Design and Economic Analysis of Boil-Off Gas Recovery in LNG Facilities

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Abstract: This study focuses on designing and evaluating a process plant for liquefied natural gas (LNG) boil-off gas (BOG) recovery. The lightest hydrocarbons included in LNG, such as methane and ethane, are often included in Boil-off Gas (BOG). Flaring and contamination of the environment are unavoidable in the absence of an effective BOG recovery system. Using Aspen HYSYS, a natural gas liquefaction process was simulated, emphasizing the recovery and utilization of BOG generated during various stages of LNG processing, including liquefaction, depressurization, storage, and shipping. The material and energy balances for the process were meticulously calculated to ensure accuracy in flow rates and heat exchange efficiencies. The simulation results indicate that the liquefied natural gas produced contains a methane 2473oncentrationn of 96.64% with minor amounts of ethane. BOG, mainly consisting of methane (100% purity), was effectively recovered and conditioned for reuse or flaring. An economic analysis was conducted to assess the profitability of BOG recovery, highlighting an estimated annual income of $138,121,200, with a gross profit margin of 97.3%. The total capital investment required for BOG recovery equipment amounted to $3,790,605. This study demonstrates that BOG recovery can significantly enhance the economic viability and environmental sustainability of LNG operations by reducing methane emissions and providing a valuable energy resource.

Keywords: *Liquefied Natural Gas (LNG); LNG Process Plant; Boil-Off Gas (BOG); BOG Recovery; Economic Analysis; Environmental Sustainability*.

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I. INTRODUCTION

Worldwide, the ability to produce liquefied natural gas (LNG) is also increasing, as the demand for energy continues to rise every day [1]. Liquefied natural gas is mostly composed of methane, with a small amount of ethane that has been cooled for transportation and safety purposes. The most cost-effective method of transporting natural gas across long distances is in its liquefied form, which is 600 times smaller than the gas itself [2]. But its boiling point is lower than -161 ºC.

Even with meticulous insulation, heat may escape from LNG due to the temperature differential with the surrounding environment [1]. Some of the LNG evaporated due to the heat leak; the resulting gas is known as boil-off gas (BOG). Regular venting is necessary to keep the LNG carriers from being over pressured. The lightest hydrocarbons included in LNG, such as methane and ethane, are often included in the BOG. Flaring and contamination of the environment are unavoidable in the absence of an effective BOG recovery system [3].

As a result of variations in pressure and temperature inside the tanks, BOG is an inevitable part of transporting and storing LNG [4]. Without proper treatment, the ever-increasing concentration of vaporised gases caused by LNG's continual evaporation poses a significant threat to both public health and the economy. Despite flaring's effectiveness in managing pressure, it contributes to the emission of greenhouse gases, which undermines the sustainability goals of the LNG industry [5]. Historically, many LNG terminals have used flaring to eliminate excess BOG [2]. The emission of carbon dioxide and other pollutants into the air makes this process particularly harmful to the environment, since it contributes to climate change and goes against the ever-tightening environmental restrictions. Among these difficulties, the need for better BOG management to reduce flaring and its environmental effects is becoming more apparent [6].

This study aims to design and perform an economic analysis of boil-off gas recovery in LNG facilities. The objectives of the study are to design a natural gas liquefaction process plant using Aspen HYSYS, perform energy and material balance calculations, and carry out economic analysis to determine the feasibility of the proposed process.

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The importance of the study is that it is likely to contribute to the efficient handling of BOG for either energy or environmental industries. These findings shall further benefit many major stakeholders that include the oil and gas industry, policymakers, environmentalists, and a number of communities living around the export terminals. This study therefore contributes to enhancing good handling of BOG, thereby helping further reduce flaring, which generally has been mostly condemned due to environmental and economic downsides.

II. MATERIALS AND METHOD

- *Study Area*

This research focused on the berth section of the Nigeria Liquefied Natural Gas (NLNG) plant where the produced liquefied natural gas (LNG) is loaded for transportation to other areas.

- *Materials*

The materials listed below are used in the Aspen HYSYS software to achieve the aim and objective of the research work.

Aspen HYSYS; component splitter, separator, LNG heat exchangers, compressors, coolers, valves, tanks, pumps, tee (header), natural gas.

- *Simulation Software (HYSYS) Description and Basis*

The entire process was designed using the Soave-Redlich Kwong cubic equation of state, and the software used for this purpose was Aspen Hysys version 8.6. Throughout this program, constant state was taken for granted.

The Soave-Redlich-Kwong equation of state;

$$P=RT/(V-b)-a/(V(V+b)) \quad (1)$$

a and b are defined as follows;

$$b = 0.08664\, RTc / Pc$$

$$a = 0.42748\, ((RTc)^2)/Pc\, [1+m(1-\sqrt{T\_r})]2$$

$$Tr = T/Tc$$

$$m = 0.480 + 1.574\omega - 0.176\omega 2$$

P = pressure, T = temperature, V = volume, R = Gas constant, Tr = Reduced temperature, Tc = Critical Temperature, Pc = Critical Pressure, ω = Acentric factor

- *Feed Data*

About 95% of the input data used were obtained from the Nigeria Liquefied Natural Gas Limited (NLNG) facts and figures 2018, as shown in tables 1 and 2 below.

Table 1 Natural Gas Feed Condition

Pressure [kPa]	5000
Temperature [°C]	25
Molar Flow [kgmole/h]	1.352×10^{5}
Mass Flow [kg/h]	2.511×10^{6}
Vapour/Phase Fraction	0.9842

Table 2 Natural Gas Feed Molar Composition

Components	Mole fraction
Methane	0.883
Ethane	0.0551
Propane	0.0080
n-Butane	0.0075
i-Butane	0.0057
C_{5+}	0.0090
H_2O	0.0166
CO_2	0.0151

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- *Process and Equipment Description*

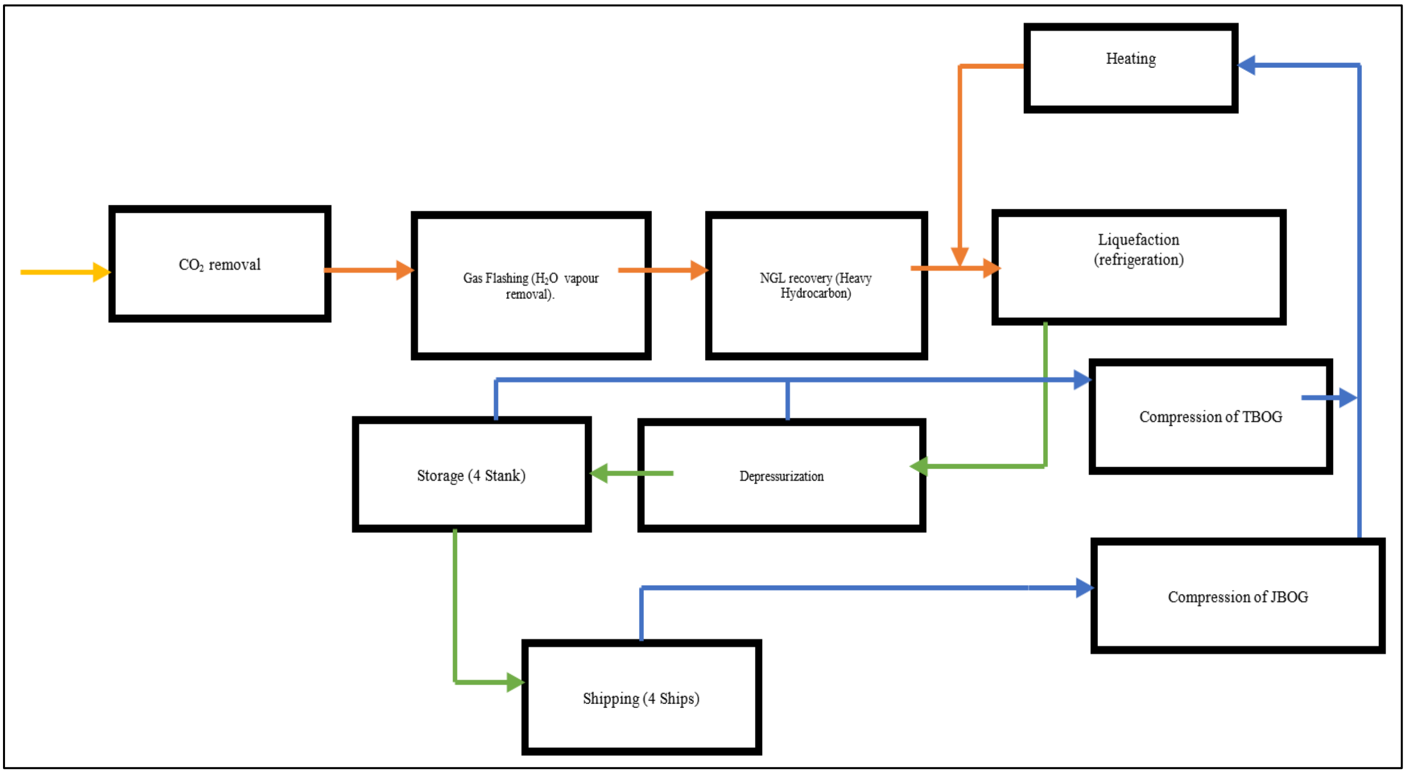

Fig 1 The Block Diagram Showing the Stages Involved in the Liquefaction and BOG Recovery

The feed (natural gas) stream was sent into a gas sweetening section to remove CO_2 and then for dehydration to exclude water, which is represented in the Aspen HYSYS design as a separator to flash out the water from the gas. In the dehydration unit, water comes out as a bottom product, and the gas evacuates the separator as an overhead. The gaseous product from the dehydration unit is sent to an NGL (Natural Gas Liquid) unit modelled with a component splitter. The NGL recovery unit removes the natural gas liquid as a bottom product, and the overhead product containing methane and ethane was sent to the refrigeration section for liquefaction.

The inlet gas temperature was reduced to -162°C by the refrigeration process. At this temperature, all the gas stream has changed to liquid, now called liquefied natural gas (LNG).

The liquefied stream was sent through a valve where the Joule Thompson effect occurred to reduce the pressure and further a flash drum to remove any form of vapour that might have arisen during pressure reduction and to prevent pump cavitation.

The produced LNG was pumped to the four storage tanks and lastly to the carriers. From the LNG KO drum, storage tanks, and the carrier, the overhead products, which are boil-off gas, are collected and sent back through compressors to the refrigeration section of the plant for re-liquefaction.

From the block diagram, the red line signifies raw gas, the green line is liquefied natural gas (LNG), while the blue line is boil-off gas.

- *Water Removal Unit (Flash Drum)*

The two-phase separator was used to remove the water vapour from the inlet gas stream (that is dehydration) to prevent corrosion of the equipment and hydrate formation that could have led to blockage of the pipe and damage of the equipment.

- *NGL Removal Unit*

The Aspen Hysys component splitter was used to separate the heavy hydrocarbons (propane down to C_5+) and some quantity of ethane from the gas stream that exited the two-phase separator. This was done to prevent freezing and hydrate formation by heavy hydrocarbon at cryogenic temperature.

- *CO_2 Removal Unit*

In the base case simulation, a component splitter removed the $CO_2$from the gas stream going to the refrigeration system. The CO_2 was removed to prevent reduction of heating value of the gas when regasification for combustion purposes has occurred.

- *Refrigeration Inlet Header*

This header received the inlet gas from the CO_2 removal unit and the BOG from the storage, depressurization and jetty section of the LNG plant and sent it through the refrigeration unit for liquefaction purpose. At this header, the stream of gas is unified to ensure suitable pressure and temperature of both streams

- *Refrigeration Unit*

The process employs a three-stage cascade refrigeration system that uses pure refrigerants such as methane, ethane, and propane to cool the inlet gas to a liquid state at the required

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temperature. The gas passes through Propane, then ethane, and finally methane.

➢ *Pressure Relieve Valve*

In order to transfer the LNG to the flash drum at a pressure lower than the entering LNG pressure, this valve was used to lower the LNG's pressure. The procedure made use of a fail shut valve.

The mathematical equation used by HYSYS is expressed as:

$$\mu_{JT} = \left(\frac{\partial T}{\partial P}\right)_H = \frac{V}{C_P}(\propto T - 1) \quad (2)$$

Where:

μ_{JT} *= J-T coefficient expressed in °C/bar (SI units: K/Pa)*

C_p *= Heat capacity at constant pressure*

V = Gas volume

$\left(\frac{\partial T}{\partial P}\right)_H$ *= Temperature and pressure gradient*

$\propto$ *= Coefficient of thermal expansion*

In J-T valve expansion the enthalpy remains constant. The enthalpy H, is defined as:

$$H = U + PV \quad (3)$$

Where:

U = Internal energy;

P = Pressure;

V = Volume

➢ *LNG Flash Drum*

Before transporting LNG to the storage tank, this drum ensured that the stream was free of any vapour that may have caused the pump to cavitate. The flashing happened because the incoming gas pressure was higher than the flash drum's operating pressure, which is just under 1 bar. The flashing method resulted in the recycling of boil-off gas.

➢ *LNG Pump and Header to Storage Tank*

The four storage tanks were supplied with the generated LNG via this header, which was pumped from the flash drum by means of the pump, which raised the pressure. Because the LNG's entrance pressure was lower than atmospheric, the pump was unable to transport the fuel far enough to reach its final destination for storage.

The head and discharge pressure of the pump was obtained with the equations;

$$H = \frac{(Q \times \rho \times g)}{\varepsilon \times p} \quad (4)$$

$$P = H \times \rho \quad (5)$$

Where;

Q = flowrate (m3/sec),

ρ = density in kg/m3,

g = acceleration due to gravity (m/s2),

p = power (kW),

H = head (m),

P = Pressure (kPa)

➢ *LNG Storage Tank*

Prior to being transported to the jetty section, the LNG was stored in four spherical tanks. The storage tank's design parameters also resulted in boil-off gas, which was subsequently recycled. The round storage tank served to alleviate the problem of liquid retention after discharge.

The volume of the storage tank was obtained using the equation below;

$$V = \frac{1}{3}\pi h^2 \,(3r - h) \quad (6)$$

$$r = \frac{d}{2} \quad (7)$$

d = diameter (m),

r = radius (m),

h = height (m) and

v = volume (m)

➢ *LNG to Jetty Pumps*

The LNG was transported from the four storage tanks to the LNG ships using a four-jetty LNG centrifugal pump. Each pump is designed to operate at 1800 rpm and 150 KW of electricity. The same mathematical equation used in equation 3 and 4 is applied to jetty pumps.

➢ *LNG Carrier*

The pumped LNG was transferred from the storage tank to four round LNG carriers. With cryogenic qualities, these carriers are intended to remain in a liquid state below 1 atm. Therefore, the receiving vessel's pressure differs from the pump's supplied pressure. The jetty area of the facility is where you may find these carriers. Flash calculations which involve the equations below was used by HYSYS.

$$y_i = kx_i \quad (8)$$

$$\Sigma kx_i = 1 \quad (9)$$

$$Fz_i = Vy_i + Lx_i \quad (10)$$

➢ *Jetty BOG Header*

The jetty boil-off gas header is a collection manifold that received all the boil-off gas the arose from each spherical LNG

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carrier and then to the compression section and temperature and pressure were raised for recycling purpose.

➢ *Jetty BOG Compressor*

This centrifugal compressor raised condition (T and P) of the jetty boil-off gas from its inlet condition before it mixed with the boil-off gas from KO and storage. Reliquefication.

The mathematical correlation for computing the process condition of the compressor is displayed as:

$$\Delta T_{1_{isentropic}} = \frac{T_1 \times [\left(\frac{P_2}{P_1}\right)^{\frac{k-1}{k}} - 1]}{\tau_{isentropic}} \tag{11}$$

$$T_2 = T_1 + \Delta T_{isentropic} \tag{12}$$

$$\Delta T_{polytropic} = T_1 \times \frac{\left[\left(\frac{P_2}{P_1}\right)^{\frac{n-1}{n}} - 1\right]}{\tau_{polytropic}} \tag{13}$$

$$T_2 = T_1 + \Delta T_{polytropic} \tag{14}$$

Where;

T_2 = discharge T (k)

T_1 = suction T (k)

P_2 = discharge P (kPa)

P_1 = suction P (kPa)

$\tau_{polytropic}$ = polytropic efficiency in decimal

$\tau_{isentropic}$ = isentropic efficiency expressed in decimal

k = isentropic exponent: c_p/c_v (ratio of the specific heat capacity)

n = polytropic exponent: $\frac{n}{n-1} = [\frac{k}{k-1}]$

➢ *Flash Drum BOG Header and Storage*

All BOG in the LNG processing sector, except the jetty, were gathered by the header. The KO drum and storage tank BOG are unified in the header. The storage BOG was tagged as TBOG while the KO drum BOG was referred to as DBOG.

➢ *TBOG AND DBOG Compressor*

This centrifugal compressor elevated the condition of TBOG and DBOG from their inflow conditions to the state that was adequate for the recycle header. See figure 16 for pictorial view.

➢ *Recycled BOG Header*

The old BOG header was used as a collector for the new BOG coming from the jetty section (JBOG), the storage tank (TBOG), and the flash drum (DBOG).

➢ *Sizing of Equipment*

The equipment sizes were determined by using the Aspen HYSYS program to calculate the volumes, diameters, and heights of the storage tanks, ships, and BOG knock-out drums.

➢ *Rating of Equipment*

A number of process equipment parameters, including motor compressor and pump heat duties, pressure, and temperature, were determined through grading.

➢ *Cost Estimation*

Cost estimation of compressors at the depressurizing (KO and valve) and jetty section were performed to know the capital cost, installation and operating cost. The capital, operating and installation cost were used to check the feasibility of the research.

III. RESULTS AND DISCUSSION

The three portions where the boil-off gas was detected in the LNG modelling processes related to liquefaction, depressurisation, storage, and shipping are;

- The knock-out drum for LNG depressurisation
- Storage tanks
- Shipping section

➢ *Refrigeration Unit*

The refrigeration unit results indicate the characteristics of the LNG produced during the liquefaction process. Table 3 shows the process conditions of the LNG produced. The phase fraction is entirely liquid (0.0000), with a temperature of **-162°C** and a pressure of **3150 kPa**. The molar flow and mass flow rates are **1.229×10^5 Kgmole/h** and **2.002×10^6 kg/h**, respectively.

The LNG stream's composition is predominantly methane, accounting for **96.64%** of the mole fraction, with the remaining **3.36%** being ethane (Table 4). The absence of heavier hydrocarbons and impurities (like CO_2 and H_2O) highlights the LNG stream's purity, which is essential for achieving efficient cryogenic conditions.

➢ *J-T Valve*

The check valve used was opened 50%. The Joule-Thomson (J-T) valve plays a key role in reducing the pressure of the LNG stream. Table 5 indicates that the pressure drops from **3150 kPa** to **1500 kPa**, while the temperature slightly increases from **-162°C** to **-161.4°C**. The outlet stream's composition remains almost identical to the inlet stream, with methane making up **98.18%** and ethane **1.82%** of the mole fraction. This shows the valve's effectiveness in controlling pressure with minimal phase changes or losses in methane concentration.

➢ *Knock-Out Drum*

The knock-out (KO) drum separates the liquid and vapor phases after the pressure drop through the J-T valve. Table 6 shows the results from the KO drum. The liquid outlet stream retains the majority of the mass flow (1.996×10^6 kg/h), while the vapor outlet (representing BOG) only accounts for **6484.93 kg/h**, with a phase fraction of **1.0000**. The methane composition in the vapor outlet reaches **100%**, indicating that the BOG primarily consists of methane, which will require recovery.

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- *Storage Pump*

Using Hysys, the pressure differential between the suction and discharge pressures was used to determine the pump heat duty, which was found to be 1.863×106 kJ/h (517.5KW).

The storage pump increases the pressure from **100 kPa** to **400 kPa** without a significant temperature change. As indicated in Table 7, the mass flow remains constant at **1.996×10^6 kg/h**, with the pump having a duty of **517.5 kW**. The methane and ethane compositions remain stable across the pump, showing efficient operation without inducing significant changes to the LNG's composition.

- *LNG Header to Storage*

The four storage tanks are supplied with LNG in an equitable distribution by the LNG header.. According to Table 8, each tank receives approximately **4.989×10^5 kg/h** of LNG, and the header maintains consistent temperature (**-161.2°C**) and pressure (**400 kPa**). The even distribution ensures stable operations and mitigates the risk of pressure imbalances in the storage tanks.

- *LNG Storage Tank*

There is a total of 84,200 cubic meters inside each tank. All four tanks provide the same outcome. Table 9a and Table 9b present the results of the storage tanks, each with a capacity of **84,200 m³**. The results are uniform across all tanks, with minor temperature and pressure variations. The BOG production from each tank is **888.2 kg/h**, and the vapor outlet consists entirely of methane (100%). The storage tanks effectively maintain the LNG in liquid form, with minimal BOG production, ensuring the safety and efficiency of the process.

- *LNG to Jetty Pumps*

There are four pumps to jetty section which supplies LNG to LNG carrier ships. From the simulation process, each pump was given the same rating and it was observed to have the same readings as shown below. Table 10 details the operation of the jetty pumps, which transfer LNG to the carrier ships. The pumps raise the pressure from **100 kPa** to **900 kPa** while maintaining a stable temperature (**-160.9°C**) and flow rate (**4.981×10^5 kg/h**). The methane and ethane compositions remain unchanged during this process, ensuring consistent LNG quality.

- *LNG Carrier at Jetty*

The spherical LNG carrier is a cryogenic container. The LNG carriers receive the LNG from the jetty pumps, and Table 11 shows the stored LNG's parameters. The temperature in the carrier remains around **-161.3°C**, while the BOG outlet has a flow of **2373 kg/h**. This is expected due to heat ingress during transportation. The stored LNG in the carrier retains a high methane composition (**98.16%**), ensuring its suitability for shipping and end-use applications.

- *Jetty Boil Off Gas Header*

All the BOG from the jetty area were collected here. With the entrance streams set to their lowest pressures, the mixer's exit stream was adjusted to match.

The BOG collected from the ships is funnelled into the jetty BOG header. Table 12 indicates the total mass flow of **9490.28 kg/h** from all ship tanks, with a constant methane composition of **100%**. This stream is compressed for further use or flaring.

- *Jetty BOG Compressor*

The motor compressor with a duty of 1120.52kW was utilised to boost the temperature and pressure of the jetty boil-off gas before the general BOG header.

The jetty BOG compressor raises the BOG pressure from **100 kPa** to **3500 kPa**, and the temperature is increased to **50.75°C** (Table 13). The compressor handles a mass flow of **9490.28 kg/h**, ensuring that the BOG is properly conditioned for reinjection or flaring.

Table 3 Process Condition of LNG Result

Fraction of Phase	**0.0000**
Temperature in °C	-162.0
Pressure (kPa)	3150
Molar flow (Kgmole/h)	1.229×10^5
Mass Flow (kg/h)	2.002×10^6

Table 4 LNG Stream Composition.

Stream Components	**Mole fraction**	**Mass Fraction Percentage**
Methane	0.9664	96.64
Ethane	0.0336	3.3600
Propane	0.0000	0.0000
n-Butane	0.0000	0.0000
i-Butane	0.0000	0.0000
C_{5+}	0.0000	0.0000
H_2O	0.0000	0.0000
CO_2	0.0000	0.0000
Total	1.0000	100.00

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Table 5 J-T Valve Results

Parameter	Inlet stream (LNG)	Outlet stream (TODEP)
Phase	0.0000	0.0000
Temperature (°C)	-162.0	-161.4
Pressure (kPa)	3150	1500

Table 6 J-T Valve Product Composition

Product Component	Mole fraction
Methane	0.9818
Ethane	0.0182
Others	0.0000
Total	1.0000

Table 7 Result of the knock-out (KO) drum

Parameter	Inlet (from JT Valve)	Liquid Outlet (to pump)	Vapour Outlet (BOG)
Phase fraction	0.0000	0.0000	1.0000
Temperature (°C)	-161.36	-161.31	-161.31
Pressure (kPa)	1500.00	100.00	100.00
Molar flow (kgmole/h)	1.229×10^{5}	1.224×10^{5}	404.213
Mass flow (Kg/h)	2.002×10^{6}	1.996×10^{6}	6484.93
Component	**Mole fraction**	**Mole fraction**	**Mole fraction**
Methane	0.9818	0.9817	1.0000
Ethane	0.0182	0.0182	0.0000
Total	1.0000	1.0000	1.0000

Table 8 Storage Pump Result

Parameter	Inlet to pump	Outlet to storage header
Vapour	0.0000	0.0000
Temperature [C]	-161.31	-161.16
Pressure [kPa]	100.00	400.0
Molar Flow [kgmole/h]	1.224×10^{5}	1.224×10^{5}
Mass Flow [kg/h]	1.996×10^{6}	1.996×10^{6}
Heat Flow [kJ/h]	-1.102×10^{10}	-1.102×10^{10}
Component	**Mole fraction**	**Mole fraction**
Methane	0.9817	0.9817
Ethane	0.0183	0.0183
Total	1.0000	1.0000

Table 9 LNG Header to Four Storage Tanks

Name	Inlet to header	To Storage 1	To Storage 2	To Storage 3	To Storage 4
Vapour	0.0000	0.0000	0.0000	0.0000	0.0000
Temperature [C]	-161.2	-161.2	-161.2	-161.2	-161.2
Pressure [kPa]	400.00	400.00	400.00	400.00	400.00
Molar Flow [kgmole/h]	1.224×10^{5}	3.061×10^{4}	3.061×10^{4}	3.061×10^{4}	3.061×10^{4}
Mass Flow [kg/h]	2.002×10^{6}	4.989×10^{5}	4.989×10^{5}	4.989×10^{5}	4.989×10^{5}
Heat Flow (KJ/h)	-1.102×10^{10}	-2.755×10^{9}	-2.755×10^{9}	-2.755×10^{9}	-2.755×10^{9}

Table 9a LNG storage tank result

Name	To Storage 1 (Inlet)	Liquid in tank	BOG from tank
Vapour	0.0000	0.0000	0.0000
Temperature [C]	-161.2	-161.3	-161.3
Pressure [kPa]	400.00	100.00	100.00
Molar Flow [kgmole/h]	3.061×10^{4}	3.061×10^{4}	55.36
Mass Flow [kg/h]	4.989×10^{5}	4.981×10^{5}	888.2
Std Ideal Liq. Vol. Flow [m3/h]	1658	1655	3.000
Molar Enthalpy [kJ/kgmole]	9.001×10^{4}	9.001×10^{4}	-8.128×10^{5}
Molar Entropy [kJ/kgmole-C]	75.61	75.57	75.61
Heat Flow (KJ/h)	-2.755×10^{9}	-2.755×10^{9}	-2.755×10^{9}

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Table 9b Molar composition of LNG in the storage tank

Component	Liq. Mole fraction	BOG Mole Fraction
Methane	0.9817	1.0000
Ethane	0.0183	0.0000
Total	1.0000	1.0000

Table 10 Jetty pumps result

Parameter	An inlet to the jetty pump	Outlet to the LNG carrier
Vapour	0.0000	0.0000
Temperature [C]	-161.3	-160.9
Pressure [kPa]	100.00	900.0
Molar Flow [kgmole/h]	3.056×10^4	3.056×10^4
Mass Flow [kg/h]	4.981×10^5	4.981×10^5
Std Ideal Liq. Vol. Flow [m3/h]	1655	1655
Molar Enthalpy [kJ/kgmole]	-9.003×10^4	-8.998×10^4
Molar Entropy [kJ/kgmole-C]	75.57	75.66
Heat Flow [kJ/h]	-2.751×10^9	-2.750×10^9
Component	Mole fraction	Mole fraction
Methane	0.9817	0.9817
Ethane	0.0183	0.0183
Total	1.0000	1.0000

Table 11 LNG Carrier result

Parameter	Outlet to the carrier	Stored LNG in the Ship	Vapour Outlet (BOG)
Vapour	0.0000	0.0000	1.0000
Temperature [C]	-160.9	-161.31	-161.31
Pressure [kPa]	900.0	100.00	100.00
Molar Flow [kgmole/h]	3.065×10^4	3.041×10^4	147.9
Mass Flow [kg/h]	4.981×10^5	4.957×10^5	2373
Molar Enthalpy [kJ/kgmole]	-8.998×10^4	-9.003×10^4	8.128×10^4
Molar Entropy [kJ/kgmole-C]	75.66	75.57	150.26
Heat Flow [kJ/h]	-2.750×10^9	-2.738×10^9	-1.202×10^7
Component	Mole fraction	Mole fraction	Mole fraction
Methane	0.9817	0.9816	1.0000
Ethane	0.0183	0.0184	0.0000
Total	1.0000	1.0000	1.0000

Table 12 Jetty BOG header result

Parameter	Ship 1 Tank 1 (BOG)	Ship 1 Tank 2 BOG	Ship 1 tank 3 BOG	Ship 1 Tank 4 BOG	BOG to J-Compressor
Vapour	1.0000	1.0000	1.0000	1.0000	1.0000
Temperature [C]	-161.31	-161.31	-161.31	-161.31	-161.31
Pressure [kPa]	100.00	100.00	100.00	100.00	100.00
Molar Flow [kgmole/h]	147.9	147.9	147.9	147.9	591.6
Mass Flow [kg/h]	2372.57	2372.57	2372.57	2372.57	9490.28
Heat Flow [kJ/h]	-1.202×10^7	-1.202×10^7	-1.202×10^7	-1.202×10^7	-4.808×10^7
Component	Mole fraction	Mole fraction	Mole fraction	Mole fraction	Mole fraction
Methane	1.0000	1.0000	1.0000	1.0000	1.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000

- *Jetty BOG Compressor*

The motor compressor with a duty of 1120.52kW was used to raise the temperature and pressure of the jetty boil-off gas before the general BOG header.

The jetty BOG compressor raises the BOG pressure from **100 kPa** to **3500 kPa**, and the temperature is increased to **50.75°C** (Table 13). The compressor handles a mass flow of **9490.28 kg/h**, ensuring that the BOG is properly conditioned for reinjection or flaring.

- *BOG From Storage and LNG Knock-Out Drum Header*

The table below displays the result of the BOG generated from the four storage tanks and the knock-out drums. The outlet of the header was set to the lowest pressure of the inlet streams.

The combined BOG from the storage tanks and KO drums results in a total flow of **10,038.4 kg/h** at a pressure of **100 kPa** (Table 14). The BOG is collected for compression and reuse, demonstrating efficient management of vaporized gas.

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- *Storage and Knock-Out Drum BOG Compressor*

The electric motor compressor has a duty of 1253.52kW.

The BOG compressor for the storage and KO drums has a duty of **1253.52 kW** (Table 15). This compressor ensures the BOG is pressurized to **3500 kPa** for handling by the general BOG system. The consistent methane composition of **100%** shows that no other hydrocarbons are present in the BOG stream, making it ideal for reuse.

- *Equipment Rating*

Equipment rating was simulated on cooler, compressor, pump and pressure control valve. Result is displayed in table 16.

The equipment rating for pumps and compressors showed the compressors' and pumps' changes in temperature, pressure, and duty. The compressors raise the pressure by **3400 kPa**, while the jetty pumps increase pressure by **800 kPa**, all operating with similar power demands around **344 kW**.

The Pressure Control Valve (PCV) is also rated in Table 17, showing a pressure change of **1650 kPa** and temperature change of **0.6°C**, ensuring precise flow control across the system.

Table 13 Jetty BOG Compressor Result

Parameter	Compressor inlet (JBOG from the header)	To GENERAL BOG header
Vapour	1.0000	1.0000
Temperature [C]	-161.31	50.75
Pressure [kPa]	100.00	3500
Molar Flow [kgmole/h]	591.6	591.6
Mass Flow [kg/h]	9490.28	9490.28
Molar Entropy [kJ/kgmole-C]	150.26	155.90
Heat Flow [kJ/h]	-4.808×10^{7}	-4.404×10^{7}
Component	**Mole fraction**	**Mole fraction**
Methane	1.0000	1.0000
Total	1.0000	1.0000

Table 14 Storage and Knock-Out Drum Header Result

Name	Knock-out BOG	Tank 1 BOG	Tank 2 BOG	Tank 3 BOG	Tank 4 BOG	Outlet to comp.
Vapour	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
Temperature [C]	-161.31	-161.31	-161.31	-161.31	-161.31	-161.31
Pressure [kPa]	100	100	100	100	100	100
Molar Flow [kgmole/h]	404.25	55.36	55.36	55.36	55.36	625.7
Mass Flow [kg/h]	6485.47	888.23	888.23	888.23	888.23	10038.4
Heat Flow [kJ/h]	-3.286×10^{7}	-4.50×10^{6}	-4.50×10^{6}	-4.50×10^{6}	-4.50×10^{6}	-5.09×10^{7}

Table 15 Storage and Knock-Out Drum (SK) BOG Compressor Result

Parameter	Compressor inlet (SKBOG from the header)	To GENERAL BOG header
Vapour	1.0000	1.0000
Temperature [C]	-161.31	50.75
Pressure [kPa]	100	3500
Molar Flow [kgmole/h]	625.7	625.7
Mass Flow [kg/h]	10038.4	10038.4
Heat Flow [kJ/h]	-5.09×10^{7}	-4.659×10^{7}
Component	**Mole fraction**	**Mole fraction**
Methane	1.0000	1.0000
Total	1.0000	1.0000

Table 16 Equipment Rating of Pumps and Compressors

Parameters	Compressors	Pumps					
	JETTY	**STORAGE**	**LNG STORAGE**	**JETTY PUMP 1**	**JETTY PUMP 2**	**JETTY PUMP 3**	**JETTY PUMP 4**
Change in Temperature (°C)	212.05	212.05	0.1	0.6	0.6	0.6	0.6
Change in Pressure (kPa)	3400	3400	300	800	800	800	800
Duty (kW)	1121.03	1185.17	344.394	344.394	344.394	344.394	344.394

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Table 17 Equipment Rating of Pressure Control Valve

Parameters	PCV
Change in Temperature (°C)	0.6
Change in Pressure (kPa)	1650
Change in Mass Flow(kg/hr)	0
Valve opening (%)	50

Where PCV=Pressure Control Valve

- *Mass Balance*

The total mass balance of the concentrated area is shown in table 18 below. The total mass balance for the system shows no significant deviation, with inflow and outflow perfectly matching at **2002329.215 kg/h** (Table 18). This balance indicates that all streams are accounted for, with zero mass loss, ensuring efficient system operation.

- *Energy Balance*

The inlet and outlet heat flow of the concentrated system are summarized in the tables 19. The energy balance (Table 19) also demonstrates near-perfect alignment, with a small error of **-0.00013%**, confirming the accuracy of the heat flow calculations throughout the system.

- *Estimated Costing of BOG Recovery Equipment*

The cost of BOG recovery compressors is outlined in Table 20, with a total direct cost of **$1,943,900** and a total cost of investment (TCI) of **$3,790,605**. This estimation includes installation, instrumentation, piping, electrical, and working capital costs.

- *R. Annual Income from BOG*

As indicated in Table 21, the annual income from BOG recycling is estimated at **$138,121,200**, highlighting the economic viability of capturing and reusing BOG. The profit margins and cost savings make BOG recovery a lucrative aspect of LNG operations.

- *Profit Analysis on BOG Recycling*

The BOG profit analysis for recycling process is shown as follows;

- ✓ Gross Profit Margin = (Net Sales-Cost of materials)/(Net Sales)×100 = **97.3%**
- ✓ Break-Even = (Fixed Cost)/(Gross Profit Margin) = $ **3,895,791.4**

The profit analysis shows a Gross Profit Margin of 97.3%, with a break-even point of $3,895,791.4. This demonstrates that BOG recycling can be a highly profitable venture, with substantial returns on investment.

Table 18 Total Mass Balance

Stream	Mass Inflow (kg/hr)	Stream	Mass outflow (kg/hr)
LNG	2002329.215	Ship 1 BOG	2374.318092
		Ship 1 LNG	495697.5648
		Ship 2 BOG	2374.288332
		Ship 2 LNG	495697.5947
		Ship 3 BOG	2374.288347
		Ship 3 LNG	495697.5947
		Ship 4 BOG	2371.650514
		Ship 4 LNG	495700.2325
		KO BOG	6485.206154
		Storage tank 1 BOG	889.1192691
		Storage tank 2 BOG	889.1192694
		Storage tank 3 BOG	889.1192694
		Storage tank 4 BOG	889.1192694
Total	**2002329.215**		**2002329.215**
Difference = Mass inflow – Mass Outflow = 0.0000			
Hence % error = 0.0%			

KO: knock-out drum, BOG: Boil Off gas, LNG: Liquefied Natural Gas

Table 19 Total Energy Balance

Inflow Stream	Heat Inflow (kJ/hr)	Outlet Stream	Heat outflow (kJ/hr)
LNG	-11056585155	JLNG1	-2737712594.82511
TCOMP	4266627.479	JLNG	-2737698391.85143
JBCQ	4035723.883	JLNG 2	-2737698449.86276
LQ	1863088.575	JLNG 3	-2737698449.83266
SQ3	1239821.537	BOG	-90667232.16038

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TH	1239821.537		
SQ1	1239821.537		
SQ	1239821.537		
Total	**-11041460429.59**		**-11041475119.53**
	%Error	-0.00013%	

Table 20 Cost of BOG Recover Compressors

Name of Equipment	Type of Equipment E	Total Direct Cost (USD)	Equipment weight (lbs)	Installed weight (lbs)
Jetty Compressor	DGC IG CENTRIF	969,600.0	21000	44524
Storage and knock-out drum compressor	DGC IG CENTRIF	974,300.0	21000	44430
The total cost of Equipment (TCE)		1,943,900.0		
Installation	10% TCE	194,390.0		
Instrumentation	5% TCE	97,195.0		
Piping	15% TCE	291,585.0		
Electrical	6%TCE	116,634.0		
Working Capital	15%TCE	291,585.0		
Construction	10% TCE	194,390.0		
Contractor's fee	5% TCE	97,195.0		
Contingency	10% TCE	194,390.0		
Yard improvement	4% TCE	77,756.0		
Service facilities	15% TCE	291,585.0		
The total cost of investment (TCI)		3,790,605.0		

BOG-Boil-Off Gas, USD-United State Dollar, lbs- Pounds DGC IG- Integral compressor

- *Total Startup cost for BOG Recovery = $ 3,790,605.0*

Table 21 Annual Income from BOG Recycling

PRODUCT	Volume (kg/day)	Price ($/kg)	Production (days)	Total Amount ($)	Total Yearly income ($)
BOG from production	240,960	1.35	330	107,347,680.00	
BOG from shipping	227,952	1.35	100	30,773,520.00	**138,121,200.00**

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