28	63.74	63.29	63.74	2.23	2.22	17.66	17.64
TP24-12	TP24	TP12	20	63.28	63.74	63.29	63.75	2.27	2.25	18.15	18.17
TP36-43	TP36	TP43	20	63.28	63.74	63.29	63.75	1.48	1.47	12.71	12.76
TP43-24	TP43	TP24	20	63.28	63.74	63.29	63.75	2.22	2.20	18.74	18.78

Note: [a] PTTEP PLC. (2010).

Table 16 shows inlet parameters such as gas flow rate and upstream pressure of each source and segment. The operating temperature of simulation case is 35 ^{o}C (normal temperature). The segment names in the table refer to sources and segment names in Figure 36. Table 17 shows that the velocity is lower than the maximum velocity by common design practice and it is lower than erosion velocity. Actually, the erosion velocity should be over the maximum velocity but the criterion of erosion velocity is limited by characteristic of gas. In case of sour gas (H_2S/CO_2 ratio > 0.05), the limitation of velocity in pipeline for carbon steel is only 20 m/s. That is the reason of low erosion velocity.

The result of Mach number is in the same trend as shown in Table 18. The criterion of Mach number is designed to 0.35 and the result of simulation is under the criteria for all scenarios. Especially for normal scenario, either flow rate of HiT or LoT, can give the value lower than the emergency case about 4 times and lead to get low Mach number, noise and also dynamic pressure in Table 19.

Noise in Table 20 is the one of criteria; however, the result is far from the criterion value because of the impact of other limitation such as erosion velocity. By this reason, the result of noise show very low value, especially for the normal flow scenario, the results are 0 dBA for all sources and segments.

Table 18 Mach number of waste gas and the maximum criteria in each segment of pipeline network for all scenario.

Segment	From	To	Mach Number				
			Criteria[a]	Normal		Emergency	
				HiT	LoT	HiT	LoT
1TP12	GSAT1	TP12	0.35	0.018	0.017	0.060	0.060
2TP24	GSAT2	TP24	0.35	0.011	0.011	0.047	0.046
3TP36	GSAT3	TP36	0.35	0.004	0.004	0.018	0.017
4TP43	GSAT4	TP43	0.35	0.009	0.009	0.065	0.064
5TP51	GSAT5	TP51	0.35	0.006	0.006	0.062	0.062
6TP36	GSAT6	TP36	0.35	0.003	0.003	0.033	0.033
BLWKO1G	BLWKO1	BLW	0.35	0.006	0.006	0.056	0.055
MHDR	TP51	BLWKO1	0.35	0.006	0.006	0.055	0.054
TP12-51	TP12	TP51	0.35	0.007	0.007	0.058	0.057
TP24-12	TP24	TP12	0.35	0.007	0.007	0.060	0.059
TP36-43	TP36	TP43	0.35	0.005	0.005	0.042	0.042
TP43-24	TP43	TP24	0.35	0.007	0.007	0.062	0.061

Note: [a] PTTEP PLC. (2010).

Table 19 Dynamic pressure of waste gas and the maximum criteria in each segment of pipeline network for all scenario.

Segment	From	To	Dynamic pressure (Pa)				
			Criteria[a]	Normal		Emergency	
				HiT	LoT	HiT	LoT
1TP12	GSAT1	TP12	50,000	37.43	35.95	547.16	529.03
2TP24	GSAT2	TP24	50,000	15.41	14.80	344.00	333.00
3TP36	GSAT3	TP36	50,000	2.03	1.95	52.44	50.82
4TP43	GSAT4	TP43	50,000	10.19	9.76	691.88	672.71
5TP51	GSAT5	TP51	50,000	4.04	3.86	537.11	520.87
6TP36	GSAT6	TP36	50,000	1.08	1.03	183.02	178.41
BLWKO1G	BLWKO1	BLW	50,000	4.06	3.89	379.48	366.27
MHDR	TP51	BLWKO1	50,000	4.06	3.89	373.06	360.28
TP12-51	TP12	TP51	50,000	6.54	6.27	467.70	452.98
TP24-12	TP24	TP12	50,000	6.78	6.50	536.15	520.44
TP36-43	TP36	TP43	50,000	2.91	2.79	292.10	284.33
TP43-24	TP43	TP24	50,000	6.49	6.21	603.38	586.63

Note: [a] PTTEP PLC. (2010).

Table 20 Noise level of waste gas in pipeline and the maximum criteria at 1 m distance in each segment of pipeline network for all scenarios.

Segment	From	To	Noise (dBA)				
			Criteria	Normal		Emergency	
				HiT	LoT	HiT	LoT
1TP12	GSAT1	TP12	80	0	0	27	27
2TP24	GSAT2	TP24	80	0	0	21	20
3TP36	GSAT3	TP36	80	0	0	1	1
4TP43	GSAT4	TP43	80	0	0	29	29
5TP51	GSAT5	TP51	80	0	0	38	38
6TP36	GSAT6	TP36	80	0	0	30	29
BLWKO1G	BLWKO1	BLW	80	0	0	37	36
MHDR	TP51	BLWKO1	80	0	0	49	48
TP12-51	TP12	TP51	80	0	0	45	44
TP24-12	TP24	TP12	80	0	0	42	42
TP36-43	TP36	TP43	80	0	0	30	30
TP43-24	TP43	TP24	80	0	0	40	40

Table 21 Comparison result of pipeline sizing and the pressure drop across each segment between Aspen Flare System Analyzer and Aspen Plus.

Segment	Connection		Pipeline					
	From	To	ID (mm)		Nominal diameter (in)		Delta P (bar)	
			Flare Analyzer	Aspen	Flare Analyzer	Aspen	Flare Analyzer	Aspen
1TP12	GSAT1	TP12	146	298	6	12	0.180142	0.004880
2TP24	GSAT2	TP24	194	298	8	12	0.057522	0.005913
3TP36	GSAT3	TP36	330	298	14	12	0.005907	0.008137
4TP43	GSAT4	TP43	248	330	10	14	0.046555	0.009463
5TP51	GSAT5	TP51	330	432	14	18	0.123086	0.025224
6TP36	GSAT6	TP36	483	432	20	18	0.062099	0.101721
BLWKO1G	BLWKO1	BLW	737	737	30	30	0.014535	0.114973
MHDR	TP51	BLWKO1	737	737	30	30	0.121795	0.114973
TP12-51	TP12	TP51	584	584	24	24	0.094320	0.087625
TP24-12	TP24	TP12	533	533	22	22	0.064505	0.059569
TP36-43	TP36	TP43	483	432	20	18	0.026443	0.040860
TP43-24	TP43	TP24	483	483	20	20	0.057037	0.051186

Table 21 shows the result of pipe sizing with design criteria. When comparing the result to the result of Aspen Plus simulation, the result shows that the pipe size by Aspen Plus is quite bigger than Flare Analyzer in some segment such as source nos : 1, 2, 4, and 5. However, some of segment of Flare Analyzer is bigger than Aspen plus as well. The reason is the criteria of each parameter cannot be specified in Aspen Plus. The software is not designed to sizing the equipment. It is designed to calculate the pressure and flow parameters with thermal effect calculation involvement. Therefore, the optimization by Aspen Plus is not possible to perform. However, the equation of solver is different due to the design of software itself and the types of involved equation. The pressure drop is the result of pipe sizing. Flare Analyzer can minimize the size of pipeline and maximize the pressure drop across the network with atmospheric pressure at target location.

In the simulation, the variation of temperature is considered as well. The constraint at the low flow rate with low temperature leads to have the condensation while at the high flow rate with high temperature leads get over velocity at the same temperature. Tables 22 and 23 show the result of pressure drop, dynamic pressure, and noise. There is no any parameter, which is higher than the maximum criteria. It means that the simulated pipe size is large enough to handle the flow capacity in case operating temperature change by either gas sources or ambient temperature.

Table 22 Gas flow rate and pressure drop of waste gas in each segment at various operating temperature conditions.

Segment	From	To	Gas Flow rate (kg/h)				Pressure drop (bar)			
			Normal		Emergency		Normal		Emergency	
			15 ^{o}C	35 ^{o}C	35 ^{o}C	45 ^{o}C	15 ^{o}C	35 ^{o}C	35 ^{o}C	45 ^{o}C
1TP12	GSAT1	TP12	423	423	1,794	1,794	0.0138	0.0155	0.1801	0.1852
2TP24	GSAT2	TP24	475	475	2,558	2,558	0.0028	0.0032	0.0575	0.0591
3TP36	GSAT3	TP36	502	502	2,999	2,999	0.0003	0.0003	0.0059	0.0061
4TP43	GSAT4	TP43	633	633	6,074	6,074	0.0008	0.0008	0.0466	0.0478
5TP51	GSAT5	TP51	712	712	8,769	8,769	0.0012	0.0013	0.1231	0.1268
6TP36	GSAT6	TP36	791	791	12,129	12,129	0.0005	0.0006	0.0621	0.0638
BLWKO1G	BLWKO1	BLW	3,363	3,537	34,324	34,324	0.0002	0.0002	0.0145	0.0150
MHDR	TP51	BLWKO1	3,537	3,537	34,324	34,324	0.0017	0.0019	0.1218	0.1259
TP12-51	TP12	TP51	2,825	2,825	25,555	25,555	0.0016	0.0018	0.0943	0.0972
TP24-12	TP24	TP12	2,402	2,402	23,761	23,761	0.0010	0.0011	0.0645	0.0663
TP36-43	TP36	TP43	1,293	1,293	15,128	15,128	0.0003	0.0004	0.0264	0.0272
TP43-24	TP43	TP24	1,926	1,926	21,202	21,202	0.0007	0.0008	0.0570	0.0586

Table 23 Dynamic pressure and noise of waste gas in each segment at various operating temperature conditions.

Segment	From	To	Dynamic pressure (Pa)				Noise (dBA)			
			Normal		Emergency		Normal		Emergency	
			15 ^{o}C	35 ^{o}C	35 ^{o}C	45 ^{o}C	15 ^{o}C	35 ^{o}C	35 ^{o}C	45 ^{o}C
1TP12	GSAT1	TP12	33.56	37.43	547.16	561.88	-	-	27	28
2TP24	GSAT2	TP24	13.82	15.41	344.00	352.86	-	-	21	21
3TP36	GSAT3	TP36	1.82	2.03	52.44	53.72	-	-	1	1
4TP43	GSAT4	TP43	9.15	10.19	691.88	709.10	-	-	29	29
5TP51	GSAT5	TP51	3.63	4.04	537.11	552.66	-	-	38	38
6TP36	GSAT6	TP36	0.97	1.08	183.02	187.51	-	-	30	30
BLWKO1G	BLWKO1	BLW	3.46	4.06	379.48	392.03	-	-	37	37
MHDR	TP51	BLWKO1	3.64	4.06	373.06	385.18	-	-	49	49
TP12-51	TP12	TP51	5.87	6.54	467.70	481.25	-	-	45	45
TP24-12	TP24	TP12	6.09	6.78	536.15	550.59	-	-	42	43
TP36-43	TP36	TP43	2.61	2.91	292.10	299.38	-	-	30	30
TP43-24	TP43	TP24	5.83	6.49	603.38	618.91	-	-	40	41

3. Process design of waste gas treatment system

3.1 Process description

For the process of gas treatment, it can be divided in 3 sub-processes; gas pressure booster, dehydration and cryogenic separation as shown in Figures 37 and 38. Waste gas from pipeline network is routed to gas treatment plant. The design pressure of inlet gas of the system is 0.5-1.0 barg. The limitation depends on the design inlet pressure of 1st booster. In case of over pressure, some amount of gas is relieved to flare system of treatment gas process.

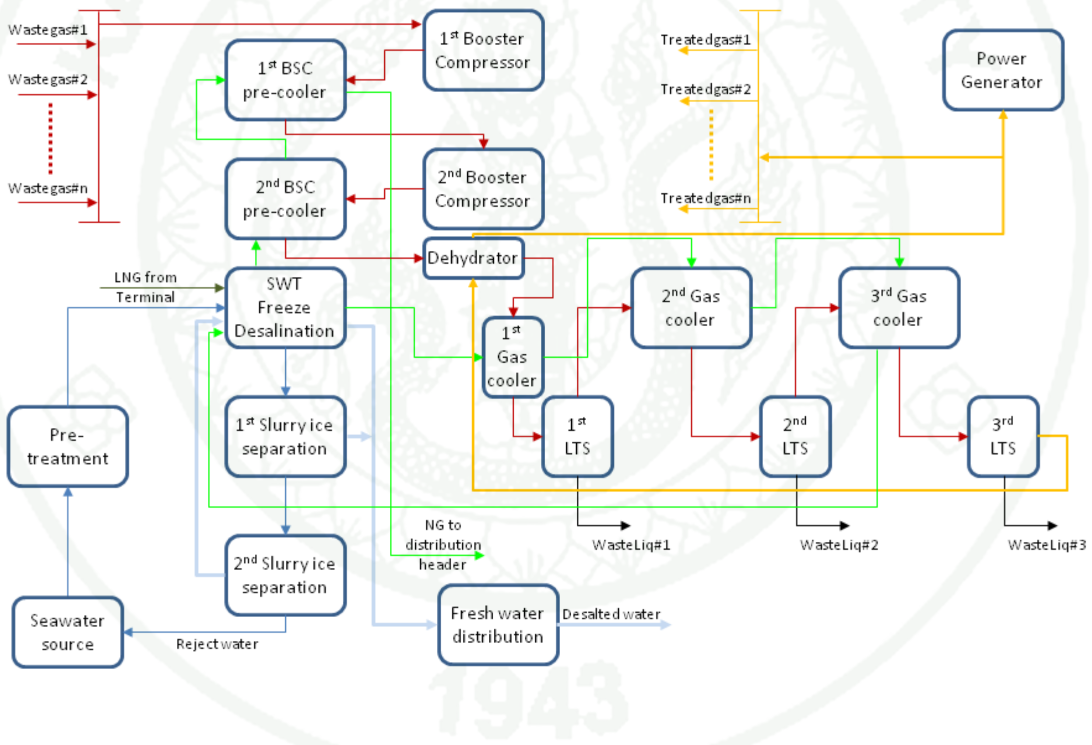


Figure 37 Process flow diagram of waste gas treatment system.

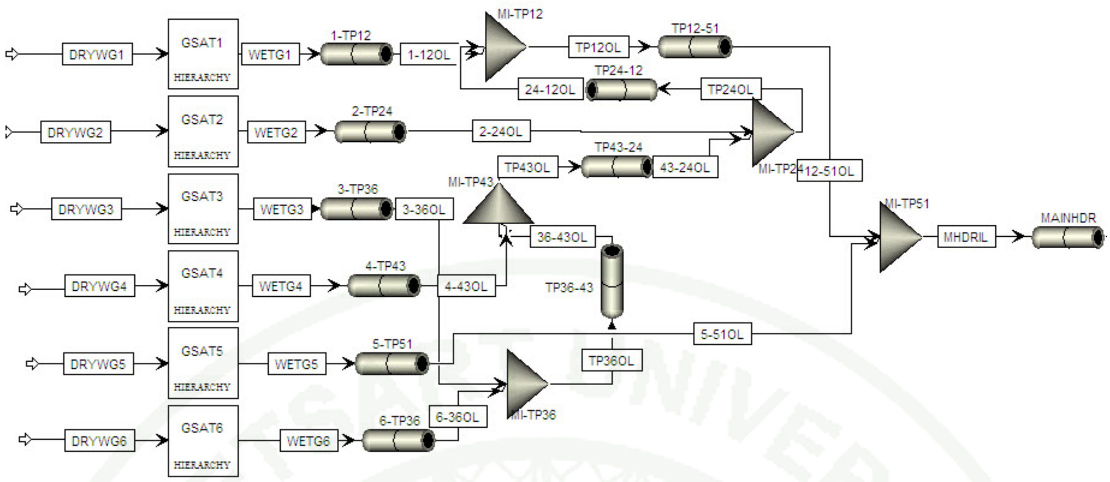


Figure 38 Simulation model in part of waste gas pipeline network.

In Figure 39, gas is compressed from 0.5 barg to 1 barg by this stage and it is increased to 14 barg by 2nd stage compressor in following stage. The design pressure is limited at 14 barg base on the investment cost. In addition, this pressure condition is the maximum working pressure rating ASME class 150 and it is sufficient for toxic removal process including gas dehydration. The cooling of compressed gas is done by using of recycle treated gas from cryogenic process before further utilization. The control temperature of outlet gas of each stage is 20 ^{o}C.

After gas passing through the compression, the high pressure gas is fed to dehydration unit for saturated water removal. The objective of this unit is to prevent gas hydration or ice formation, which may be occurred in the low temperature process. In this design, the cryogenic separation is used, therefore, the dehydration unit is strongly recommended. In general, water content in outlet gas of dehydration unit should be 1-2 lb per MMSCF, which is suitable the process involved with cryogenic condition. In the design, the minimum temperature of downstream process is -80 ^{o}C, and ice formation gas hydration do not occur at -80 ^{o}C.

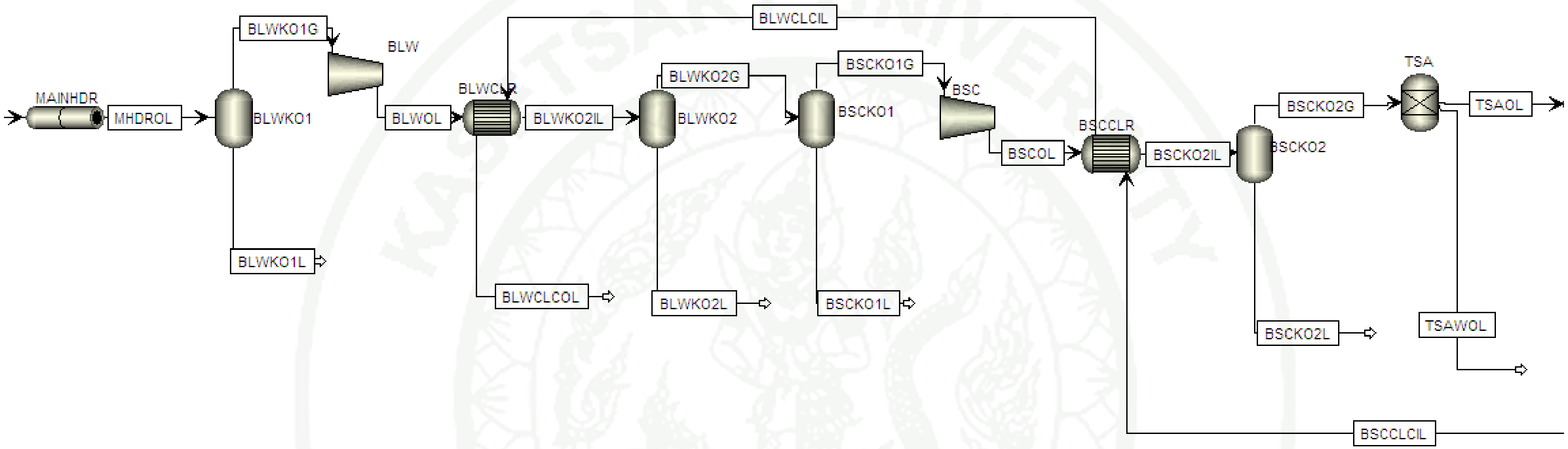


Figure 39 Simulation model for gas pressure booster.

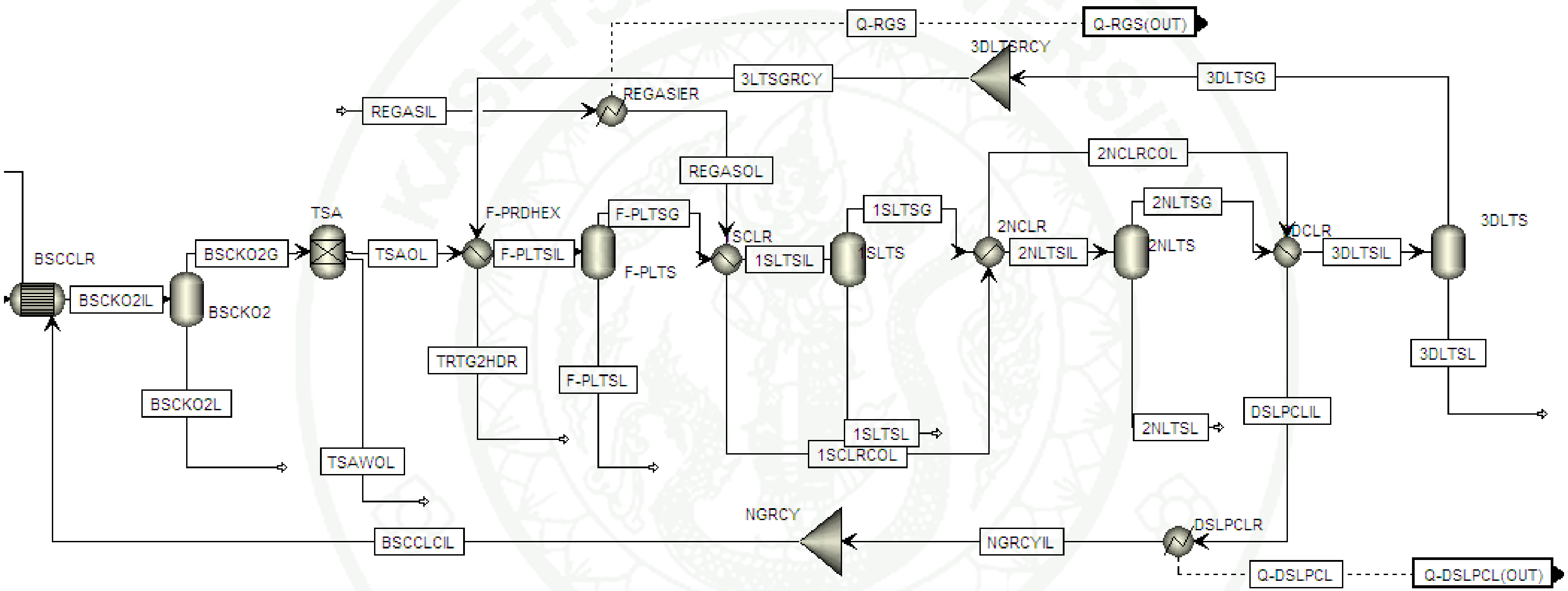


Figure 40 Simulation model for gas treatment process.

The main objective of this integrated system is un-interrupting LNG regasification process purpose. Therefore, the design of LNG to vaporizer and using seawater as heating media was keep as per primary design. This concept also has an advantage covering the phase of modification and construction. In case the re-route of LNG vaporizer by waste gas from the network, the tie-in period and modification would take long time for a year and it's not feasible in aspect of add-on facility concept. Therefore, in the study, LNG is designed to vaporize by seawater, then; NG from the vaporizer will be routed to waste gas treatment process.

Refer to Figure 40, gas from dehydration unit feeds to "Feed-Product cooler" for pre-cooling cryogenic inlet gas and heating-up product treated gas vice versa. Feed gas temperature is decreased to 4 ^{o}C to remove heavy components e.g. BTEX, VOCs and CS_2. LTS technique is selected for this process because the composition of feed gas from waste gas network is not constant as specification. Hence, this method is suitable for wide range of composition type and fluctuation of composition content in feed gas. In addition, this method is also easy for designation, installation, operation and maintenance comparing to other low temperature technique e.g. cryogenic distillation. In view of investment, CAPEX of this method is also lower than other method. However, the disadvantage of this method is the product purity. Since the objective of the design does not aim to get the purified specific component but it aims to separate the toxic component from the waste gas with the opportunity to utilize the treated gas as much as possible. Therefore, other low temperature methods such as low temperature distillation may not be necessary. Cold side stream of this heat exchanger is the treated gas from the 3rd stage LTS. The inlet temperature of cold side is -80 ^{o}C and the outlet is designed to maintain the temperature less than -29 ^{o}C due to the temperature limitation of carbon steel. The temperature depends on the flow rate of each scenario. The average temperature of outlet is about -10 ^{o}C. Although the downstream of this gas is gas generator and/or as purge gas, the temperature is not an issue because the ambient air along the transfer line is adequate for warming up the temperature of this stream.

Gas from F-P cooler (Hot side) is fed to F-P LTS for liquid removal as above explanation. Gas is carried to 1st stage cooler and decrease the gas temperature from 4 to -45 ^{o}C. The objective of this stage is to remove the LPG portion and Vinyl Chloride. The liquid product of this stage is removed by 1st LTS. The liquid product aims to use as fuel for any purpose if there is high VOC content. Therefore, this product must be carefully used. After 1st stage, gas is fed to 2nd stage cooler and 3rd stage; consequently. The design temperature of 2nd stage is -65 ^{o}C. The objective of this stage is to remove the sulfur component such as COS (Carbonyl sulfide) and H_2S. Therefore, the liquid product of this stage must be sent out as disposal for further specific handling. In addition, another objective of this stage is to use as variable stage separation. Actually the concept of the gas treatment in this study is to separate liquid product into 4 groups; 1) Heavy components and VOCs, 2) LPG, 3) Sulfur compounds and 4) Liquid Carbon Dioxide (LCO_2) at atmospheric condition. The waste product group is the first and the third group. The cut point temperature of the first group and the second group is too narrow for adjustment, therefore the setting temperature has to be fixed at -45 ^{o}C but the third group has significant gap. In case there is the waste gas which has boiling point less than the setting temperature (-65 ^{o}C), it can be adjusted (Decreasing) to remove that gas. However, the range of operation must be kept in -65 to -73 ^{o}C.

In the last stage (3rd stage), gas from the 2nd stage is delivered to the 3rd stage to remove CO_2 as liquid for LCO_2 production. Since the CO_2 can be used as feedstock for downstream industry and it also reduces heating value of treat gas and leads to difficulty in engine combustion, these are the reasons for CO_2 removal at the last stage in process. At this stage, the temperature must be reduced to -80 ^{o}C. After the 3rd stage, the main composition in gas is CH_4, C_2H_8, N_2 and H_2 therefore; gas can be used for combustion without any toxic component. Gas from the 3rd stage is fed to F-P cooler again. Gas from F-P cooler is heated up and fed to gas distribution network system for further utilization.

In view of maximization energy recovery efficient, the designed route of cold NG from vaporizer is not the best configuration. The cold NG from vaporizer should feed to 3rd stage heat exchanger as primary recovery, then feed to 2nd stage heat exchanger and 1st stage heat exchanger, respectively. However, the objective of this project is to reduce the emission, especially VOCs, which is designed to remove in 1st stage LTS. For CO_2, the removal of this component is optional. If there is excess energy to remove, it is the advantage for remove it. The result shows that there is excess energy for from the process and likely that the first stage of cold NG feed is the 3rd heat exchanger. In future, the waste gas flow rate may increase by the combination of waste gas from flare header to the waste gas treatment system. It would be a better control performance for VOCs removal if the situation is occurred.

3.2 Simulation result

The result is separated to 2 parts; treated gas product and by-product liquid. In gas part, compositions of waste gas are classified to atmospheric gas, hydrocarbon, LPG, BTEX, Sulphur compound and VOCs. The summation of BTEX, Sulphur and VOCs so-called "Toxic gas". Table 24 shows that the recovery efficiency of BTEX, sulphur compound and VOCs is more than 99%. Especially VOCs, which is the main target in the study, the process can remove them from the waste gas = 100% although LNG feed capacity varies since the minimum capacity (turn-down operation) until maximum capacity. Moreover, at the maximum feed of waste gas as the emergency flow rate condition and high concentration of toxic content in waste gas, the recovery efficiency still closes to 100% for toxic gas. It means that the waste energy is adequate for treating the waste gas. However, sulphur component such as H_2S is still remaining in the treat gas because the condition of the 3rd stage LTS is not low enough to remove all of H_2S. Despite H_2S still remains in the treated gas, the quality of treated gas meets the minimum screening air pollution shown in Table 25. In the future, if 100% H_2S removal is strictly required, the operating temperature of the 3rd stage LTS can be adjusted to reach such a requirement but the feed capacity of waste gas through this system is reduced.

Table 24 Recovery efficiency of the waste treatment process between feed waste gas and treated gas for all scenarios.

Component	LNG feed (MTPA)	Recovery efficiency (%)			
		Normal flow		Emergency flow	
		HiT	LoT	HiT	LoT
Methane	3	11.89	11.88	11.89	11.88
	5	11.89	11.88	11.89	11.88
Ethane	3	76.58	76.64	76.58	76.65
	5	76.58	76.64	76.58	76.65
Propane	3	99.43	99.43	99.43	99.43
	5	99.43	99.43	99.43	99.43
n-butane & i-butane	3	99.99	99.99	99.99	99.99
	5	99.99	99.99	99.99	99.99
Pentane Plus	3	100.00	100.00	100.00	100.00
	5	100.00	100.00	100.00	100.00
LPG	3	99.65	99.65	99.65	99.65
	5	99.65	99.65	99.65	99.65
BTEX	3	100.00	100.00	100.00	100.00
	5	100.00	100.00	100.00	100.00
VOC	3	100.00	100.00	100.00	100.00
	5	100.00	100.00	100.00	100.00
Sulfur Compound	3	96.56	94.98	96.60	95.02
	5	96.56	94.98	96.60	95.02
Carbon dioxide	3	55.90	55.60	55.91	55.61
	5	55.90	55.60	55.91	55.61
Total Toxic Gas	3	99.39	99.26	99.41	99.31
	5	99.39	99.26	99.41	99.31

When comparing the level of toxic content to the air pollution standard as shown in Table 25, the result shows that the VOCs content in the gas is lower than the minimum screening level of VOCs in air. Therefore, this treated gas is able to combust as fuel or burnt at flare stack with incomplete combustion; then, the remaining of VOCs contain in air environment will be at the low concentration level. VOCs are also lower than the simulation result because the concentration of each gas is dispersed by air and the concentration of these VOCs is kept in low level.

The conditions and properties of product; treated gas, it is likely the same for all scenarios but the compositions contented in treated gas are a little different and insignificant impacting on the usage as fuel gas (Table 26). The pressure of the product gas is designed to supply as fuel gas for power generation unit either reciprocating type or turbine type. In general, the supply pressure of fuel gas to power generator is approximate 160-200 psig or 11-13.5 barg. Therefore, product gas is able to supply to power generation unit, directly.

Table 25 The VOC content comparison between gas outlet for waste gas treatment process and screening level standard of air pollution based on 5 MTPA LNG scenario.

Composition	Minimum Screening level	Quantity level (ug/m3)							
		Treated gas				Deviation under screening level			
		Normal flow		Emergency flow		Normal flow		Emergency flow	
		HiT	LoT	HiT	LoT	HiT	LoT	HiT	LoT
BENZENE	0.25	1.75E-15	1.15E-15	1.76E-15	1.14E-15	0.25	0.25	0.25	0.25
TOLUENE	95.8	0	0	0	0	95.8	95.8	95.8	95.8
P&M-XYLENE	208	0	0	0	0	208	208	208	208
O-XYLENE	730	0	0	0	0	730	730	730	730
VINYL-CHLORIDE	0.22	1.51E-08	1.12E-08	1.51E-08	1.12E-08	0.22	0.22	0.22	0.22
1,2-DICHLOROETHANE	0.074	0	0	0	2.38E-16	0.074	0.074	0.074	0.074
CHLOROFORM	0.084	4.92E-14	1.35E-14	4.93E-14	1.35E-14	0.084	0.084	0.084	0.084
CARBONYL-SULFIDE	0.8	2.09E-06	1.67E-06	2.09E-06	1.67E-06	0.8	0.8	0.8	0.8
MTBE	3.7	2.81E-14	1.92E-14	2.82E-14	1.92E-14	3.7	3.7	3.7	3.7
MEK	390	0	0	0	0	390	390	390	390
MIK	205	0	0	0	0	205	205	205	205
ACETONE	370	1.57E-13	3.23E-13	1.58E-13	3.23E-13	370	370	370	370
CARBON-DISULFIDE	3	7.98E-12	4.34E-12	8.02E-12	4.36E-12	3	3	3	3
ETHYL-CHLORIDE	4.1	1.61E-10	5.70E-11	1.61E-10	5.70E-11	4.1	4.1	4.1	4.1
ETHYLBENZENE	200	0	0	0	0	200	200	200	200
DICHLOROMETHANE	2.3	4.38E-12	1.51E-12	4.38E-12	1.51E-12	2.3	2.3	2.3	2.3
STYRENE	11	0	0	0	0	11	11	11	11
TETRACHLOROETHYLENE	0.33	0	0	0	0	0.33	0.33	0.33	0.33
TRICHLOROETHYLENE	0.017	0	0	0	0	0.017	0.017	0.017	0.017
HYDROGEN-SULFIDE	1	7.91E-06	8.20E-06	7.91E-06	8.20E-06	1	1	1	1

Even though the gas temperature is quite low but it is acceptable for pipe routing with carbon steel because the design temperature of carbon steel is -29 ^oC. In view of supply temperature to power generator, it is quite low compared to the standard gas supply temperature (15 ^oC). In this case, however, heat transfer from ambient temperature along gas transfer line to utility system is enough to heat up gas temperature to 15 ^oC.

Table 26 Operating condition and basic physical properties of treated gas for all scenarios.

Parameter	Unit	LNG feed (MTPA)	Scenario			
			Normal flow		Emergency flow	
			HiT	LoT	HiT	LoT
Volume flow	MMSCFD	3	1.12	1.14	10.78	10.92
		5	1.12	1.14	10.78	10.92
Mass flow	kg/h	3	1,038.43	1,052.15	9,965.49	10,097.14
		5	1,038.43	1,052.15	9,965.49	10,097.14
Temperature	^oC	3	- 23.93	- 28.59	- 23.93	- 28.59
		5	- 23.93	- 28.59	- 23.93	- 28.59
Pressure	Barg	3	14	14	14	14
		5	14	14	14	14
Density	kg/m^3	3	14.00	14.31	14.00	14.31
		5	14.00	14.31	14.00	14.31
MW	-	3	18.56	18.57	18.56	18.57
		5	18.56	18.57	18.56	18.57

For the quality of gas, the major components in the gas are methane, ethane and hydrogen. At the level, gas heating value and Methane Number (MN) of gas can be used as fuel gas because carbon dioxide is only 9% approximately (Tables 27 and 28). At this composition level, MN is around 80 and Wobbe Index (WI) is about 33-34 MJ/m^3 for all scenarios. Although the WI of gas is lower than the conventional NG but it can be used as medium grade of fuel gas. And at this level of MN, there is no engine knocking occurring during gas engine operation as per California Air Resource Board (CARB) engine specification.

Table 27 Mole fraction of component in treated gas for all scenarios.

Composition	3MTPA				5MTPA			
	Normal flow		Emergency flow		Normal flow		Emergency flow	
	HiT	LoT	HiT	LoT	HiT	LoT	HiT	LoT
WATER	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CARBON-DIOXIDE	9.39	9.45	9.39	9.45	9.39	9.45	9.39	9.45
METHANE	47.03	47.01	47.03	47.01	47.03	47.01	47.03	47.01
ETHANE	7.47	7.45	7.47	7.45	7.47	7.45	7.47	7.45
PROPANE	0.07	0.07	0.07	0.07	0.07	0.07	0.07	0.07
ISOBUTANE	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
N-BUTANE	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
2-METHYL-BUTANE (i-C_5)	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
N-PENTANE	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
N-HEXANE	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HYDROGEN	21.23	21.22	21.23	21.22	21.23	21.22	21.23	21.22
HYDROGEN-SULFIDE	0.29	0.30	0.29	0.30	0.29	0.30	0.29	0.30
NITROGEN	4.16	4.16	4.16	4.16	4.16	4.16	4.16	4.16
CARBON- MONOXIDE	10.33	10.32	10.33	10.32	10.33	10.32	10.33	10.32
BENZENE	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
TOLUENE	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
P-XYLENE	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
M-XYLENE	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
O-XYLENE	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
VINYL-CHLORIDE	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
1,2-DICHLORO ETHANE	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CHLOROFORM	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CARBONYL-SULFIDE	0.02	0.02	0.02	0.02	0.02	0.02	0.02	0.02
METHYL-TERT-BUTYL-ETHER (MTBE)	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
METHYL-ETHYL-KETONE (MEK)	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
METHYL-ISOBUTYL-KETONE (MIK)	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
ACETONE	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CARBON-DISULFIDE	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
ETHYL-CHLORIDE	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
ETHYLBENZENE	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
DICHLOROMETHANE	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
STYRENE	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
TETRACHLORO-ETHYLENE	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
TRICHLORO ETHYLENE	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
TOTAL	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00

Table 28 Individual component mass flow rate of inlet waste gas and treated gas and individual component removal efficiency of all scenarios.

Composition	Normal flow						Emergency flow					
	Inlet waste gas		Treated gas		Efficiency		Inlet waste gas		Treated gas		Efficiency	
	HiT	LoT	HiT	LoT	HiT	LoT	HiT	LoT	HiT	LoT	HiT	LoT
WATER	52.26	54.20	-	-	100	100	360.93	361.56	-	-	100	100
CARBON-DIOXIDE	524.39	530.75	231.25	235.67	55.90	55.60	5,033.80	5,094.85	2,219.23	2,261.64	55.91	55.61
METHANE	479.24	484.88	422.28	427.29	11.89	11.88	4,599.31	4,653.51	4,052.46	4,100.54	11.89	11.88
ETHANE	536.93	543.23	125.76	126.91	76.58	76.64	5,154.43	5,215.13	1,206.93	1,217.96	76.58	76.65
PROPANE	311.39	315.04	1.76	1.81	99.43	99.43	2,992.23	3,027.84	16.93	17.36	99.43	99.43
ISOBUTANE	134.03	135.70	0.01	0.01	99.99	99.99	1,290.25	1,306.78	0.11	0.11	99.99	99.99
N-BUTANE	65.79	66.62	0.00	0.00	100	100	634.55	642.79	0.02	0.02	100	100
2-METHYL-BUTANE (i-C_5)	76.36	77.51	0.00	0.00	100	100	741.12	752.91	0.00	0.00	100	100
N-PENTANE	73.28	74.41	0.00	0.00	100	100	714.17	726.03	0.00	0.00	100	100
N-HEXANE	67.25	69.08	0.00	0.00	100	100	671.16	690.79	0.00	0.00	100	100
HYDROGEN	24.10	24.38	23.95	24.24	0.61	0.61	231.31	234.02	229.88	232.60	0.62	0.61
HYDROGEN-SULFIDE	40.34	41.65	5.47	5.76	86.44	86.17	387.49	400.17	52.51	55.29	86.45	86.18
NITROGEN	66.98	67.76	65.25	66.03	2.58	2.55	642.78	650.31	626.19	633.69	2.58	2.56
CARBON-MONOXIDE	167.43	169.39	161.86	163.75	3.33	3.33	1,606.76	1,625.61	1,553.30	1,571.51	3.33	3.33
BENZENE	45.46	19.37	0.00	0.00	100	100	467.15	201.23	0.00	0.00	100	100
TOLUENE	24.03	110.38	-	-	100	100	259.09	1,207.11	-	-	100	100
P-XYLENE	10.72	7.53	-	-	100	100	119.52	85.16	-	-	100	100
M-XYLENE	10.37	7.26	-	-	100	100	115.75	82.30	-	-	100	100
O-XYLENE	8.50	4.28	-	-	100	100	95.53	48.93	-	-	100	100
VINYL-CHLORIDE	71.54	49.33	0.01	0.00	99.99	99.99	689.37	475.72	0.05	0.04	99.99	99.99
1,2-DICHLOROETHANE	49.35	272.91	-	-	100	100	515.26	2,889.86	-	0.00	100	100
CHLOROFORM	93.19	17.40	0.00	0.00	100	100	934.38	175.88	0.00	0.00	100	100
CARBONYL-SULFIDE	70.80	56.72	0.82	0.66	98.84	98.83	680.27	545.14	7.88	6.37	98.84	98.83

Table 28 (Continued)

Composition	Normal flow						Emergency flow					
	Inlet waste gas		Treated gas		Efficiency		Inlet waste gas		Treated gas		Efficiency	
	HiT	LoT	HiT	LoT	HiT	LoT	HiT	LoT	HiT	LoT	HiT	LoT
METHYL-TERT-BUTYL-ETHER (MTBE)	78.80	36.84	0.00	0.00	100	100	777.12	363.99	0.00	0.00	100	100
METHYL-ETHYL-KETONE (MEK)	39.46	10.18	-	-	100	100	407.82	106.25	-	-	100	100
METHYL-ISOBUTYL-KETONE (MIK)	19.22	8.79	-	-	100	100	210.14	97.23	-	-	100	100
ACETONE	46.42	66.32	0.00	0.00	100	100	464.54	668.15	0.00	0.00	100	100
CARBON-DISULFIDE	72.03	29.58	0.00	0.00	100	100	709.93	293.24	0.00	0.00	100	100
ETHYL-CHLORIDE	70.00	20.88	0.00	0.00	100	100	678.34	202.71	0.00	0.00	100	100
ETHYLBENZENE	12.34	5.50	-	-	100	100	136.89	61.90	-	-	100	100
DICHLOROMETHANE	81.77	21.19	0.00	0.00	100	100	802.89	209.01	0.00	0.00	100	100
STYRENE	8.82	2.06	-	-	100	100	98.86	23.46	-	-	100	100
TETRACHLORO-ETHYLENE	35.87	3.23	-	-	100	100	390.30	35.74	-	-	100	100
TRICHLOROETHYLENE	68.54	12.70	-	-	100	100	710.45	133.17	-	-	100	100

For the by-product liquid, the results in Table 29 show that the quantity is very large number and the amount of product from 1st stage is more than 3rd stage product, 2nd stage product and pre-cooling stage product, respectively. However, it depends on the composition at actual condition. In real operation, the waste gas or excess gas that is routed may affect the main composition in the network.

Table 29 Liquid product characteristic of Pre-cooling stage, 1st LTS, 2nd stage LTS and 3rd stage LTS based on 5 MTPA LNG feed scenario.

Product type	Parameter	Unit	Scenario			
			Normal flow		Emergency flow	
			HiT	LoT	HiT	LoT
Pre-cooling stage product	Flow rate	kg/h	136.9	97.43	1313.37	934.9
	Temperature	°C	4	4	4	4
	Pressure	Barg	14	14	14	14
	Density	kg/m^3	738.6	681.3	2.684	681.5
	HC Dew point	°C	119.8	126.3	120	126.3
	HHV	BTU/scf	3172.488	3210.064	3172.48	3210.064
1st stage product	Flow rate	kg/h	725.2	734.6	6955.078	7050
	Temperature	°C	-45	-45	-45	-45
	Pressure	Barg	14	14	14	14
	Density	kg/m^3	652.1	633	1.827	633
	HC Dew point	°C	52.17	52.52	52.18	52.52
	HHV	BTU/scf	74.96	2024.5412	2011.9264	2024.5412
2nd stage product	Flow rate	kg/h	275.8	285.9	2646.62	2744
	Temperature	°C	-65	-65	-65	-65
	Pressure	Barg	14	14	14	14
	Density	kg/m^3	632	627.7	1.509	627.7
	HC Dew point	°C	3.112	2.691	3.111	2.691
	HHV	BTU/scf	1434.598	1439.6976	1434.598	1439.6976
	Wobbe Index	BTU/scf	1292.8828	1300.6664	1179.4667	1300.6664
3rd stage product	Flow rate	kg/h	293.1	301.6	2813.118	2894
	Temperature	°C	-80	-80	-80	-80
	Pressure	Barg	14	14	14	14
	Density	kg/m^3	666.4	667.1	1.419	667.1
	HC Dew point	°C	-26.14	-26.47	-26.14	-26.47
	HHV	BTU/scf	1065.8164	1060.7168	1065.8164	1060.7168
	Wobbe Index	BTU/scf	990.396	985.8332	903.63134	985.8332

The HHV (High Heating Value) of the pre-cooling stage is higher than the other product because of the heavy hydrocarbon base e.g. BTEX. However, it is not suitable to use the product as liquid fuel directly because there are other substances such styrene and other VOCs contained in this product. The re-processing of this product or blending with fuel oil for further application is more suitable than disposal because the number of this product is quite large (approximately 140 kg/h). Therefore, it leads to give the expensive cost for transportation. In addition, improper handling product storage may be cause the problem of disposal area such as spillage or high VOCs vaporization.

Since the process of gas treatment is to compress the gas before sending to cryogenic process, therefore, the pressure of liquid product is 14 barg as well. At this pressure, gas or light gas is compressed in liquid product. Once the pressure is released to atmospheric condition, only the expected product is kept as liquid phase in the product. Therefore, all liquid product pressure must be reduced to atmospheric before unloading. For the 1st stage product, H_2S is released out from LPG portion. For 2nd stage product, the most of LPG releases from liquid sulfur compound. Therefore, the monitoring of liquid sulfur could be considered as dangerous flammable substance. The 3rd stage substance, light hydrocarbon e.g. ethane released from LCO_2. However, some H_2S may contaminate in the product, the monitoring is required for this product. After decreasing the pressure, some liquid product can be used as fuel or some light gas can be used as fuel gas e.g. light gas from 3rd stage product. The reason is the sufficient amount of HHV of the liquid for engine combustion. The properties of all liquid by-products are detailed in Tables 30 and 31.

Table 30 Liquid product mass flow rate and volumetric flow rate of Pre-cooling stage, 1st LTS, 2nd stage LTS and 3rd stage LTS.

Product type	LNG feed (MTPA)	Mass flow rate (kg/h)				Volumetric flow rate (m^3/h)			
		Normal flow		Emergency flow		Normal flow		Emergency flow	
		HiT	LoT	HiT	LoT	HiT	LoT	HiT	LoT
Pre-cooling stage product	3	136.85	105.68	1,313.37	1,014.20	0.18	0.15	1.77	1.45
	5	136.85	105.68	1,313.37	1,014.20	0.18	0.15	1.77	1.45
1st stage product	3	725.13	722.87	6,955.08	6,937.36	1.10	1.14	10.58	10.90
	5	725.13	722.87	6,955.08	6,937.36	1.10	1.14	10.58	10.90
2nd stage product	3	275.79	288.74	2,646.62	2,770.91	0.43	0.45	4.09	4.31
	5	275.79	288.74	2,646.62	2,770.91	0.43	0.45	4.09	4.31
3rd stage product	3	293.13	302.11	2,813.12	2,899.23	0.42	0.43	4.04	4.16
	5	293.13	302.11	2,813.12	2,899.23	0.42	0.43	4.04	4.16

Table 31 Liquid product mass density and molecular weight of Pre-cooling stage, 1st LTS, 2nd stage LTS and 3rd stage LTS for all scenarios.

Product type	LNG feed (MTPA)	Mass density (kg/m3)				Molecular Weight			
		Normal flow		Emergency flow		Normal flow		Emergency flow	
		HiT	LoT	HiT	LoT	HiT	LoT	HiT	LoT
Pre-cooling stage product	3	740.0	697.6	740.3	697.7	61.12	59.94	61.13	59.94
	5	740.0	697.6	740.3	697.7	61.12	59.94	61.13	59.94
1st stage product	3	657.6	636.5	657.7	636.6	42.55	41.87	42.56	41.87
	5	657.6	636.5	657.7	636.6	42.55	41.87	42.56	41.87
2nd stage product	3	647.1	642.4	647.1	642.4	35.37	35.18	35.36	35.18
	5	647.1	642.4	647.1	642.4	35.37	35.18	35.36	35.18
3rd stage product	3	696.3	697.4	696.3	697.4	33.32	33.32	33.32	33.32
	5	696.3	697.4	696.3	697.4	33.32	33.32	33.32	33.32

4. Process design of seawater desalination plant

4.1 Deviation of freezing point and eutectic point between real data and simulated data.

In part of seawater desalination, freeze desalination technique is applied as mention earlier. The key parameter of freeze desalination system is the freezing point and eutectic point. In the study, real seawater from GOT is acquired for preliminary testing on the freezing. However, the comparison of lab test and simulation is required. With the initial, a single solution system (NaCl/Water) was tested by simulation. The result is shown Tables 32 and 33 as below:

Table 32 Simulation result of Freezing point test in salt solution (Water/NaCl).

%Mass in feed			Freezing point (^oC)					
			Operating pressure (barg)					
H2O	**NaCl**	**Total**	**0**	**1**	**3**	**5**	**7**	**10**
100	0	100	0	0	0	0	0	0
99	1	100	-0.9	-1	-0.9	-0.9	-0.9	-0.9
95	5	100	-3.2	-3.2	-3.2	-3.2	-3.2	-3.2
90	10	100	-6.7	-6.7	-6.7	-6.7	-6.7	-6.7
85	15	100	-11.2	-11.2	-11.2	-11.3	-11.2	-11.2
80	20	100	-17.1	-17.1	-17.1	-17.1	-17.1	-17.1
77	23	100	-21.5	-21.5	-21.5	-21.5	-21.5	-21.5
76.7	23.3	100	-29.5	-29.5	-29.5	-29.3	-29.6	-29.7

Table 33 Simulation result of Eutectic point test in salt solution (Water/NaCl).

%Mass in feed			Eutectic point (^{o}C) Operating pressure (barg)					
H2O	**NaCl**	**Total**	**0**	**1**	**3**	**5**	**7**	**10**
100	0	100	0.0	0.0	0.0	0.0	0.0	0.0
99	1	100	-29.4	-29.4	-29.4	-29.4	-29.4	-29.4
95	5	100	-29.4	-29.4	-29.4	-29.4	-29.4	-29.4
90	10	100	-29.4	-29.4	-29.5	-29.4	-29.4	-29.4
85	15	100	-29.4	-29.4	-29.5	-29.4	-29.4	-29.4
80	20	100	-29.4	-29.4	-29.4	-29.4	-29.4	-29.4
77	23	100	-29.4	-29.4	-29.5	-29.4	-29.4	-29.4
76.7	23.3	100	-29.4	-29.5	-29.5	-29.3	-29.6	-29.7

The results show that the freezing point of the NaCl/Water system start at 0 ^{o}C with the absent of NaCl. During increasing of the concentration of NaCl from 0 to 23.3 %wt., the simulation shows that the trend of freezing point of eutectic solution reduced until -29.4 ^{o}C and it did not change under pressure. In general, the eutectic point of NaCl/Water system is -21.1 ^{o}C at 101.13 kPa but in the simulation the result shows the eutectic point at 29.4 ^{o}C (the deviation = 8.3 ^{o}C). This deviation is quite wide band for using as reference for design the system. The reason of this deviation may be divided in 3 possibilities:

1. The dissolve gas is saturated in water before its phase change such as CO_2 or O_2.

2. The library of electrolyte pair-coefficient data from Aspen Plus has some tolerance of error.

3. The actual of chemical reaction and component involved in the eutectic solution system are different.

These reasons influence to the result deviation. However, to handling these reasons are difficult and may not be possible because of the chemical reaction of each component in real seawater. There are many kinds of composition and radical in the solution though, the acceptable method is to compare the real freezing point to the simulation.

4.2 Effect of component on freezing point and eutectic point

In general, the contaminant in pure substance leads to higher boiling point or lower freezing. Figure 41 shows that the different of freezing point and eutectic point between NaCl/Water system and Sea-salt/seawater system. The freezing point begins at the same temperature (-0.8 ^{o}C and -1 ^{o}C) but once the concentration of salt in seawater increases until eutectic point, the eutectic temperature is lower than the eutectic point of NaCl/Water system. This characteristic of seawater in the simulation aligns to the theory as mentioned above and the discussion in section 4.1.

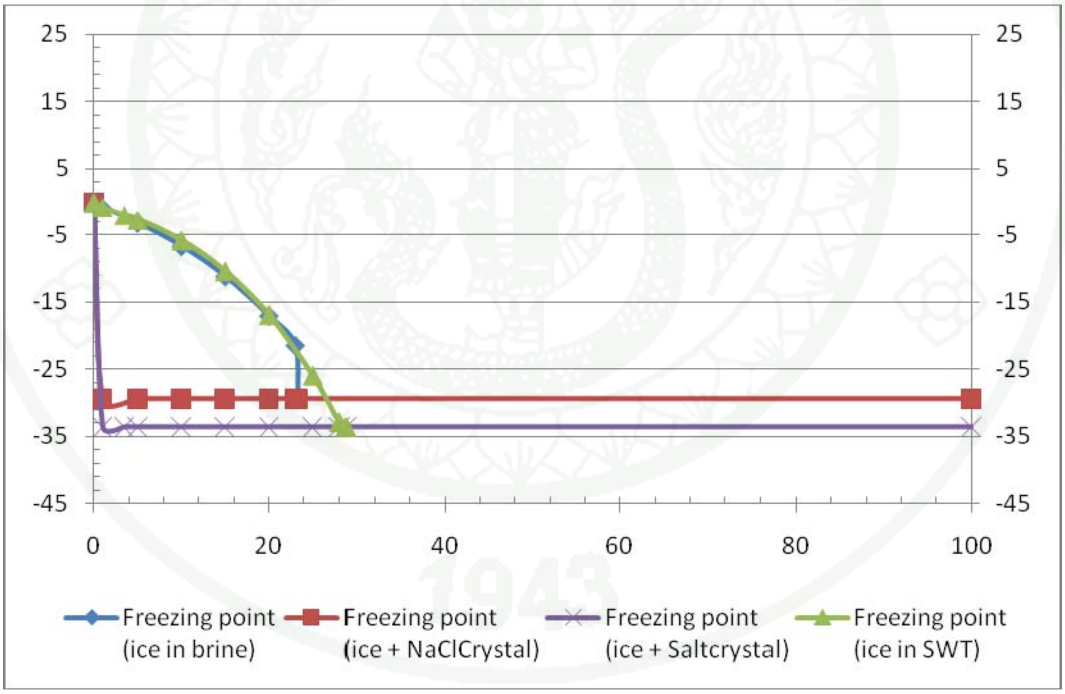


Figure 41 Freezing temperature curve and eutectic temperature curve of NaCl/Water system and sea-salt/seawater system.

Moreover, the comparison of the eutectic point of other eutectic solution systems such as NaCl/KCl/Water, NaCl/Na_2SO_4/Water, $CaCl_2$/Water, $MgCl_2$/Water, NaCl/$CaCl_2$/Water and NaCl/$MgCl_2$/Water with the reference in Table 9. The results show that the eutectic point at different electrolyte system is also different significantly. In addition, a kind of salt affects to the eutectic point of the electrolyte system as well. Example, NaCl-water system, the eutectic point of the system = -21.2 ^oC but if $MgCl_2$ was added in to this system, the eutectic point of the system will reduced to -35 ^oC. Hence, the eutectic point of the electrolyte system is affected by the composition in the system. Table 34 shows that different of eutectic point between actual data and simulated data is unpredictable. In some system; e.g. $CaCl_2$/Water, the eutectic temperature of simulation is lower than the actual test. Therefore, the simulation of salt solution by electrolyte feature in some specific software can be performed if there is lab test result compared to the simulation result.

Table 34 The comparison between simulation result and laboratory test result from reference on freezing point and eutectic point

Scenario	Temperature (^oC)											
	1		2		3		4		5		6	
Component	**Ref.**	**Sim.**	**Ref.**	**Sim.**	**Ref.**	**Sim.**	**Ref.**	**Sim.**	**Ref.**	**Sim.**	**Ref.**	**Sim.**
NaCl(S)	-21.2	-29.5	-22.9	-30.7						-52.6	-35	-49.5
KCl(S)				-6.8								
$KHCO_3$(S)												
$CaCl_2$(S)					-49.8				-55			
$CaCO_3$(S)												
$MgCl_2$(S)							-33.6	-60.1				-38.7
$MgSO_4$(S)												
H_2O (ICE)(S)		-22.1		-22.4		-38.5		-45		-40.4		-45.9
Eutectic Temperature	-21.2	-29.5	-22.9	-30.7	-49.8	-38.5	-33.6	-60.1	-55	-52.6	-35	-49.5

Note: "Ref." stands for reference data.

"Sim." stands for data from simulation.

4.3 Comparison of freezing point between laboratory test and simulated data.

By performing freezing point of seawater sample from GOT with petroleum pour point test, the results are in Table 35.

Table 35 Freezing point measurement of GOT seawater

Run No.	Freezing point (^{o}C)
1	- 16
2	- 16
3	- 16

The result shows that the freezing point starts at - 16 ^{o}C for 3 runs; however, the eutectic point cannot be demonstrated by this method because of 2 reasons below:

1. The size of sample vessel of petroleum pour point test is very small.

2. The temperature controlling performance at water bath in pour point test does not sufficient for freezing point testing.

In general, the temperature at inside vessel wall is equal to the outside wall temperature and the temperature at center of vessel is normally higher than the outside in case of cryogenic condition. If the vessel is designed to have more bigger in diameter, the temperature difference between center of vessel and the wall of vessel would be more than a smaller vessel and it would be easier to detect the freezing point by that equipment. Moreover, if sample vessel size is too small then, the temperature probe cannot be placed and measured; accurately. Therefore, in this study, the initial freezing point of the seawater is adequate for the conceptual process design. Moreover, the operating temperature of the design system is not expected to operate the freeze desalination unit close to eutectic temperature because the salinity of the fresh water product would increase due to the operating temperature goes close to eutectic temperature.

According to the simulation, there are 5 cases of seawater. The concentration of salt in water is varied with constant salt to salt ratio. The concentration is varied from 96.47% water (base case) to 80% water (high salt concentration case).

Table 36 The simulation result of freezing point and eutectic point at various seawater compositions with possible salt in seawater solution

Case	**1**	**2**	**3**	**4**
Reference case	**7[a]**	**8**	**10**	**12**
Component	**Temperature (Deg.C)**			
NaCl(S)	-31.8	-31.8	-31.8	-31.8
KCl(S)	-34.3	-34.3	-34.3	-34.3
$KHCO_3$(S)	-	-	-	-
$CaCl_2$(S)	-	-	-	-
$CaCO_3$(S)	-	-	-	-
$MgCl_2$(S)	-	-	-	-
$MgCl_2$*H_2O	-17.5	-17.5	-17.5	-17.5
$MgCl_2$*4H_2O	-	-	-	-
$MgSO_4$(S)	-	-	-	-
$CaSO_4$	-17.5	-17.5	-17.5	-17.5
$CaSO_4$*1:2W:A	-37	-37	-37	-38.5
ICE(S)	-2.1	-0.9	-5.9	-16.8
Eutectic Temperature	-42.3	-39.8	-40.7	-47.1

Note: [a] means data set of base case.

Table 36 shows that initial freezing point start at -2.1 ^oC (base case) and it gradually decreases while the concentration of water decreases. The initial freezing point is quite reliable because the initial point is the same to NaCl/water system and the trend of initial freezing point in other case is reduced along the concentration of water. This characteristic of initial freezing point is same to NaCl/water system or other eutectic solution phase diagram. As the temperature reduction until -17.5 ^oC, $MgCl_2$ and $CaSO_4$ begin to crystallize for all cases. At this condition, there is a risk of high salinity in fresh water, however, the operating temperature cannot be decided before simulation completed.

The result also shows that the further temperature be decreased to -31.8 ^{o}C, the higher salt component (NaCl) in seawater crystallize. At this temperature, the salinity of water product must not be over the fresh water specification because the NaCl is the salt contained 78.27% wt of total salt in seawater. Then, the operating temperature selection must not less than -31.8 ^{o}C. According to the operating temperature, these 2 conditions are considered because the suitable condition depends on the salinity of the fresh water product. However, the freezing point cannot be used as operating temperature directly because the fluctuation in the actual operation may occur due the temperature controller. The temperature might fluctuate in upside or downside of freezing point. If the temperature is less than the freezing point, the water product salinity will be higher than the specification. Therefore, the range of -16 ^{o}C and -30 ^{o}C are considered as the operating temperature conditions in this study.

4.4 Effect of salinity of seawater to desalination feed supply flow rate

The desalination system feed water supply at different process configuration will be illustrated in this sector. Table 37 is the process used hydrocyclone for ice-seawater separation while Table 38 is the process used ice filter for the separation. The flow rate requirement in either hydrocyclone configuration or filter configuration is the same salinity. However, the different of seawater requirement between low case and high case is 80.2 - 82.8 % by used minimum flow in each LNG feed scenario as a baseline. It is 60.70 - 61.30% of the base case. Although the deviation in flow rate is too wide range but it is not possible to occur because the high case, the salinity is very high and has never been found in GOT. Moreover, it is also more than the conventional salinity of seawater in the world as well. In spite of concentration in thermocline zone or deep zone in GOT, the salinity of seawater of that zone is still lower than the salinity in high case.

Table 37 The feed seawater flow rate for LNG re-gasification with operating temperature = -16 °C at different salt concentration in seawater in the process with hydrocyclone configuration

LNG feed (MTPA)	SWT flow rate (m^3/h@STD)					
	Low salt		Base Case		High salt	
	LoT	HiT	LoT	HiT	LoT	HiT
3	522	565	560	631	687	951
3.5	614	659	656	739	809	1114
4	705	755	752	850	926	1278
4.5	795	853	848	953	1046	1440
5	888	951	943	1061	1165	1600

Table 38 The feed seawater flow rate for LNG re-gasification with operating temperature = -16 °C at different salt concentration in seawater in the process with multi-stage filter configuration

LNG feed (MTPA)	SWT flow rate (m^3/h@STD)					
	Low salt		Base Case		High salt	
	LoT	HiT	LoT	HiT	LoT	HiT
3	522	565	560	631	687	951
3.5	614	659	656	739	809	1114
4	705	755	752	850	926	1278
4.5	795	853	848	953	1046	1440
5	888	951	943	1061	1165	1600

4.5 Process description

In the study, many attempts have been performed in many process designs for seawater desalination section. Since there are many process configurations, they lead to have various alternative designs. It is not only the product specifications and their conditions need to be carefully concerned but also the economic result of each design. However, in the study, there are three reasonable techniques for ice-seawater separation unit:

1. Hydrocyclone
2. Multi-stage filter
3. Hydrocyclone with reverse osmosis combination

And there are 4 techniques for ice slurry melting:

1. Electric heating
2. Sunlight radiation
3. Mixing with make-up water
4. Exchange heat with seawater by heat exchanger

The process diagrams of these configurations/combinations are similar but the advantages and disadvantages process operation are different. The reasons have been described in each part of discussion. The process descriptions of all configurations are illustrated in the following sections:

4.5.1 Hydrocyclone unit with electric heater/sunlight radiation configuration

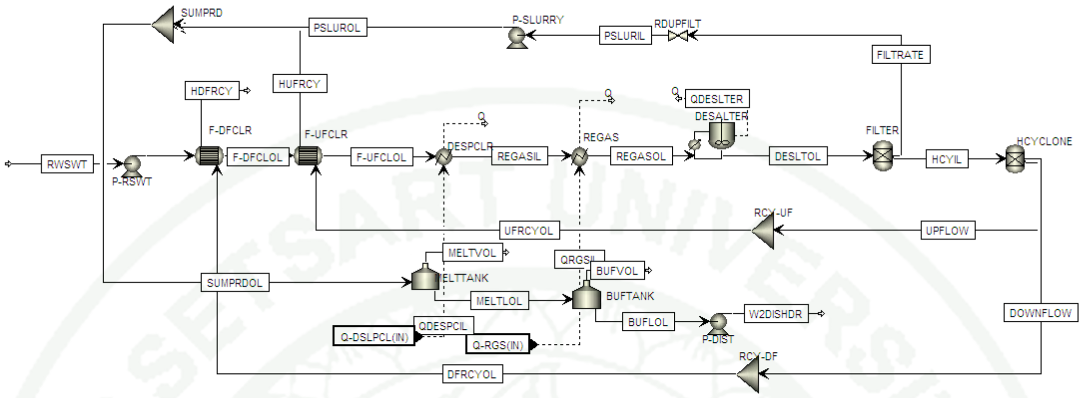


Figure 42 Process simulation model of freeze desalination system with hydrocyclone and electric heater configuration.

Figure 42 shows the detail of this configuration. The process starts with seawater supply line. Seawater feed the seawater storage tank. Pre-filter of junk and mud is filtrated and precipitated at this tank. Seawater is pumped by feed supply pump. The pressure increases from about 1 barg (depends on seawater in storage tank) to 5 barg. This pressure included the pressure drop across the piping line in the system. Seawater is sent to reduce the temperature as pre-cooling stages; F-DFCLR and F-UFCLR before delivered to LNG vaporizer. These units are designed to use upflow (product stream) and downflow (rejected stream) streams; respectively, from hydrocyclone to pre-cool the feed water. In the same time, both product and rejected streams need to be heated up also. The outlet temperature is about 25 and 22.5 ^{o}C; respectively. F-UFCLOL stream is fed to the 3rd stage pre-cooler; DESPCLR. The temperature outlet from this unit is designed about -2 ^{o}C. The cooling side of this unit is NG outlet from 3rd LTS in waste gas treatment plant. At this temperature, crystallization will not be occurred. The crystallization appears after seawater passed through LNG vaporizer unit. The slurry product from LNG vaporizer will be filtered by ice-filter. This filter is designed to separate a main portion of ice slurry (water product) with coarse mesh screen. The remaining slurry fraction is expected to 0.2

then sent to hydrocyclone unit for fine separation purpose. There are 2 outlet streams from hydrocyclone; upflow and downflow streams. Upflow stream is the product containing the most of ice portions in the stream. Downflow stream is the rejected water with the high salinity and less of ice. The range of temperature outlet of these streams is among (-16) – (-30) ^{o}C depending on the operating condition.

Downflow stream is fed back to D-FCLR pre-cooler to heat up the stream from (-16) – (-30) ^{o}C to 30 ^{o}C or equal 5 ^{o}C differ from the rejected seawater temperature. The purpose of this design is to prevent the effect of temperature different between disposal water to marine life in the sea (Phongsuwan, 2007). Upflow stream from hydrocyclone is fed to U-FCLR pre-cooler then mixing with ice slurry from filter before sending to the melting tank. The melting tank or pond is designed to increase the temperature of ice slurry product from about -12 ^{o}C to 0.1 ^{o}C. The reason is to change ice slurry phase to liquid phase before sending to distribution network. The conventional technique selected to apply on this purpose is the electric heating from the electric heater or renewable energy from sunlight radiation, the preliminary assessment comparison of the benefits between these techniques is shown in Table 39. After ice slurry is changed to liquid phase, it is sent to distribution pump designed to 10 barg for further to public storage unit.

Table 39 Preliminary assessment comparison of the benefit between electric heating technique and sunlight radiation technique.

Consideration issue	**Electric heating technique**	**Sunlight radiation technique**
Power consumption	High power consumption.	High power consumption
Power availability	No issue of power shortage the source of power generation is come from the treated gas from waste gas treatment plant	There is problem in night time and cloudy season
Power stability	No problem. In case of power shortage, back up power from PEA can be supplied	There is problem when raining season and cloudy weather condition.

Table 39 (Continued)

Consideration issue	**Electric heating technique**	**Sunlight radiation technique**
Cost of power	High cost	No cost
Installation area	Medium size, however, it depends on the capacity of the melting tank.	Large area is required because open pond with circulation system is required
Temperature controllability	No problems	There is problem because of ambient temperature depending on season
Maintenance cost	Quite high because the corrosion due to the water service condition. In case material is design to water service condition, the investment cost of electric heater will be expensive.	Cost is quite low because the open pond should be designed as concrete.
Investment cost	Since the large capacity and many point of heater are required. So the investment cost of electric heater should be expensive. Moreover, it also depends on the type of material of heating coil.	Significant cheaper than electric heating technique because of source of power and maintenance cost.

4.5.2 Multi-stage filter configuration.

Description in Figure 43 shows the start point is same as the hydrocyclone configuration. Feed seawater is delivered to the storage tank. Pre-filter of junk and mud is filtrated and precipitated at this tank. Seawater will be pumped by the feed supply pump. The pressure can be increased from about 1 barg (depends on seawater in storage tank) to 5 barg. This pressure included pressure drop across the piping line in the system. Seawater is sent to reduce the temperature at 2 stages pre-cooling (F-RJCLR and DESPCLR) before delivered to LNG vaporizer. The outlet temperature of 2nd stage pre-cooler is designed at -2 ^{o}C before fed to LNG vaporizer. The operating temperature is designed at the range of (-16) – (-30) ^{o}C. Slurry product such as ice from vaporizer is fed to the multi-stages filter, which is different from the hydrocyclone configuration.

In Figure 44, the filtrate stream (ice) from ice filter is mixed with make-up fresh water and sent to buffer tank before distributed to further public storage. The purpose of make-up water is to increase the temperature from the range of (-16) – (-30) ^{o}C to 0.1 ^{o}C.

On the other hand, Figure 45 shows that the rejected water stream (concentrated salt stream) passes through F-RJCLR to the blending tank. The cold side outlet temperature of F-RJCLR is around (-6) – (-11) ^{o}C (depending on operating temperature of desalination unit). The outlet from F-RJCLR will mix with make-up seawater to increase temperature of rejected water stream to 30 ^{o}C before disposed to sea. Rejected water will be sent to buffer tank before disposed to sea by rejected pump.

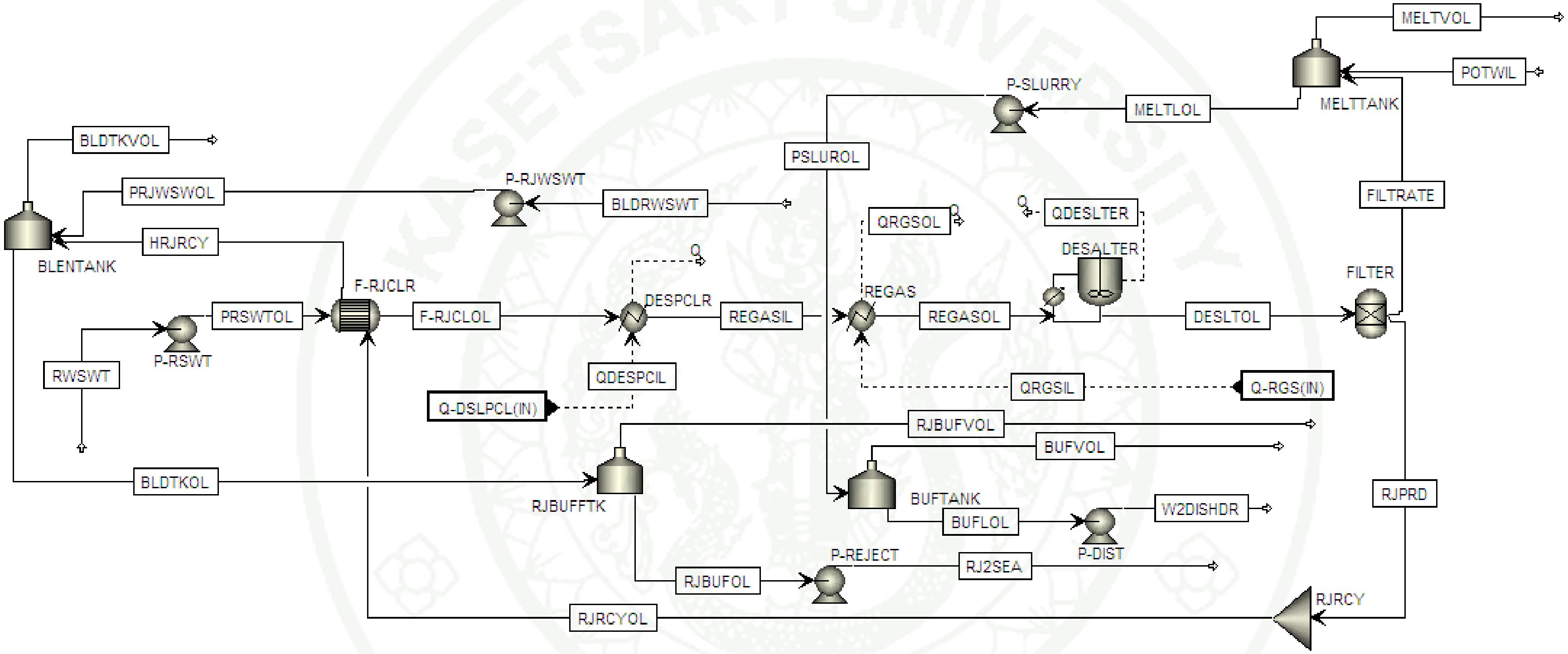


Figure 43 Process simulation model of freeze desalination system with multi-stage filter configuration

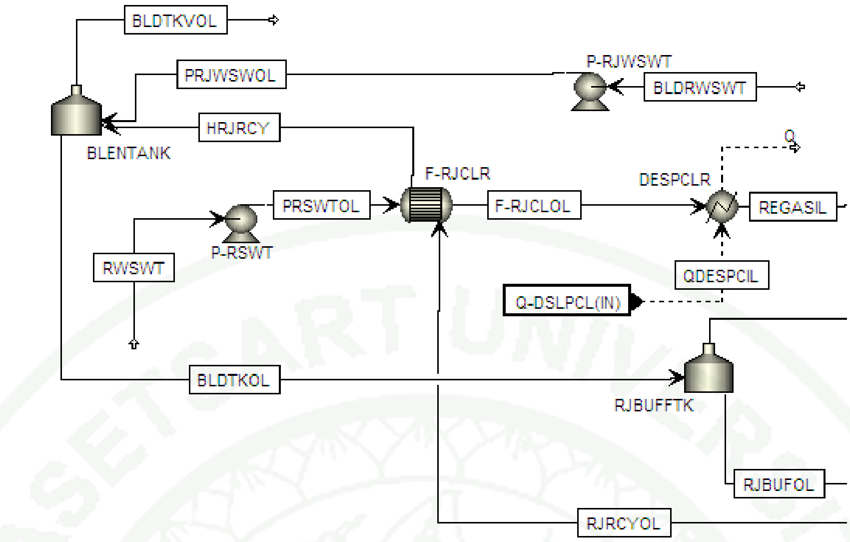


Figure 44 Feed pre-cooling and rejected water treatment section in freeze desalination system with multi-stage filter configuration

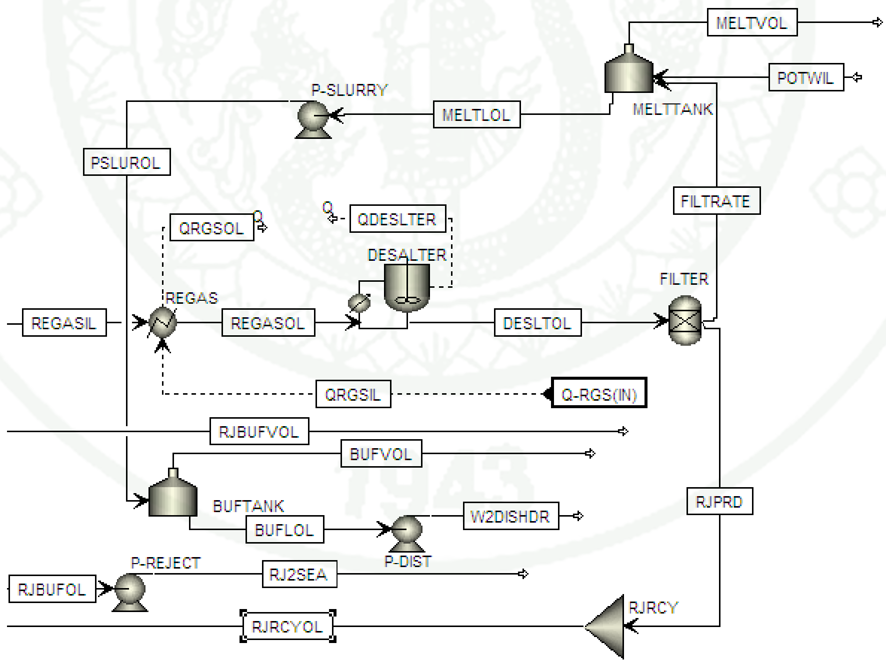


Figure 45 Freeze desalination, fresh water melting and product distribution section in freeze desalination system with multi-stage filter configuration

4.5.3 Hydrocyclone with reverse osmosis combination configuration

The objective of this combination is to produce the drinking water (not only raw water quality). The main process is similar to hydrocyclone configuration but there is RO package to treat upflow stream (Hydrocyclone product) before distribution to community. According to Figure 46, all ice products from PSLUROL and HUFRCY, this product in form of slurry will be sent to heat exchanger (MELTHEX) for melting the ice slurry in the product stream. The hot media for this heat exchanger is the seawater. The design outlet temperature of the product stream is 0.1 ^{o}C, which is the same value to other configurations. Heating outlet stream is sent to mix with the rejected water from RO and downflow from hydrocyclone as well. However, the operating condition of RO is too high, then the high-pressure pump is required to increase the pressure to 80 barg.

Since there is no RO module for the simulation, the combination of flow splitter and component splitter are applied (ROFLOSEP and ROSALSEP). The objective of ROFLOSEP is to perform as the flow recovery of RO unit. In the design, the recovery of flow is designed at 80%. The objective of ROSALSEP is to split the salt content in the product stream to meet the requirement. In this design, TDS is the design criteria and it is designed to less than 200 mg/l. After RO unit, both product water and rejected water will be sent to buffer tanks (BUFTANK and RJBUFFTK) before distribute to communities or disposal rejected water to sea.

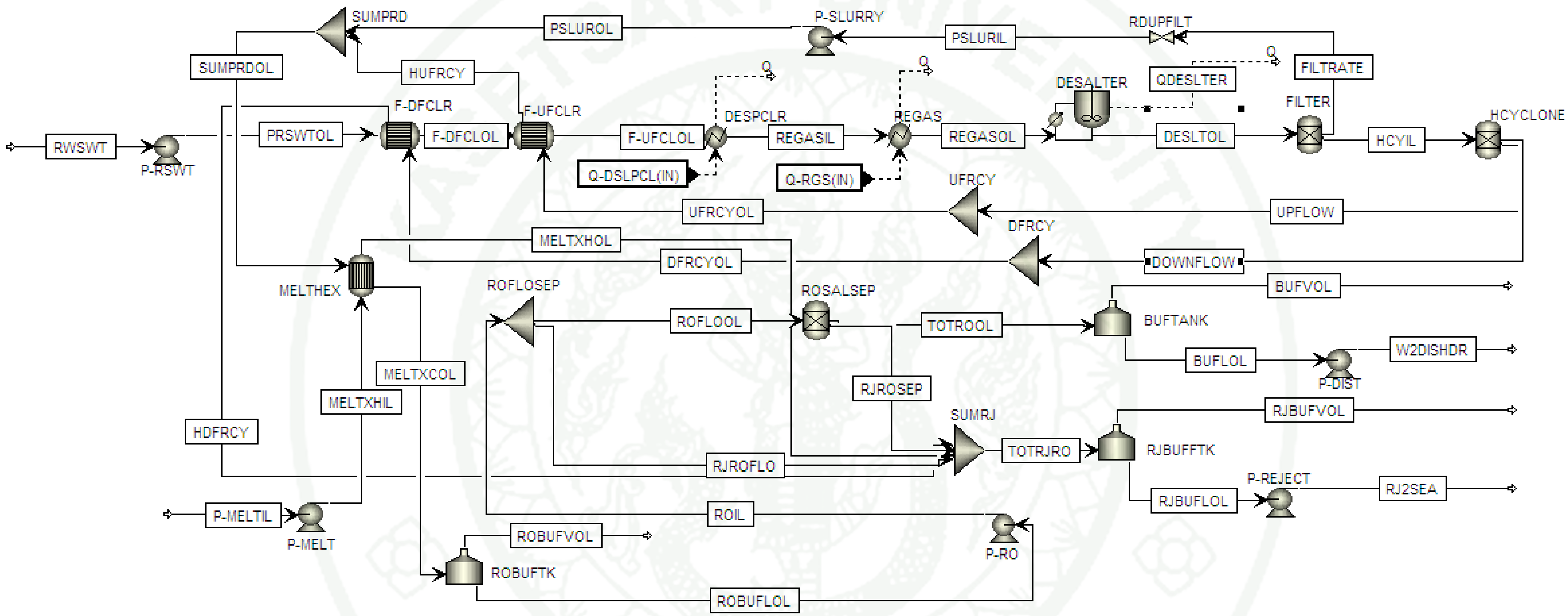


Figure 46 Process simulation model of freeze desalination system with hydrocyclone and RO combination configuration

4.6 Simulation result

4.6.1 Hydrocyclone with electric heating/sunlight radiation configuration

Referring to Table 40, the salinity of water product is high and also higher than drinking water salinity specification. In general, the salinity for drinking water is about 0 - 0.5 ppt and brackish water salinity is about 0.5 – 30 ppt (Wikipedia, n.d.). In this configuration, the salinity of water product is 8.93 and 13.87 ppt at -16 and -30 ^{o}C; respectively. This high salinity in fresh water product is the the result from the hydrocyclone performance. According to CFD simulation, the maximum of separation efficiency is about 80%; therefore, the seawater portion remains in fresh water product and it is also high enough to give the salinity of fresh water to reach 8.93 ppt at -16 ^{o}C. Therefore, the hydrocyclone application in freeze seawater desalination for drinking water purpose is not feasible.

In case that operating temperature = -30 ^{o}C, the salinity of water product is higher than -16 ^{o}C case. This result is consistency to the simulation result of freezing point of feed seawater. The less temperature, the higher salinity the water produces. By this result, it demonstrates that the operating temperature of the freeze desalination unit must be -16 ^{o}C. For the rejected water by-product, the salinity is very high comparing to salinity of feed seawater. The different of salinity is about 140.97 ppt at -16 ^{o}C and 191.92 ppt at -30 ^{o}C. By this result, there is very high risk of salinity changing (immediately) at the disposal area impacts to marine life in such an area. Although the depth of disposal level is located at deep zone or thermocline layer but the salinity gap between surface layer and thermocline layer in GOT is not more than 2 - 3 ppt; therefore, this process may not feasible in aspect of environmental issue.

The energy requirement is about 50-85 MW for melting ice slurry from -16 ^{o}C to 0.1 ^{o}C both electric heating or sunlight radiation. Although treated gas can be used as fuel gas for power generator, the estimated power production at the maximum capacity scenario (Emergency flow) = 43.12 MW. It means that the electricity must be required for drinking water from freeze desalination system and causes the high operation cost. This is the major reason of the unfeasibility of this process configuration.

Table 40 Summary of feed, product, by-product, energy consumption and performance of freeze desalination process with hydrocyclone and electric heating/ sunlight radiation configuration at product temperature = 0.1 ^{o}C

Water source	Parameter	Unit	LNG feed (MTPA)	Crystallization temp = -16 ^{o}C				Crystallization temp = -30 ^{o}C			
				Normal flow		Emergency flow		Normal flow		Emergency flow	
				HiT	LoT	HiT	LoT	HiT	LoT	HiT	LoT
Feed	Raw Seawater	ppt	3-5	35.86	35.86	35.86	35.86	35.86	35.86	35.86	35.86
Fresh water	Salinity	ppt	3	8.93	8.93	8.93	8.93	13.87	13.87	13.87	13.87
			5	8.93	8.93	8.93	8.93	13.87	13.87	13.87	13.87
	Mass flow rate	kg/h	3	543,543	543,543	530,922	530,922	517,986	517,986	505,353	505,353
			5	906,186	906,186	892,724	892,724	864,514	864,514	850,978	850,978
	Volumetric flow rate	m3/h	3	541.52	541.52	528.95	528.95	514.30	514.30	501.76	501.76
			5	902.82	902.82	889.40	889.41	858.36	858.36	844.92	844.92
	Density	kg/m^3	3	1,003.7	1,003.7	1,003.7	1,003.7	1,007.2	1,007.2	1,007.2	1,007.2
			5	1,003.7	1,003.7	1,003.7	1,003.7	1,007.2	1,007.2	1,007.2	1,007.2
Rejected water	Salinity	ppt	3	176.83	176.83	176.83	176.83	227.78	227.78	227.78	227.78
			5	176.83	176.83	176.83	176.83	227.78	227.78	227.78	227.78
	Mass flow rate	kg/h	3	116,017	116,017	113,323	113,323	68,063	68,063	66,403	66,403
			5	193,422	193,422	190,549	190,549	113,596	113,596	111,818	111,818
	Volumetric flow rate	m^3/h	3	103.75	103.75	101.35	101.35	59.17	59.17	57.72	57.72
			5	172.98	172.98	170.41	170.41	98.75	98.75	97.20	97.20
	Density	kg/m^3	3	1,118.2	1,118.2	1,118.2	1,118.2	1,150.4	1,150.4	1,150.4	1,150.4
			5	1,118.2	1,118.2	1,118.2	1,118.2	1,150.4	1,150.4	1,150.4	1,150.4
Energy	Energy consumption	MW	3	51.30	51.30	50.10	50.10	52.71	52.71	51.42	51.42
			5	85.52	85.52	84.25	84.25	87.97	87.97	86.60	86.60
Performance	Desalination efficiency	%	3	75.08	75.08	75.08	75.08	61.31	61.31	61.31	61.31
			5	75.08	75.08	75.08	75.08	61.31	61.31	61.31	61.31

4.6.2 Multi-stage filter configuration

Table 41 shows that the salinity of the fresh water product is 0.153 ppt at -16 ^{o}C and 2.077 ppt at -30 ^{o}C, which follows drinking water specification. The flow rate of product is quite high comparing to product flow rate in hydrocyclone configuration. The reason comes from the configuration is designed to use make-up water to heat up the temperature of the ice slurry product from multi stage filter to reach 0.1 ^{o}C. The real product from ice filter is 416.16 – 710.33 m^3/h at -16 ^{o}C and 402.25 - 688.15 m^3/h at -30 ^{o}C. Table 41 shows the flow rate requirement for melting water. This configuration is feasible for drinking water product for freeze desalination.

Table 41 Flow rate of make-up water and product salinity and disposal water from desalination process with multi-stage filter at product temperature =0.1 ^{o}C

Parameter	Source	LNG feed (MTPA)	Crystallization temp = -16 ^{o}C				Crystallization temp = -30 ^{o}C			
			Normal flow		Emergency flow		Normal flow		Emergency flow	
			HiT	LoT	HiT	LoT	HiT	LoT	HiT	LoT
Volumetric Flow rate (m^3/h)	Melting water	3	1089.8	1089.8	1064.6	1064.5	1115.4	1115.4	1087.9	1088.2
		5	1817.0	1817.0	1789.8	1789.8	1861.8	1864.8	1832.4	1832.5
	Blending water	3	1831.6	1755.2	1777.7	1787.5	1839.4	1837.4	1792.4	1792.4
		5	3055.9	3051.0	3006.9	3006.9	3070.5	3066.6	2948.1	3016.7
Salinity (ppt)	Fresh water	3	0.153	0.153	0.153	0.153	2.077	2.077	2.078	2.077
		5	0.153	0.153	0.153	0.153	2.077	2.074	2.077	2.077
	Reject water	3	43.48	43.78	43.53	43.49	41.95	41.95	41.96	41.96
		5	43.48	43.49	43.49	43.49	41.95	41.95	42.09	41.96

For the rejected water, the concept of stream temperature heat-up identical to water product stream, therefore, seawater blending is required. The total flow rate of reject water is about 1,755.2 - 1,839.6 m^3/h at -16 and -30 ^{o}C but the real rejected water is only 201.47 -338.77 m^3/h at -16 ^{o}C and 144.37- 246.97 m^3/h at -30 ^{o}C. Moreover, the salinity of this configuration is still lower than the rejected stream in hydrocyclone configuration.

Table 42 Summary of process condition of feed seawater, fresh water product and rejected water, energy consumption, performance and fresh water operating cost in freeze desalination process with multi-stage filter configuration

Water source	Parameter	Unit	LNG feed (MTPA)	Crystallization temp = -16 ^{o}C				Crystallization temp = -30 ^{o}C			
				Normal flow		Emergency flow		Normal flow		Emergency flow	
				HiT	LoT	HiT	LoT	HiT	LoT	HiT	LoT
Feed	Raw Seawater	ppt	3-5	35.859	35.859	35.859	35.859	35.859	35.859	35.859	35.859
Fresh water	Salinity	ppt	3	0.153	0.153	0.153	0.153	2.077	2.077	2.078	2.077
			5	0.153	0.153	0.153	0.153	2.077	2.074	2.077	2.077
	Mass flow rate	kg/h	3	1,516,980	1,516,980	1,481,820	1,481,720	1,530,960	1,530,960	1,493,270	1,493,570
			5	2,529,200	2,529,200	2,491,460	2,491,460	2,555,250	2,558,250	2,515,030	2,515,130
	Volumetric flow rate	m^3/h	3	1,515.85	1,515.87	1,480.73	1,480.62	1,527.77	1,527.77	1,490.14	1,490.44
			5	2,527.34	2,527.34	2,489.63	2,489.63	2,549.92	2,552.92	2,509.75	2,509.85
	Density	kg/m^3	3	1,000.74	1,000.73	1,000.74	1,000.75	1,002.09	1,002.09	1,002.10	1,002.10
			5	1,000.74	1,000.74	1,000.74	1,000.74	1,002.09	1,002.09	1,002.10	1,002.10
Rejected water	Salinity	ppt	3	43.48	43.78	43.53	43.49	41.95	41.95	41.96	41.96
			5	43.48	43.49	43.49	43.49	41.95	41.95	42.09	41.96
	Mass flow rate	kg/h	3	2,096,080	2,018,080	2,035,830	2,045,830	2,042,090	2,040,090	1,990,080	1,990,080
			5	3,496,910	3,491,910	3,441,310	3,441,310	3,408,860	3,404,860	3,279,570	3,349,570
	Volumetric flow rate	m^3/h	3	2,037.81	1,961.59	1,979.18	1,988.95	1,987.36	1,985.41	1,936.74	1,936.74
			5	3,399.73	3,394.83	3,345.65	3,345.65	3,317.51	3,313.60	3,191.35	3,259.78
	Density	kg/m^3	3	1,028.59	1,028.80	1,028.62	1,028.60	1,027.54	1,027.54	1,027.54	1,027.54
			5	1,028.59	1,028.60	1,028.59	1,028.59	1,027.53	1,027.54	1,027.64	1,027.54
Energy	Energy consumption	MW	3	1.80	1.76	1.76	1.76	1.78	1.78	1.73	1.73
			5	3.01	3.01	2.96	2.96	2.97	2.82	3.17	2.92
Performance	Desalination efficiency	%	3	99.57	99.57	99.57	99.57	94.21	94.21	94.21	94.21
			5	99.57	99.57	99.57	99.57	94.21	94.21	94.21	94.21
Benefit	Desalination cost	THB/L	3	0.003569	0.003487	0.003556	0.003567	0.003494	0.003492	0.003492	0.003492
			5	0.003570	0.003567	0.003568	0.003568	0.003494	0.003319	0.003788	0.003491

Table 42, the salinity of the rejected water is about 41.96 – 43.78 ppt for -16 and -30 °C which are lower than hydrocyclone configuration about 133.3 – 185.83 ppt or about 4.1 – 5.4 times. The different of salinity between the rejected water and disposal point is still about 5 - 7 ppt. It is much better than the salinity from hydrocyclone configuration. In this case, the effect of salinity different may occur but it can be reduced the risk of salinity different by deeper the point of rejected water disposal. However, this value is very low comparing to current project in many other countries.

Table 43 Breakdown of energy consumption from each source in freeze desalination process with multi stage filter configuration

Power load user	**Relate source media**	**LNG feed (MTPA)**	**Energy consumption (MW)**							
			Crystallization temperature (°C)							
			-16				**-30**			
			Normal flow		**Emergency flow**		**Normal flow**		**Emergency flow**	
			HiT	**LoT**	**HiT**	**LoT**	**HiT**	**LoT**	**HiT**	**LoT**
P-RSWT	Feed seawater	3	0.090	0.090	0.088	0.088	0.080	0.080	0.078	0.078
		5	0.150	0.150	0.147	0.147	0.133	0.133	0.131	0.131
P-SLURRY	Melting water	3	0.215	0.215	0.210	0.210	0.217	0.217	0.211	0.211
		5	0.358	0.358	0.353	0.353	0.362	0.217	0.356	0.356
P-RJWSWT	Blending water	3	0.260	0.249	0.252	0.253	0.261	0.260	0.254	0.254
		5	0.433	0.432	0.426	0.426	0.435	0.435	0.418	0.428
P-REJECT	Rejected water	3	0.809	0.779	0.786	0.790	0.789	0.788	0.769	0.769
		5	1.350	1.348	1.328	1.328	1.317	1.316	1.552	1.294
P-DIST	Fresh water	3	0.430	0.430	0.420	0.420	0.433	0.433	0.423	0.423
		5	0.717	0.717	0.706	0.706	0.723	0.724	0.712	0.712
Total		3	1.803	1.762	1.755	1.761	1.779	1.778	1.735	1.735
		5	3.008	3.005	2.961	2.961	2.970	2.825	3.169	2.921

Note: P-RSWT means seawater supply pump.

P-SLURRY means mixed product water transfer pump to product distribution pump buffer tank.

P-RJWSWT means blending seawater supply pump.

P-REJECT means rejected water disposal pump.

P-DIST means water product distribution pump.

According to Table 43, the total power consumption is less than hydrocyclone configuration because the natural energy from mixing water and blending water are used. With this consumption, the utilization of internal generated power is adequate and has the remaining power for sale to trade off to water make-up cost.

(a) Effect of blending water on product and disposal treatment and its cost

Table 44 shows that the flow rate requirement for temperature treatment is 3,050.96 m^3/h to reach 30 ^oC. The total power consumption is 1.78 MW which is mainly for the blending water supply pump and the disposal pump. However, the maximum internal generated power is about 43.12 MW (Emergency flow scenario). There is also margin to utilize them for reducing the salinity in disposal water. Referring to Table 45, sensitivity analysis of blending water rate to salinity has been performed. The result shows that although the flow rate can be increased extremely to 96,587.46 m3/h or 3,166.8 times the salinity can be reduced to 36.12 ppt only (0.6 ppt differs from blending water salinity). Moreover, the total power requirement is 52.16 MW, which is over the capacity of internal power generation. The reason is the concentration of disposal and blending water is close together, therefore, the blending capacity requirement will be much higher than the capacity at high different salinity. Although the cost of treatment is not too expensive (0.082547 THB/Liter) as shown in Table 46, the relative CAPEX and OPEX e.g. supply blending pipeline, piping system, pump size, the number of pump due to sparing pump, large size of switch gear room, maintenance cost is very expensive and not economic for the implementation. Therefore, the utilization of internal power production to reduce the salinity is not suitable for selection.

Table 44 Blending water requirement for reject water temperature treatment and its energy consumption in freeze desalination process with multi-stage filter based on 5 MTPA LNG at -16 ^{o}C and flow rate = 2,527.34 m^3/h at 0.1 ^{o}C and salinity = 0.15 ppt.

Step size	Blend Water (TAG ID: BLDRWSWT)					Reject Water (TAG ID: RJBUFOL)						Energy consumption (MW)		
	Mass flow	Vol. flow	Density	Temp.	Salinity	Mass flow	Vol. flow	Density	Temp.	Salinity		P-RJWSWT	P-REJECT	Total
kg/h	kg/h	m^3/h	g/ml	^{o}C	ppt	kg/h	m^3/h	g/ml	^{o}C	g/L	ppt			
-	-	-	1.02	35.00	35.86	376,911	347.22	1.09	(6.90)	67.73	122.36	-	0.14	0.14
100,000	100,000	97.94	1.02	35.00	35.86	476,911	447.36	1.07	(4.34)	51.65	93.32	0.01	0.18	0.19
100,000	200,000	195.89	1.02	35.00	35.86	576,911	544.43	1.06	2.81	46.01	83.13	0.03	0.22	0.24
200,000	400,000	391.78	1.02	35.00	35.86	776,911	739.34	1.05	11.41	39.14	70.71	0.06	0.29	0.35
134,000	534,000	523.02	1.02	35.00	35.86	910,911	870.22	1.05	15.00	36.25	65.49	0.07	0.35	0.42
66,000	600,000	587.66	1.02	35.00	35.86	976,911	934.72	1.05	16.40	35.12	63.45	0.08	0.37	0.45
200,000	800,000	783.55	1.02	35.00	35.86	1,176,911	1,130.30	1.04	19.66	32.48	58.68	0.11	0.45	0.56
200,000	1,000,000	979.44	1.02	35.00	35.86	1,376,911	1,325.98	1.04	21.94	30.62	55.32	0.14	0.53	0.66
406,000	1,406,000	1,377.09	1.02	35.00	35.86	1,782,911	1,723.38	1.03	25.00	28.14	50.84	0.20	0.68	0.88
94,000	1,500,000	1,469.16	1.02	35.00	35.86	1,876,911	1,815.40	1.03	25.51	27.72	50.08	0.21	0.72	0.93
500,000	2,000,000	1,958.88	1.02	35.00	35.86	2,376,911	2,304.97	1.03	27.56	26.05	47.06	0.28	0.91	1.19
500,000	2,500,000	2,448.60	1.02	35.00	35.86	2,876,911	2,794.59	1.03	28.89	24.96	45.10	0.35	1.11	1.46
615,000	3,115,000	3,050.96	1.02	35.00	35.86	3,491,911	3,396.88	1.03	30.00	24.06	43.46	0.43	1.35	1.78

Table 45 Blending water requirement for reject water salinity treatment and its energy consumption for the treatment in the process with multi-stage filter based on 5 MTPA LNG at -16 ^{o}C and flow rate = 2,527.34 m^{3}/h at 0.1 ^{o}C and salinity = 0.15 ppt.

Step size	Blend Water (TAG ID: BLDRWSWT)					Reject Water (TAG ID: RJBUFOL)						Energy consumption (MW)		
	Mass flow	Vol. flow	Density	Temp.	Salinity	Mass flow	Vol. flow	Density	Temp.	Salinity		P-RJWSWT	P-REJECT	Total
kg/h	kg/h	m^{3}/h	g/ml	^{o}C	ppt	kg/h	m^{3}/h	g/ml	^{o}C	g/L	ppt			
-	3,115,000	3,050.96	1.02	35.00	35.86	3,491,911	3,396.88	1.03	30.00	24.06	43.46	0.43	1.35	1.78
100,000	3,215,000	3,148.90	1.02	35.00	35.86	3,591,911	3,494.81	1.03	30.14	23.94	43.25	0.45	1.39	1.83
100,000	3,315,000	3,246.84	1.02	35.00	35.86	3,691,911	3,592.75	1.03	30.28	23.83	43.05	0.46	1.43	1.89
200,000	3,515,000	3,442.73	1.02	35.00	35.86	3,891,911	3,788.63	1.03	30.53	23.62	42.68	0.49	1.50	1.99
200,000	3,715,000	3,638.62	1.02	35.00	35.86	4,091,911	3,984.50	1.03	30.75	23.44	42.34	0.52	1.58	2.10
500,000	4,215,000	4,128.34	1.02	35.00	35.86	4,591,911	4,474.20	1.03	31.23	23.04	41.63	0.59	1.78	2.36
500,000	4,715,000	4,618.06	1.02	35.00	35.86	5,091,911	4,963.91	1.03	31.62	22.73	41.06	0.65	1.97	2.62
500,000	5,215,000	5,107.78	1.02	35.00	35.86	5,591,911	5,453.63	1.03	31.93	22.47	40.60	0.72	2.16	2.89
800,000	6,015,000	5,891.33	1.02	35.00	35.86	6,391,911	6,237.18	1.02	32.33	22.14	40.00	0.83	2.48	3.31
800,000	6,815,000	6,674.88	1.02	35.00	35.86	7,191,911	7,020.74	1.02	32.65	21.89	39.54	0.95	2.79	3.73
800,000	7,615,000	7,458.43	1.02	35.00	35.86	7,991,911	7,804.31	1.02	32.89	21.68	39.17	1.06	3.10	4.15
1,000,000	8,615,000	8,437.87	1.02	35.00	35.86	8,991,911	8,783.77	1.02	33.14	21.48	38.80	1.20	3.49	4.68
2,000,000	10,615,000	10,396.75	1.02	35.00	35.86	10,991,911	10,742.70	1.02	33.50	21.18	38.26	1.47	4.26	5.74
5,000,000	15,615,000	15,293.95	1.02	35.00	35.86	15,991,911	15,640.08	1.02	34.01	20.76	37.51	2.17	6.21	8.37
8,000,000	23,615,000	23,129.47	1.02	35.00	35.86	23,991,911	23,475.91	1.02	34.38	20.46	36.96	3.28	9.32	12.59
10,000,000	33,615,000	32,923.87	1.02	35.00	35.86	33,991,911	33,270.72	1.02	34.60	20.28	36.63	4.67	13.20	17.87
15,000,000	48,615,000	47,615.47	1.02	35.00	35.86	48,991,911	47,962.94	1.02	34.76	20.15	36.40	6.75	19.03	25.78
20,000,000	68,615,000	67,204.26	1.02	35.00	35.86	68,991,911	67,552.59	1.02	34.87	20.06	36.24	9.52	26.81	36.33
30,000,000	98,615,000	96,587.46	1.02	35.00	35.86	98,991,911	96,937.06	1.02	34.95	20.00	36.12	13.69	38.47	52.16

Table 46 Cost of reject water salinity treatment in the freeze desalination process with multi-stage filter based on 5 MTPA LNG at -16 ^{o}C and flow rate = 2,527.34 m^3/h at 0.1 ^{o}C and salinity = 0.15 ppt.

Blend Water (TAG ID: BLDRWSWT)	Reject Water (TAG ID: RJBUFOL)				Treatment cost (Electricity cost = 4 THB/kWh)	
Volumetric flow	Volumetric flow	Density	Temperature	Salinity		
m^3/h	m^3/h	g/ml	^{o}C	ppt	MMTHB/yr	THB/L
3,051	3,397	1.03	30.00	43.46	58.97	0.002818
3,149	3,495	1.03	30.14	43.25	60.71	0.002901
3,247	3,593	1.03	30.28	43.05	62.46	0.002985
3,443	3,789	1.03	30.53	42.68	65.95	0.003152
3,639	3,985	1.03	30.75	42.34	69.45	0.003319
4,128	4,474	1.03	31.23	41.63	78.18	0.003736
4,618	4,964	1.03	31.62	41.06	86.92	0.004153
5,108	5,454	1.03	31.93	40.60	95.65	0.004571
5,891	6,237	1.02	32.33	40.00	109.63	0.005239
6,675	7,021	1.02	32.65	39.54	123.60	0.005907
7,458	7,804	1.02	32.89	39.17	137.58	0.006574
8,438	8,784	1.02	33.14	38.80	155.05	0.007409
10,397	10,743	1.02	33.50	38.26	189.99	0.009079
15,294	15,640	1.02	34.01	37.51	277.34	0.013253
23,129	23,476	1.02	34.38	36.96	417.11	0.019932
32,924	33,271	1.02	34.60	36.63	591.81	0.028281
47,615	47,963	1.02	34.76	36.40	853.87	0.040804
67,204	67,553	1.02	34.87	36.24	1,203.28	0.057501
96,587	96,937	1.02	34.95	36.12	1,727.40	0.082547

4.6.3 Hydrocyclone with reverse osmosis combination configuration

Table 47 shows that the salinity of the product water is about 0.09 ppt at operating temperature = -16 ^{o}C and 0.14 ppt at operating temperature = -30 ^{o}C. This specification can be applied for the drinking water. The water product flow rate is less than the multi-stage filter configuration about 111.06 m^3/h at 3 MTPA of LNG and 185.16 m^3/h at 5 MTPA. The rejected water flow rate is very high as shown in Table 48 because the main portion of the rejected water is from the seawater from water for melting the slurry ice. It's approximate 4.7- 4.8 times for all scenarios. This configuration use heat exchanger for increasing the temperature of the product, therefore, the amount of water requirement is too high if comparing to direct blending water. The energy consumption of this configuration is better than the multi-stage configuration because there are 2 sets of feed-product heat exchanger reduced the energy consumption. Therefore, the energy consumption different between these configurations should be in the reasonable range. However, the major power consumption is the high-pressure pump in RO unit as shown in Table 49. The pressure for this unit is required at 80 barg, therefore, the power consumption for increasing the pressure from atmospheric (Tank pressure) to 80 barg is very high and the power consumption is close to the power consumption by melting water pump.

While the comparison of energy consumption between the multi-stage filter and this configuration, it shows that the energy consumption of multi-stage filter is higher than this configuration about 4.8 – 4.9 times. Moreover, the salinity of the rejected water is less than the multi-stage filter configuration also. The salinity is about 37.5 ppt. It is different to the feed water about 2 ppt. The reason is the melting water is very high and when it is mixed to all rejected water from all parts in the process such as downflow product from hydrocyclone and rejected stream from RO, it help to reduce the salinity of these streams. The advantage point of this salinity is the impaction to the environment at the disposal point.

Table 47 Summary of process condition of feed seawater, fresh water product and rejected water, energy consumption, performance and fresh water operating cost in freeze desalination process with hydrocyclone-RO combination configuration

Water source	Parameter	Unit	LNG feed (MTPA)	Operating temperature (^{o}C)							
				-16				-30			
				Normal flow		Emergency flow		Normal flow		Emergency flow	
				LoT	HiT	LoT	LoT	HiT	HiT	LoT	HiT
Feed	Raw Seawater	ppt	3-5	35.86	35.86	35.86	35.86	35.86	35.86	35.86	35.86
Fresh water	Salinity	ppt	3	0.09	0.09	0.09	0.09	0.14	0.14	0.14	0.14
			5	0.09	0.09	0.09	0.09	0.14	0.14	0.14	0.14
	Mass flow rate	kg/h	3	430,698	430,698	420,697	420,697	408,857	408,857	398,885	398,885
			5	718,052	718,052	707,384	707,384	682,378	682,378	671,694	671,694
	Volumetric flow rate	m3/h	3	430	430	420	420	409	409	399	399
			5	718	718	707	707	682	682	671	671
	Density	kg/m3	3	1,001	1,001	1,001	1,001	1,001	1,001	1,001	1,001
			5	1,001	1,001	1,001	1,001	1,001	1,001	1,001	1,001
Rejected water	Salinity	ppt	3	37.54	37.54	37.51	37.51	37.47	37.47	37.43	37.43
			5	37.52	37.52	37.50	37.28	37.45	37.45	37.42	37.42
	Mass flow rate	kg/h	3	9,877,240	9,877,240	9,871,920	9,871,920	9,825,560	9,825,560	9,821,240	9,821,240
			5	16,717,400	16,717,400	16,711,800	19,264,200	16,631,600	16,631,600	16,627,000	16,627,000
	Volumetric flow rate	m3/h	3	9,642	9,642	9,637	9,637	9,591	9,591	9,588	9,588
			5	16,319	16,319	16,314	18,813	16,236	16,235	16,232	16,232
	Density	kg/m3	3	1,024	1,024	1,024	1,024	1,024	1,024	1,024	1,024
			5	1,024	1,024	1,024	1,024	1,024	1,024	1,024	1,024
Energy	Energy consumption	MW	3	8.67	8.67	8.63	8.63	8.56	8.56	8.52	8.52
			5	14.64	14.64	14.60	16.44	14.45	12.15	14.41	14.41
Performance	Desalination efficiency	%	3	99.75	99.75	99.75	99.75	99.61	99.61	99.61	99.61
			5	99.75	99.75	99.75	99.75	99.61	99.61	99.61	99.61
Benefit	Desalination cost	THB/L	3	0.0604	0.0604	0.0616	0.0616	0.0628	0.0628	0.0641	0.0641
			5	0.0612	0.0612	0.0619	0.0698	0.0636	0.0534	0.0644	0.0644

Table 48 Flow rate (m^3/h) melting water and rejected water from desalination process with hydrocyclone-RO combination at product temperature = 0.1 ^oC

Source	LNG feed (MTPA)	Operating temperature (^oC)							
		-16				-30			
		Normal flow		Emergency flow		Normal flow		Emergency flow	
		HiT	LoT	HiT	LoT	HiT	LoT	HiT	LoT
Melting water	3	9,450	9,450	9,450	9,450	9,450	9,450	9,450	9,450
	5	16,000	16,000	16,000	18,500	16,000	16,000	16,000	16,000
Rejected water	3	9,642	9,642	9,637	9,637	9,591	9,591	9,588	9,588
	5	16,319	16,319	16,314	18,813	16,236	16,235	16,232	16,232

Table 49 Breakdown of energy consumption from each source in freeze desalination process with hydrocyclone-RO combination configuration

Source	LNG feed (MTPA)	Operating temperature (^oC)							
		-16				-30			
		Normal flow		Emergency flow		Normal flow		Emergency flow	
		HiT	LoT	HiT	LoT	HiT	LoT	HiT	LoT
P-SWSUBM	3	2.003	2.003	2.000	2.000	1.989	1.989	1.986	1.986
	5	3.388	3.388	3.385	3.881	3.365	3.365	3.362	3.362
P-DIST	3	0.122	0.122	0.119	0.119	0.116	0.116	0.113	0.113
	5	0.204	0.204	0.200	0.200	0.193	0.193	0.190	0.190
P-MELT	3	1.339	1.339	1.339	1.339	1.339	1.339	1.339	1.339
	5	2.268	2.268	2.268	2.622	2.268	2.268	2.268	2.268
P-REJECT	3	3.828	3.828	3.826	3.826	3.808	3.808	3.807	3.807
	5	6.480	6.480	6.478	7.470	6.447	4.143	6.445	6.445
P-RO	3	1.223	1.223	1.195	1.195	1.162	1.162	1.134	1.134
	5	2.040	2.040	2.009	2.009	1.939	1.939	1.909	1.909
P-RSWT	3	0.090	0.090	0.088	0.088	0.080	0.080	0.078	0.078
	5	0.150	0.150	0.147	0.147	0.133	0.133	0.131	0.131
P-SLURRY	3	0.067	0.067	0.065	0.065	0.065	0.065	0.063	0.063
	5	0.112	0.112	0.110	0.110	0.108	0.108	0.106	0.106
Total Energy	3	8.673	8.673	8.633	8.633	8.559	8.559	8.520	8.520
	5	14.640	14.640	14.598	16.440	14.453	12.149	14.411	14.411

Note: P-RO means high-pressure pump in RO unit

The result of all configuration shows that the multi-stage filter configuration gives the best in power consumption and the amount of water handling. The hydrocyclone with RO combination gives the best performance in the result of product and rejected water salinity. However, the selected configuration cannot be decided at this stage because it depends on the investment cost and economic evaluation result in each configuration, which is shown in the following sections.

5. CFD modeling and analysis of hydrocyclone

5.1 Effect of inlet velocity

Theoretically, the higher velocity increases the centrifugal force; therefore, higher inlet velocity can increase the separation efficiency as illustrated in Figure 47.

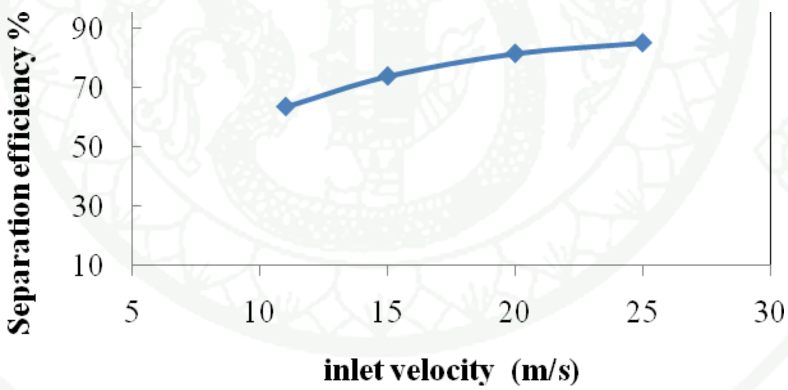


Figure 47 Effect of inlet velocities on the separation efficiency Condition: ice fraction = 0.2, temperature = -16°C and pressure = 5 barg

Figure 48 shows total pressure distribution along the radius distance inside hydrocyclone. The total pressure gradually decreases when the radius decreases inwards to the center of hydrocyclone. The maximum pressure occurs near hydrocyclone wall due to tangential entry of hydrocyclone feed.

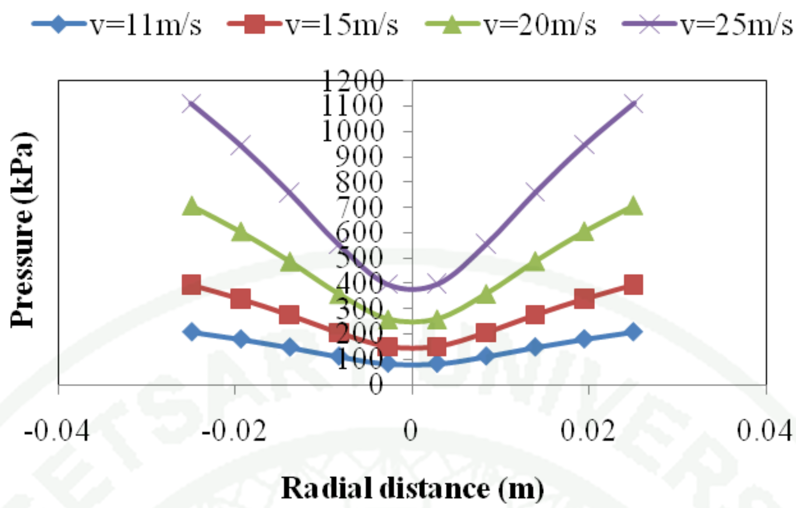
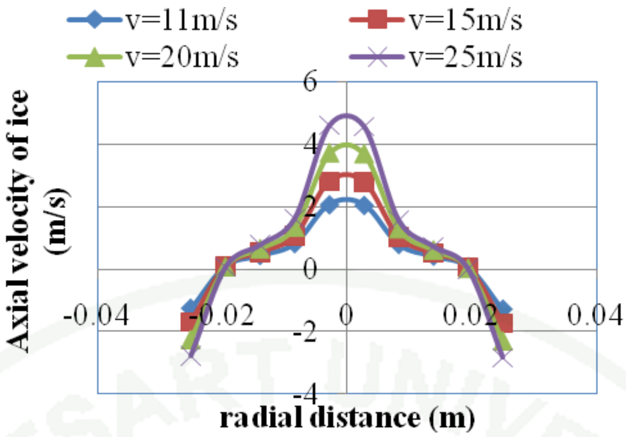
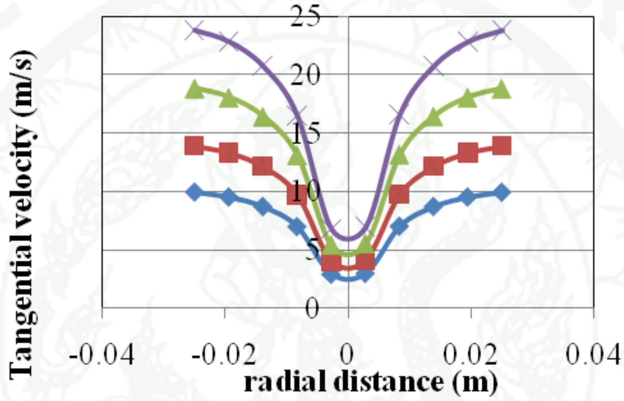


Figure 48 Effect of inlet velocity on pressure profile below 4.5 cm of vortex finder

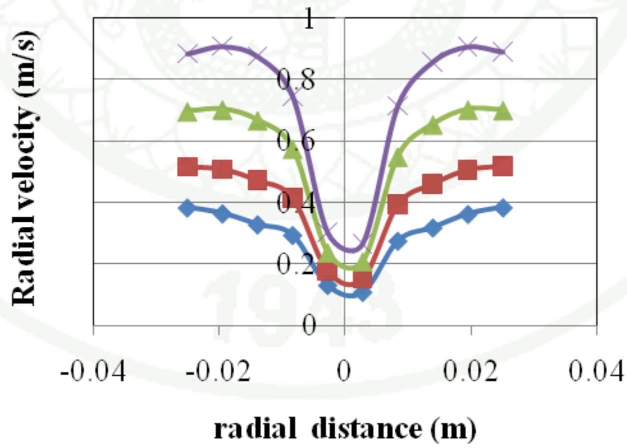
The flow characteristics of ice-sea water are represented by the radial velocity, axial velocity, and the tangential velocity profile at level of 4.5 cm below the vortex finder. Figure 49 (a) shows both positive axial velocity and the negative axial velocity. The positive axial velocity means the material goes to the overflow and the negative value means that the material goes to the underflow side. The negative axial velocity flow removes the particle to the underflow. The tangential velocity magnitude increases with the radius. The higher tangential velocity represents higher centrifugal force. The radial velocity profile is similar to the tangential velocity profile. The magnitudes of radial velocity profile are much smaller than the axial velocity and the tangential velocity.



(a)



(b)



(c)

Figure 49 Effect of inlet velocity profile of ice (a) axial velocity (b) tangential velocity (c) radial velocity

5.2 Effect of solid concentration

Higher solid concentration in hydrocyclone feed limits the efficiency of separation, because the increment solid volume fraction increases the resistance to flow. Increasing solid concentration also raises the bulk viscosity of mixture slurry. Figure 50 shows the relationship between the separation efficiency and the inlet ice fraction.

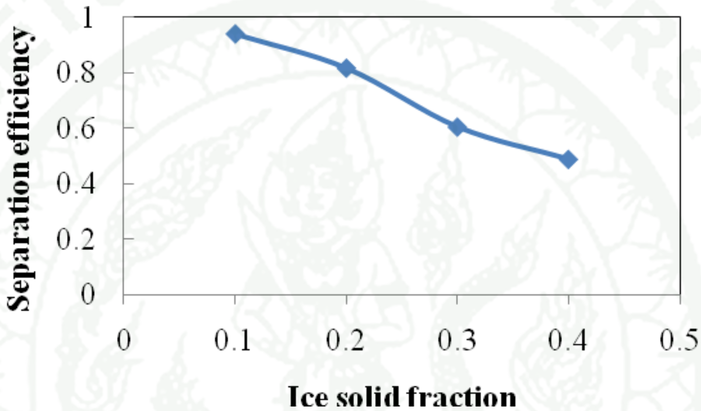


Figure 50 Effect of ice solid fraction on the separation efficiency

5.3 Effect of overflow diameter and underflow diameter

The overflow diameter and the underflow diameter are varied to determine their effects on the flow pattern of ice, the split ratio, and purity of ice. The influence of overflow diameter adjustment on the separation is similar to that of underflow diameter adjustment as shown in Figures 51, 52, 53 and 54. The increase of D_o/D_u results in the higher split fraction of slurry to overflow side, but the lower ice purity is found in the overflow stream. The reason is that seawater recovery in the overflow stream continuously rises at higher value of D_o/D_u, while the ice recovery gradually increases.

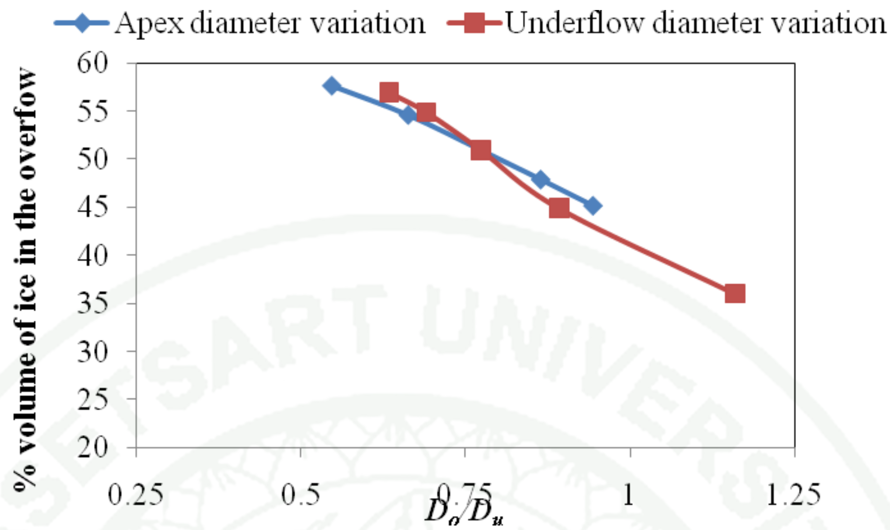


Figure 51 Purity of ice at the overflow opening operating at the inlet velocity at 20 m/s and ice concentration at 20 % v/v

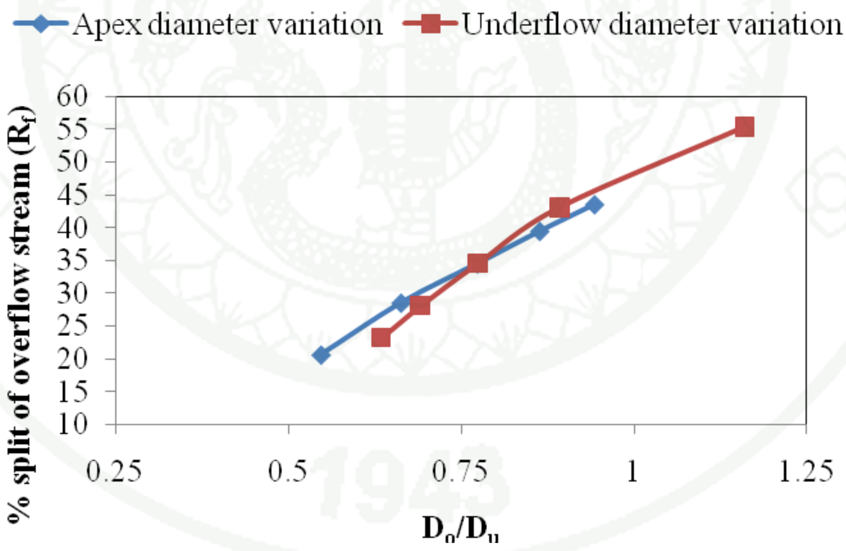


Figure 52 Percent split fraction of the overflow stream operating at the inlet velocity at 20 m/s and ice concentration at 20 % v/v

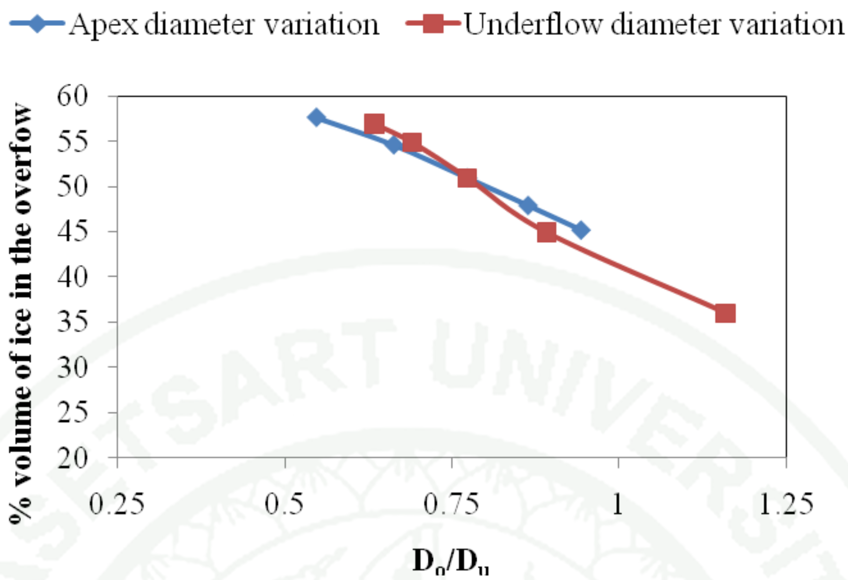


Figure 53 Ice recovery in the overflow stream at the overflow opening operating at the inlet velocity at 20 m/s and ice concentration at 20 % v/v

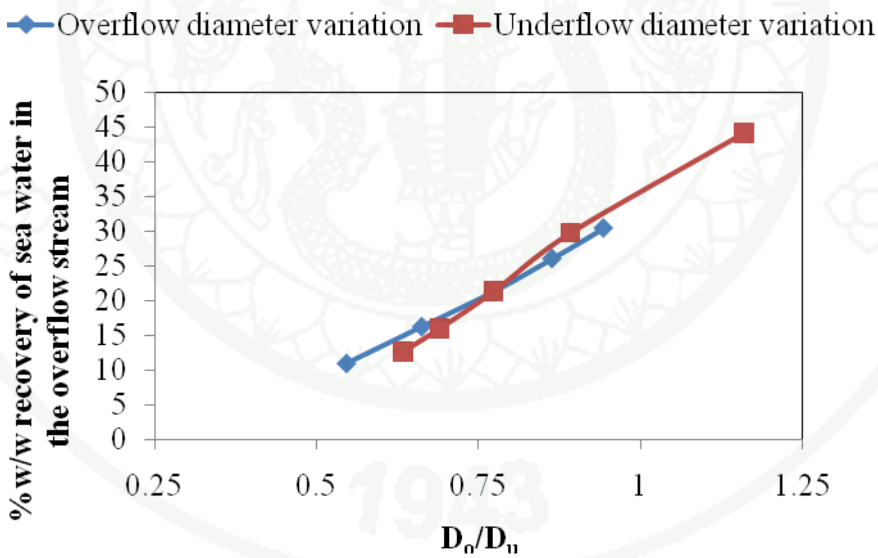


Figure 54 Seawater recovery in the overflow stream at the overflow opening operating at the inlet velocity at 20 m/s and ice concentration at 20 % v/v

5.4 Transient behavior

For transient study of separation in hydrocyclone, transient simulation of hydrocyclone at 20 % v/v of ice and inlet velocity at 20 m/s were performed. The operating condition reaches steady state condition in time less than 0.2 s. The mass flow rate at overflow and underflow are observed. Figure 55 shows the mass flow rate of overflow and underflow at steady state when reach 0.23 s. The contour of ice volume fraction is shown in Figure 56.

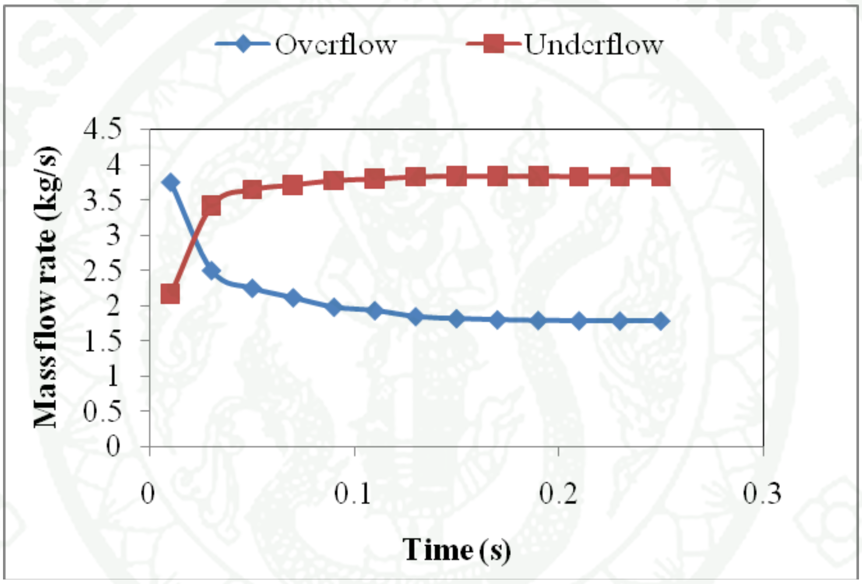
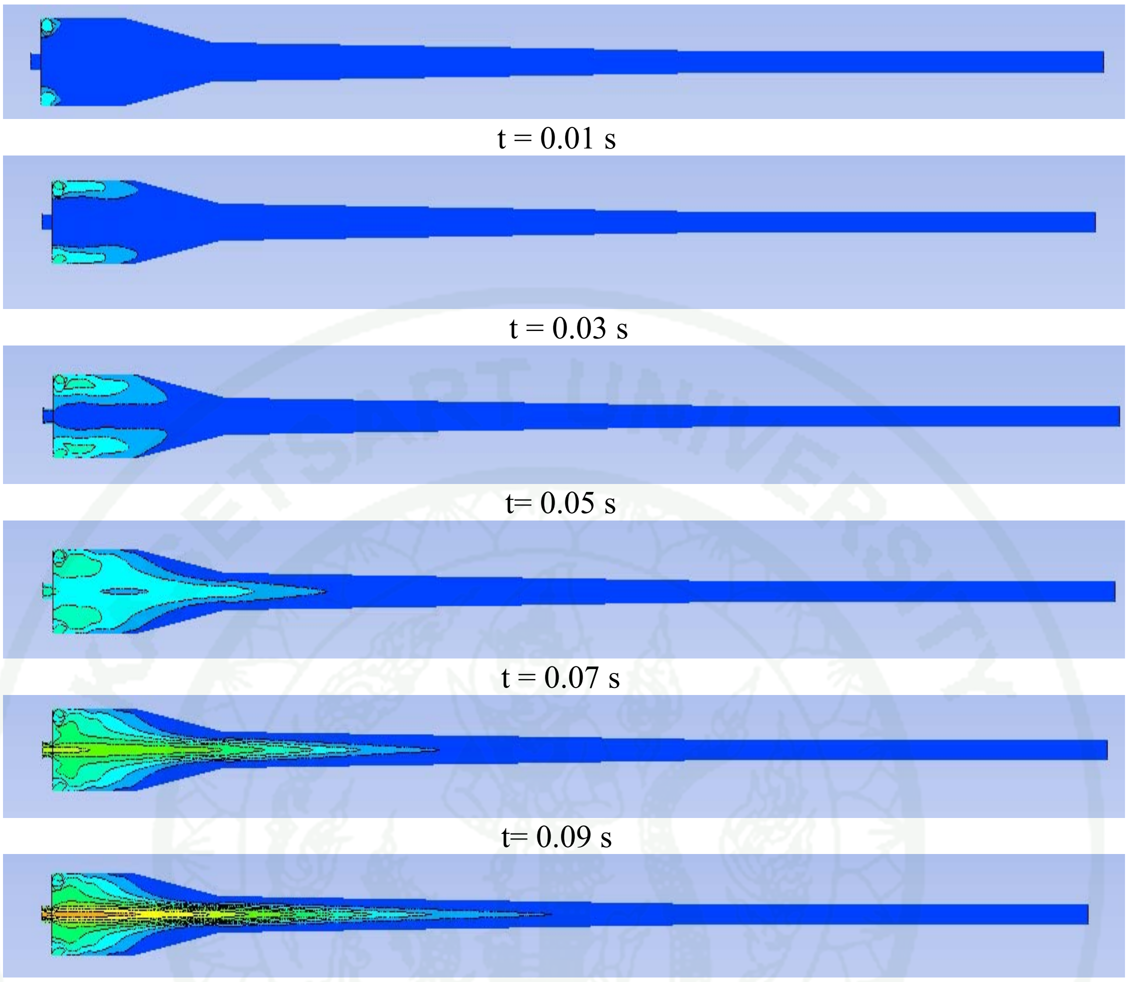


Figure 55 Transient profile of mass flow rate at overflow and underflow



t = 0.01 s

t = 0.03 s

t= 0.05 s

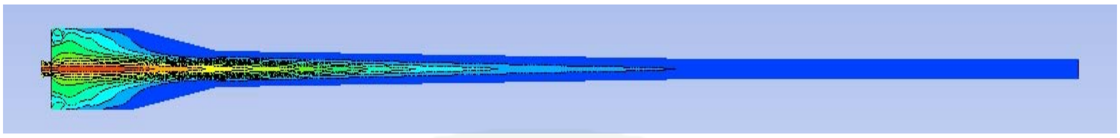
t = 0.07 s

t= 0.09 s

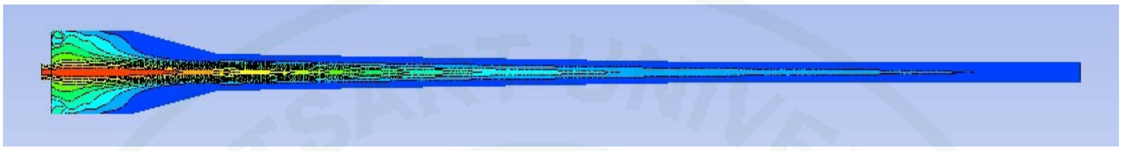
t = 0.11 s

Figure 56 Transient behavior of ice-sea water slurry inside hydrocyclone

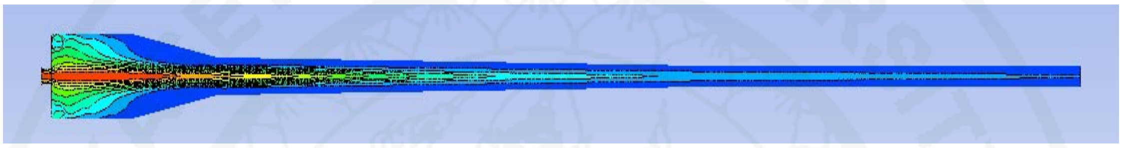
t = 0.13 s



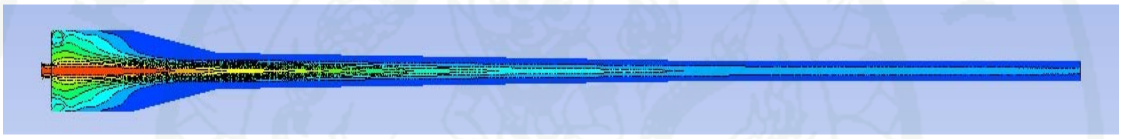
t = 0.15 s



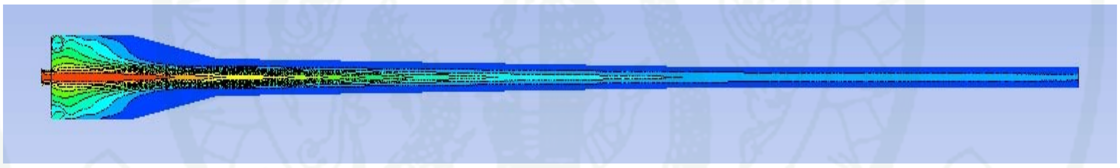
t = 0.17 s



t = 0.19 s



t = 0.21 s



t = 0.23 s

Figure 56 (Continued)

6. Cost Estimation

According to the process simulation results, there are 2 configurations under consideration, multi-stage filter and hydrocyclone-RO combination. Therefore, this part will break down the estimate cost of each configuration. The cost estimation was performed in 6 parts:

- Main equipment cost
- Construction cost
- Operating cost (included maintenance cost)
- Product sale income
- Raw material cost

Main equipment cost included cost of pipeline, power generator, instrument package and electrical package attached with main equipments (Tables 50 and 51). The cost of instrumentation and electrical is included in the margin. Construction cost includes land cost, soil and concrete work and rental cost of heavy machine. The total of main equipment cost and construction cost are included in CAPEX (Capital Expenditure).

The result also shows that the investment cost of hydrocyclone-RO combination is more expensive than the multi-stage filter configuration amount 2.28 times. The major reason is the size of rejected water pipeline and the number of rejected water pump and melting water pump. The concern of the rejected water salinity leads to the investment cost is too high, however, with 2 ppt salinity deviation between product and disposal seawater is not a concern, the investment cost will be reduced of 600 MMTHB. Another major cost is tank. Although the amount of water requirement is increased due to the melting purpose, but the material leads the price different. Since the product from RO combination configuration is designed as drinking water, therefore, carbon steel is not applied for this purpose due to the contaminate of iron and its oxide during long term operation, therefore, Fiberglass Reinforced Plastic (FRP) is applied on this purpose instead. The different cost of tank is about 178 MMTHB and it leads the total different cost between these configurations is 590 MMTHB, approximately.

Table 50 Summary of main equipment cost of multi-stage filter and hydrocyclone and RO combination configuration

Item No.	Equipment description	Estimated cost (THB)	
		Hydrocyclone + RO	Multi-stage filter
1	Heat Exchanger	47,008,406	26,557,406
2	2 phase separator	9,387,710	9,387,710
3	Dehydrator	2,957,873	2,957,873
4	Pump	59,844,150	43,350,510
5	Compressor	64,596,000	64,596,000
6	Pipe	288,871,250	288,871,250
7	Pipeline	607,500,000	303,750,000
8	Filter	48,381,000	96,762,000
9	Hydrocyclone	60,000,000	0
10	RO	5,941,801	0
11	Tank	386,373,000	207,906,000
12	Gas turbine Generator	855,000,000	855,000,000
	Total	2,435,861,190	1,899,138,749
	Margin (10%)	243,586,119	189,913,875
	Grand Total	2,679,447,309	2,089,052,624

Table 51 Detail of main equipment cost of multi-stage filter configuration and hydrocyclone and RO combination configuration

Configuration	Hydrocyclone + RO combination			Multi-stage Filter		
	Heat Exchanger					
Name	Estimated unit cost (THB)	Qty (unit)	Estimated cost (THB)	Estimated unit cost (THB)	Qty (unit)	Estimated cost (THB)
DESALINATION PRE-COOLER	9,801,703	1	9,801,703	9,801,703	1	9,801,703
LNG VAPORIZER	9,801,703	1	9,801,703	9,801,703	1	9,801,703
GNTWTRT.1SCLR	336,000	1	336,000	336,000	1	336,000
GNTWTRT.2NCLR	123,000	1	123,000	123,000	1	123,000
GNTWTRT.3DCLR	147,000	1	147,000	147,000	1	147,000
GNTWTRT.BLWCLR	1,803,000	1	1,803,000	1,803,000	1	1,803,000

Table 51 (Continued)

Configuration	Hydrocyclone + RO combination			Multi-stage Filter		
Heat Exchanger						
Name	**Estimated unit cost (THB)**	**Qty (unit)**	**Estimated cost (THB)**	**Estimated unit cost (THB)**	**Qty (unit)**	**Estimated cost (THB)**
GNTWTRT.BSCCLR	2,310,000	1	2,310,000	2,310,000	1	2,310,000
GNTWTRT.F-PRDHEX	132,000	1	132,000	132,000	1	132,000
SWTTRT.F-DFCLR	22,554,000	1	22,554,000	2,103,000	1	2,103,000
SWTTRT.F-UFCLR	4,968,000	1	4,968,000	0	0	0
SWTTRT.MELTHEX	26,112,000	3	78,336,000	0	0	0
Total			47,008,406			26,557,406
2 phase separator						
Name	**Estimated unit cost (THB)**	**Qty (unit)**	**Estimated cost (THB)**	**Estimated unit cost (THB)**	**Qty (unit)**	**Estimated cost (THB)**
GNTWTRT.1SLTS	1,919,077	1	1,919,077	1,919,077	1	1,919,077
GNTWTRT.2NLTS	1,436,028	1	1,436,028	1,436,028	1	1,436,028
GNTWTRT.3DLTS	1,436,028	1	1,436,028	1,436,028	1	1,436,028
GNTWTRT.BLWKO1	1,316,800	1	1,316,800	1,316,800	1	1,316,800
GNTWTRT.BLWKO2	974,383	1	974,383	974,383	1	974,383
GNTWTRT.BSCKO1	1,316,800	1	1,316,800	1,316,800	1	1,316,800
GNTWTRT.BSCKO2	448,727	1	448,727	448,727	1	448,727
GNTWTRT.F-PLTS	539,869	1	539,869	539,869	1	539,869
Total			9,387,710			9,387,710
Dehydrator						
Name	**Estimated unit cost (THB)**	**Qty (unit)**	**Estimated cost (THB)**	**Estimated unit cost (THB)**	**Qty (unit)**	**Estimated cost (THB)**
GNTWTRT.TSA	1,478,936	2	2,957,873	1,478,936	2	2,957,873
Total			2,957,873			2,957,873
Pump						
Name	**Estimated unit cost (THB)**	**Qty (unit)**	**Estimated cost (THB)**	**Estimated unit cost (THB)**	**Qty (unit)**	**Estimated cost (THB)**
P-SWSUBM	2,116,410	1	2,116,410	2,116,410	1	2,116,410
SWTTRT.P-DIST	4,123,410	2	8,246,820	4,123,410	2	8,246,820
SWTTRT.P-MELT	4,123,410	4	16,493,640	0	0	0
SWTTRT.P-REJECT	4,123,410	4	16,493,640	4,123,410	2	8,246,820
SWTTRT.P-RJWSWT	0	0	0	4,123,410	2	8,246,820

Table 51 (Continued)

Configuration	**Hydrocyclone + RO combination**			**Multi-stage Filter**		
Pump						
Name	**Estimated unit cost (THB)**	**Qty (unit)**	**Estimated cost (THB)**	**Estimated unit cost (THB)**	**Qty (unit)**	**Estimated cost (THB)**
SWTTRT.P-RO	Cost is included in the RO package			0	0	0
SWTTRT.P-RSWT	4,123,410	2	8,246,820	4,123,410	2	8,246,820
SWTTRT.P-SLURRY	4,123,410	2	8,246,820	4,123,410	2	8,246,820
Total			59,844,150			43,350,510
Compressor						
Name	**Estimated unit cost (THB)**	**Qty (unit)**	**Estimated cost (THB)**	**Estimated unit cost (THB)**	**Qty (unit)**	**Estimated cost (THB)**
GNTWTRT.BLW	6,234,000	1	6,234,000	6,234,000	1	6,234,000
GNTWTRT.BSC	58,362,000	1	58,362,000	58,362,000	1	58,362,000
Total			64,596,000			64,596,000
Gas Pipeline						
Name	**Estimated unit cost (THB)**	**Qty (unit)**	**Estimated cost (THB)**	**Estimated unit cost (THB)**	**Qty (unit)**	**Estimated cost (THB)**
GNTWTRT.1-TP12	4,714,800	1	4,714,800	4,714,800	1	4,714,800
GNTWTRT.2-TP24	2,750,300	1	2,750,300	2,750,300	1	2,750,300
GNTWTRT.3-TP36	2,750,300	1	2,750,300	2,750,300	1	2,750,300
GNTWTRT.4-TP43	1,298,850	1	1,298,850	1,298,850	1	1,298,850
GNTWTRT.5-TP51	23,372,000	1	23,372,000	23,372,000	1	23,372,000
GNTWTRT.6-TP36	51,418,400	1	51,418,400	51,418,400	1	51,418,400
GNTWTRT.MAINHDR	114,000,000	1	114,000,000	114,000,000	1	114,000,000
GNTWTRT.TP12-51	44,541,750	1	44,541,750	44,541,750	1	44,541,750
GNTWTRT.TP24-12	20,400,000	1	20,400,000	20,400,000	1	20,400,000
GNTWTRT.TP36-43	11,686,000	1	11,686,000	11,686,000	1	11,686,000
GNTWTRT.TP43-24	11,938,850	1	11,938,850	11,938,850	1	11,938,850
Total			288,871,250			288,871,250
Seawater Pipeline						
Name	**Estimated unit cost (THB)**	**Qty (unit)**	**Estimated cost (THB)**	**Estimated unit cost (THB)**	**Qty (unit)**	**Estimated cost (THB)**
PIL-SWT	3,750,000	2	7,500,000	3,750,000	1	3,750,000
RJPIPLN	300,000,000	2	600,000,000	300,000,000	1	300,000,000

Table 51 (Continued)

Configuration	Hydrocyclone + RO combination			Multi-stage Filter		
Seawater Pipeline						
Name	**Estimated unit cost (THB)**	**Qty (unit)**	**Estimated cost (THB)**	**Estimated unit cost (THB)**	**Qty (unit)**	**Estimated cost (THB)**
Total			607,500,000			303,750,000
Filter						
Name	**Estimated unit cost (THB)**	**Qty (unit)**	**Estimated cost (THB)**	**Estimated unit cost (THB)**	**Qty (unit)**	**Estimated cost (THB)**
SWTTRT.FILTER	48,381,000	1	48,381,000	48,381,000	2	96,762,000
Total			48,381,000			96,762,000
Hydrocyclone						
Name	**Estimated unit cost (THB)**	**Qty (unit)**	**Estimated cost (THB)**	**Estimated unit cost (THB)**	**Qty (unit)**	**Estimated cost (THB)**
SWTTRT.HCYCLONE	60,000,000	1	60,000,000	0	0	0
Total			60,000,000			0
RO PACKAGE						
Name	**Estimated unit cost (THB)**	**Qty (unit)**	**Estimated cost (THB)**	**Estimated unit cost (THB)**	**Qty (unit)**	**Estimated cost (THB)**
RO PACKAGE	1,537,647,000	1	1,537,647,000	0	0	0
Total			1,537,647,000			0
Tank						
Name	**Estimated unit cost (THB)**	**Qty (unit)**	**Estimated cost (THB)**	**Estimated unit cost (THB)**	**Qty (unit)**	**Estimated cost (THB)**
SWTTRT.BLENTANK	0	0	0	16,728,000	1	16,728,000
SWTTRT.BUFTANK	136,110,000	2	272,220,000	7,566,000	1	7,566,000
SWTTRT.MELTTANK	0	0	0	7,569,000	1	7,569,000
SWTTRT.ROBUFTK	15,765,000	5	78,825,000	0	0	0
SWTTRT.RJBUFFTK	17,664,000	2	35,328,000	16,728,000	1	16,728,000
Total			386,373,000			48,591,000

Table 51 (Continued)

Configuration	Hydrocyclone + RO combination			Multi-stage Filter		
Power Generator						
Name	**Estimated unit cost (THB)**	**Qty (unit)**	**Estimated cost (THB)**	**Estimated unit cost (THB)**	**Qty (unit)**	**Estimated cost (THB)**
GAS TURBINE GENERATOR	285,000,000	3	855,000,000	142,500,000	3	427,500,000
Total			855,000,000			427,500,000

As the result in Table 52, the operating cost included maintenance cost of all of main equipments and office works. Employee salary is estimated by using the estimated number of engineer in petrochemical in MTP. The number of employee is 500 staffs with the estimation from the big petrochemical plant in MTP. The miscellaneous cost in operating cost e.g. cost for office work and Corporate Social Responsibility (CSR) are included in this part as well. The total of operating cost called OPEX (Operating Expenditure).

With different configuration, the detail of OPEX is also different, especially for OPEX of seawater desalination plant. The cost of raw material, the waste gas is the only feedstock for the plant in both configurations. The cost of waste gas is assumed to 10 THB/MMBTU based on the LHV of production gas. This price = 10% of NG sale price in Thailand. The cost of treated gas is 2 THB/MMBTU based on the LHV of treated gas. The calculation is estimated by 20% of waste gas price. Therefore the net gas cost for feedstock = 8 THB/MMBTU. In addition, the treated gas is designed to use as fuel gas for power generation unit. The generated electricity is used for internal process and the excess margin will be sold back to PEA at 3.0 THB/kWh

In desalination process, the multi-stage filter is designed to use the raw water for melting the slurry ice; therefore, the cost of this water is the main OPEX for this configuration. The imported make-up raw water price is assumed to 10 THB/m^3 based on the price from East Water (Kaohoon, 2013) After treatment, raw water from desalination process is assume to sold back as raw water at price = 5 THB/m^3.

In case of hydrocyclone-RO combination configuration, there is no import raw water for the process. Only raw seawater is required for the melting purpose. However, the operation cost is 8.1 THB/m^3 (Table 13). The product of this configuration is drinking water and the sale price is designed to 500 THB/m^3. The current maximum drinking water price is approximate 9,300 THB/m^3 (OKnation, 2010). The margin between sale price and operating price is wide enough for the investment. Therefore, the assumption is possible for economic evaluation.

Table 52 Breakdown estimated cost of waste gas treatment plant and seawater desalination plant for hydrocyclone-RO and multi-stage configuration at 5 MTPA LNG feed scenario.

Parameter	Hydrocyclone-RO		Multi-stage filter	
	Normal HiT	**Emergency HiT**	**Normal HiT**	**Emergency HiT**
Flow rate; MMscfd	1.12	10.78	1.12	10.78
Wobbe; BTU/scf	893.50	893.50	893.50	893.50
HHV;BTU/scf	715.55	715.55	715.55	715.55
LHV;BTU/scf	645.50	645.50	645.50	645.50
SG (60/60)	0.64	0.64	0.64	0.64
Calc Wobbe; BTU/scf	894.02	894.02	894.02	894.02
Calc SG	0.64	0.64	0.64	0.64
Est. Power gen; MW	4.02	38.53	4.02	38.53
Internal usage; MW				
GT Plant	0.28	3.05	0.28	3.05
Desal. Plant	14.41	14.34	4.65	4.57
Total internal usage; MW	14.69	17.39	4.92	7.62
Misc power; MW	1.00	1.00	1.00	1.00
Margin power; MW	1.00	1.00	1.00	1.00
Grand Total power usage	16.69	19.39	6.92	9.62
Remain power for sale; MW	-12.67	19.14	-2.91	28.91
GT Plant income; THB/yr	-315,009,788	473,484,849	-72,453,369	716,036,910
Sale power, THB/yr	-314,801,254	475,485,421	-72,244,835	718,037,482
Waste Gas buying cost from local company; THB/yr	208,534	2,000,572	208,534	2,000,572

Table 52 (Continued)

Parameter	Hydrocyclone-RO		Multi-stage filter	
	Normal HiT	Emergency HiT	Normal HiT	Emergency HiT
Waste Gas buying cost from local company; THB/yr	208,534	2,000,572	208,534	2,000,572
Desal. Plant income; THB/yr	2,924,192,044	2,880,750,104	-45,816,097	-45,128,608
RO feed water; m3/yr	7,448,004	7,337,356	-	-
Sale Drinking water; m3/yr	5,942,229	5,853,951	-	-
RO operating cost; THB/yr	46,922,426	46,225,342	-	-
Sale Drinking water; THB/yr	2,971,114,470	2,926,975,446	-	-
Total income; THB/yr	2,609,182,256	3,354,234,952	-	-
Import raw water; m3/yr	-	-	15,044,785	14,819,925
Sale raw water; m3/yr	-	-	20,926,350	20,614,128
Import raw water; THB/yr	-	-	150,447,848	148,199,249
Sale raw water; THB/yr	-	-	104,631,752	103,070,641
Total income; THB/yr	-	-	-118,269,466	670,908,302
Estimated CAPEX				
Total equipment cost; THB	2,435,861,190	2,435,861,190	1,899,138,749	1,899,138,749
Total Land cost; THB	78,125,000	78,125,000	78,125,000	78,125,000
Total soil work cost; THB	280,000,000	280,000,000	280,000,000	280,000,000
Total concrete cost; THB	112,500,000	112,500,000	112,500,000	112,500,000
Total CAPEX	2,906,486,190	2,906,486,190	2,369,763,749	2,369,763,749
Margin 10%	290,648,619	290,648,619	236,976,375	236,976,375
Grand Total CAPEX; THB	3,197,134,809	3,197,134,809	2,606,740,124	2,606,740,124
Estimated OPEX				
Employee salary; THB	240,000,000	240,000,000	240,000,000	240,000,000
Maintenance cost; THB	10,000,000	10,000,000	10,000,000	10,000,000

Table 52 (Continued)

Parameter	Hydrocyclone-RO		Multi-stage filter	
	Normal HiT	**Emergency HiT**	**Normal HiT**	**Emergency HiT**
Misc; THB	10,000,000	10,000,000	10,000,000	10,000,000
Total OPEX	260,000,000	260,000,000	260,000,000	260,000,000
Margin 10%	26,000,000	26,000,000	26,000,000	26,000,000
Grand Total OPEX;THB/yr	286,000,000	286,000,000	286,000,000	286,000,000

According to Table 52, the results show that the total CAPEX of multi-stage filter configuration is less than hydrocyclone-RO configuration about 600 MMTHB. This value is close to the different of main equipment cost because the other CAPEX is the same. The OPEX of both configurations show the same value because the cost of desalination plant OPEX is separated in income calculation. Therefore, the number of total OPEX is salary, maintenance cost and management cost as mention prior. The treated gas flow rate of these configuration is the same because the result base on the same scenario both waste gas feed and LNG feed flow rate. The different between these configurations are the detail of desalination plant OPEX. The operating cost of hydrocyclone-RO combination configuration is operating cost of RO and the price of drinking water. Multi-stage filter configuration, the import raw water price and price of sale back raw water. These figures lead the result of economic evaluation is totally different as shown in Table 53.

Table 53 Comparison of estimated payback period of the each configuration at 3 and 5 MTPA LNG feed scenarios.

Configuration	5 MTPA		3 MTPA	
	Normal HiT	**Emergency HiT**	**Normal HiT**	**Emergency HiT**
Hydrocyclone-RO combination	1.44	1.12	2.56	1.62
Multi-stage filter	Negative	6.87	Negative	6.43

7. Preliminary economic evaluation and sensitivity analysis

Although the total investment cost of hydrocyclone-RO configuration is higher than multi-stage configuration, the payback period of this configuration is better than multi-stage filter configuration because of the operating cost (Table 53). The cost of imported water is very high if comparing to RO operating cost. The cost of imported water is 10 THB/m^3 but the sale price of the product water is 5 THB/m^3. It means that the operating cost of multi-stage filter configuration is 5 THB/m^3. On the other hand, the operating cost of RO is 6.3 THB/m^3 but the drinking water product is 500 THB/m^3. This margin gives faster payback period. The payback period of hydrocyclone-RO configuration is 1.12-1.44 years depending on waste-gas flow rate. With same waste gas feed, multi-stage filter configuration gives loss in investment every year of operation in case normal flow operation. Although feed of LNG is reduce to turndown load, the payback period of hydrocyclone-RO configuration also gives the payback period in 1.62-2.52 years. In actual, the flow rate of waste gas is flow as normal rate, therefore, the multi-stage filter is not suitable as alternative selection. In Table 54, hydrocyclone-RO configuration is selected for sensitivity analysis

Table 54 Sensitivity analysis at various parameter of hydrocyclone-RO combination configuration at 3 and 5 MTPA LNG scenarios

LNG load (MTPA)	Case	Treated gas flow rate	Electricity price	Waste gas price	RO operating cost	Sale drinking water price
		MMSCFD	THB/kWh	THB/MMBTU	THB/m^3	THB/m^3
3	1	3.20	3.00	10.00	262.65	500.00
	2	3.20	-5,755.96	10.00	6.30	500.00
	3	3.20	3.00	15,367.22	6.30	500.00
	4	1.12	3.00	10.00	225.32	500.00
	5	1.12	3.00	10.00	6.30	225.48
	6	1.12	3.00	37,670.01	6.30	500.00
	7	1.12	5.00	100.00	262.65	579.17
	8	0.00	5.00	100.00	148.00	476.85
5	1	4.98	3.00	10.00	317.10	500.00
	2	4.98	-7,469.18	10.00	6.30	500.00
	3	4.98	3.00	19,935.81	6.30	500.00
	4	1.12	3.00	10.00	275.26	500.00
	5	1.12	3.00	10.00	6.30	162.88

Table 54 (Continued)

LNG load (MTPA)	Case	Treated gas flow rate	Electricity price	Waste gas price	RO operating cost	Sale drinking water price
		MMSCFD	THB/kWh	THB/MMBTU	THB/m^3	THB/m^3
	6	1.12	3.00	77,112.80	6.30	500.00
	7	1.12	5.00	100.00	317.10	588.18
	8	0.00	5.00	100.00	148.00	401.15

Table 54 shows the variation of cost impact parameter in operation. There are 5 parameters:

- Treated gas flow rate: It relates to the treated gas rate. The treated gas is used as the fuel gas for power generation. The analysis consider on optimum flow, normal flow and no flow case. The optimum condition mean the treated gas flow rate given no profit of waste gas treatment plant.

- Electricity price: Electricity, which is generated by gas turbine generator from treated gas, is used as main power for electrical load user in the waste gas treatment process and seawater desalination process such as pump. The excess electricity after deducting from internal usage will be sold to PEA. On the other hands, in case of lack of power generation, the electricity has to be imported from PEA with the same price. The possible electricity cost is assumed at 5 THB/kWh

- Waste gas price: In base case, waste gas price is assumed to 10% of NG, therefore, if the NG price is increased from the current price, the operating cost of gas treatment plant will be increased. The possible maximum gas price is 100 THB/MMBTU.

- RO operating cost: The operating cost of RO depends on RO elements. The cycle period of membrane element depends on TDS of feed water. In base case, TDS is 9,700 mg/l and the operating cost is 8.1 THB/m^3. However, the feed seawater TDS and salinity cannot be predicted accurately or assume as constant along the time because it depends on global nature. It may change or not change. The maximum figure of TDS considers at 170,000 mg/l and the operating cost is 148 THB/m^3.

- Sale drinking water price: In base case, this price is about 500 THB/ m^3. However, the price may change because of the drought in the future.

By the reasons above, the sensitivity cases are list as below:

- Case 1: Optimum treated gas flow rate and vary RO operating cost that give payback period in 10 years.

- Case 2: Optimum treated gas flow rate and vary electricity price that give payback period in 10 years.

- Case 3: Optimum treated gas flow rate and vary waste gas price that give payback period in 10 years.

- Case 4: Normal treated flow rate and vary RO operating cost that give payback period in 10 years.

- Case 5: Normal treated gas flow rate and vary drinking water price that give payback period in 10 years.

- Case 6: Normal treated gas flow rate and vary waste gas price that give payback period in 10 years.

- Case 7: Normal treated gas flow rate, possible maximum electricity price, maximum waste gas price, maximum RO operating cost in case 1 – 6 and vary drinking water price that give payback period in 10 years.

- Case 8: No treated gas flow rate, possible maximum electricity price, maximum waste gas price, possible maximum RO operating cost in case 1 – 6 and vary drinking water price that give payback period in 10 years.

Table 54 shows that the variation value of parameter in each case is over the possible maximum either normal or optimum treated gas flow rate. The major profit part is the drinking water price. Although there is no treated gas produced from the gas treatment system, overall system can operate with profit and get return in 10 years. In addition, in worse case, the overall can operate with profit to give the return in 10 years while all parameters reach possible maximum figure. Therefore, hydrocyclone-RO combination configuration is suitable for implementation because the system can operate either there is no waste gas feed or not. In additional, it can make the profit and give return in 1.5 years.

CONCLUSIONS

According to the current emission problem in MTP, the waste energy recovery and utilization of LNG terminal by integrated system was proposed to address this particular solution. The simulation and design shows that the proposed system is not only able to reduce MTP emission by freeze desalination but is also feasible for economic point of view.

The result shows that VOCs content in treated gas has qualified the international air pollution standard. The removal efficiency is likely either normal or emergency scenarios: 100.00, 99.65, 100.00, 100.00, 96.56 and 55.90 % for C_{5+}, LPG, BTEX, VOCs, sulfur compounds and CO_2, respectively. In addition, some of by-product of the integrated system is able to sale as the low quality feedstock or fuel.

The eutectic solution concept is applied in freeze desalination of seawater. The result shows the real seawater from Gulf of Thailand is possible to apply to freeze desalination technique in the integrated system. Temperature at -16 ^{o}C is selected as the best operating condition of the system. In addition, the system is also able to operate under wide range of feed seawater salinity.

With many different types of process configuration, the hydrocyclone-RO combination configuration is selected as the proposed in the integrated system. The CFD result shows that modified hydrocyclone can separate the ice-brine mixture at about 80% efficiency. The product of this system is raw drinking water (approximately 720 m^3/h), which qualifies the salinity and TDS standard drinking water specifications.

The estimated total CAPEX of whole system is about 3,200 MMTHB. The main equipment cost is about 82% of total cost. However, the system can be operated with 10 years payback period criteria whether the availability of waste gas feed or the lacks of waste gas feed. As per the based case, the payback period 1.45 and 1.50 years in case normal and emergency flow scenarios. Therefore, the proposed designed process in the study is feasible for real emission situation at MTP. In addition, the sensitivity analysis also shows that the payback period of this system is feasible in 10 years although the involved parameters; such as treated gas flow rate, waste gas price, electricity price, operating cost of RO and drinking water price, are under variation of the extreme range.

RECOMMENDATIONS

1. The detail analysis of freezing point of seawater should be checked and compared to the simulation model if this project is further considered to study in basic of design stage.

2. Since there is some excess cold energy of NG stream; therefore, it can be utilized for other applications in the future e.g. the reduction of inlet air temperature of air cooler and the reduction refrigerant temperature in air conditioning system in office. It would be studied to maximize the energy recovery of the integrated process.

3. There are many methods for freeze-desalinated product handling such as to distribute ice directly to fresh market or fishery for fresh food preservation. However, the considerations of TDS and salinity for the product have to be concerned.

4. The pinch analysis is not included in the scope of study. Therefore, the re-investigation of the integrated system with pinch analysis would give the maximum heat recovery efficiency.

5. In case of real implementation, the components of waste gas impact to the operating condition of LTS section, potential of treated gas utilization and economic evaluation model. Hence, the component types from each waste gas source have to be re-analyzed to ensure the range of process operation and economic of the system. The re-investigation of the component type also included the new comer of the waste gas source in the future.

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APPENDIX

The information is contained in CD. The content is below:

1. Stream summary for waste gas treatment plant and seawater desalination plant for all scenarios
2. Stream summary for waste gas pipeline network for all scenario
3. Equipment summary list
4. Equipment process specification
5. Main equipment cost list
6. Economic evaluation for 3 MTPA LNG scenario and 5 MTPA LNG scenarios
7. Sensitivity analysis for 3 MTPA LNG scenario and 5 MTPA LNG scenarios

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