Multidisciplinary Output Research For Actual and International Issue (MORFAI Journal) ISSN (e): 2808-6635
Volume 6 No. 4 (2026)

TECHNO-ECONOMIC ANALYSIS OF THE NATURAL GAS PIPELINE CONSTRUCTION AND REFINING PROCESS FROM GREATER SUNRISE OFFSHORE TO NATARBORA ONSHORE, TIMOR-LESTE

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Received	: 01 April 2026	Accepted	: 05 May 2026
Revised	: 10 April 2026	Published	: 16 May 2026

Abstract

The Greater Sunrise gas field, with a production capacity of 750 MMSCFD, is a strategic energy source for Timor-Leste. Its development is planned through an offshore refining facility and a 250 km pipeline to Natarbora, requiring comprehensive technical and economic studies. Technically, gas purification is carried out to remove CO_2, H_2S, heavy hydrocarbons, and water vapor to prevent corrosion, hydrate formation, and equipment damage during transportation. The processes used include a three-phase separator, acid gas separation (Acid Gas Removal Unit, AGRU), heavy hydrocarbon separation (Dew Point Control Unit, DPCU), and a water separation unit. Simulation results show that the gas is dominated by 85.1% methane, with the AGRU able to reduce CO_2 from 0.05 to 0.02, resulting in a marketable gas of 724.4 MMSCFD. At the DPCU with a flow rate of approximately 723 MMSCFD, it produces 11 barrels/day of condensate, and the dehydration process produces 723 MMSCFD of dry gas. Economically, a 24-inch diameter pipe was chosen as the most optimal option with a compressor requirement of 5 units. The total CAPEX of the project is 3,337 million USD with OPEX of 51 million USD, including the cost of the pipe, vessel installation, compressor, platform, and gas purification facilities. Based on the results of the gas price sensitivity calculation of 4-10 USD/MMBTU, the project is not feasible at a price of 4-5 USD/MMBTU, which is indicated by a negative NPV, IRR of 2-9%, and a Payback Period of 10-17 years. The project begins to be feasible at a price of 6 USD/MMBTU with an NPV of 1040 million USD, an IRR of 16%, and a Payback Period of 4 years. Furthermore, at a price of 7-10 USD/MMBTU, economic performance improves significantly, with an NPV of up to 6102 million USD, an IRR of up to 47%, and a shorter payback period of 1-3 years. At a gas price of around 7 USD/MMBTU, the project is declared economically feasible and prospective.

Keywords: Greater Sunrise, Compressor, Natural Gas Purification, Gas Pipeline.

INTRODUCTION

Natural gas is a primary energy source that plays a vital role in meeting Timor-Leste's energy needs and supporting national development (Wibowo, 2022). Currently, Timor-Leste still relies on imported fuel oil to meet its electricity needs. Following the end of Bayu-Undan Field production in 2023, state revenues from the oil and gas sector will decline, making the development of domestic energy sources crucial (Scheiner, 2021). The Greater Sunrise Field, located in the Timor Sea, between Timor-Leste and Australia, has gas reserves of approximately 7.6 Tcf and has the potential to become a long-term strategic energy source (Seggie et al., 2000). The field is approximately 250 km from the Natarbora mainland of Timor-Leste and approximately 450 km from Darwin, Australia, necessitating the construction of an undersea pipeline infrastructure (Seggie et al., 2003). The resulting gas, predominantly CH_4 methane, is planned for processing and development into Liquefied Natural Gas (LNG). Before being transported, gas must undergo a purification process to remove heavy hydrocarbons, CO_2, H_2S, and water vapor to prevent hydrate formation and corrosion (Kidnay et al., 2019). The selection of the diameter and route of the pipeline significantly impacts the technical efficiency and cost of the project. A comprehensive techno-economic analysis of the purification process and construction of the 250 km gas pipeline is required to ensure the feasibility, efficiency, and sustainability of the project in supporting Timor-Leste's economic growth (Syukur, 2023). The infrastructure to be built requires engineering and economic analysis to ensure the feasibility and sustainability of development in supporting Timor-Leste's national economic growth.

Several issues require attention in the process of refining and constructing the Greater Sunrise gas pipeline to the Timor-Leste mainland. To ensure the feasibility of developing the project, a comprehensive techno-economic analysis must be conducted. Several key issues require attention in the refining and construction of the natural gas pipeline. The main technical challenges include the need for an optimal gas purification process, the risk of hydrate formation, the potential for corrosion affecting the selection of pipe diameter and thickness, and various operational risks that can impact the stability and reliability of the submarine pipeline system (Pinandito, 2021). Economically, issues include the high investment costs (CAPEX), operating costs (OPEX), Net Present Value (NPV) calculations, the internal rate of return (IRR), and the payback period, which are key parameters in assessing the project's financial feasibility (Bahri et al., 2024; Michael, 2021). The Greater Sunrise project requires significant investment in logistics and infrastructure, including limited onshore facilities such as LNG reception at Natarbora, and the need for chemical and petroleum engineering experts who require intensive training for the operation and maintenance of the pipeline system. To determine whether the project is technically and economically feasible, a comprehensive techno-economic study is required as a basis for strategic decision-making to support the sustainable development of Timor-Leste's national energy system.

METHOD

A. Data collection

Tables 1 and 2 show the basic data used for the techno-economic analysis which includes gas characterization and the necessary economic aspects.

Table 1 Technical and economic variables

No	Variable	Data	Unit
1	Natural Gas Production Flow Rate	750	mmscfd
2	Length of Pipeline From GS to Natarbora	250	km
3	Gas Pressure (Operating Pressure On GS Platform)	180– 200	barg
4	Gas Temperature (Operating Temperature)	50	ºC
5	Project	25	Years
6	Discount Rate Assumptions	10-12	%

Table 2 Gas composition

Component	Composition (% mol)	Unit
Water	0.008	mol%
Carbon dioxide	3.46	mol%
Nitrogen	82.73	mol%
methane	4.89	mol%
Ethane	2.35	mol%
propane	0.62	mol%
i-Butane	0.92	mol%
n-Butane	0.04	mol%
Pentane Plus (C5+)	3.46	mol%

B. Process Simulation

Process simulation using UniSim for gas purification simulation (separator, AGRU, DPCU, and adsorption dehydration) and determination of pipe design parameters including outer diameter (OD) and wall thickness. Microsoft Excel is used to calculate inner diameter (ID) and pipe weight (W). Meanwhile, Microsoft Excel is used to analyze the financial feasibility of the project through the calculation of CAPEX, OPEX, NPV, IRR, and Payback Period, as well as presentation of the results in the form of NPV, IRR, and Payback Period graphs.

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COMPLETION OF NATURAL GAS PURIFICATION PROCESS

1. Separator

The initial separator functions to separate the production fluid into three phases, namely natural gas, hydrocarbons, and water. The produced gas is flowed to the Acid Gas Removal Unit with a flow rate of approximately 750 MMSCFD and a composition dominated by methane (CH_4) of 82.7 mol%. Under operating conditions, the fluid is in 100% gas phase, so the formation of liquid hydrocarbons is not significant due to the low content of heavy components C_2-C_5+. The separated free water content is also very small, which is approximately 0.008 mol%, which indicates the dominance of the gas phase in the system. The gas has a high calorific value and is suitable for LNG and gas transmission applications, but the CO_2 content of 5% still exceeds the sales specifications and requires further processing in the Acid Gas Removal Unit.

The presence of C_2-C_5+ components indicates that the gas has the potential to produce value-added condensate products. Although the water content is relatively low, a dehydration process is still required to prevent hydrate formation and corrosion in the piping system. The resulting free water flow (FW-2) is dominated by H_2O around 97%, with a CO_2 content of around 2.8% and other gases such as N_2 and CH_4, so it is categorized as Wastewater. The presence of CO_2 in the water phase has the potential to form corrosive carbonic acid (H_2CO_3), so further processing is required to reduce the CO_2 content and maintain the integrity of the process equipment.

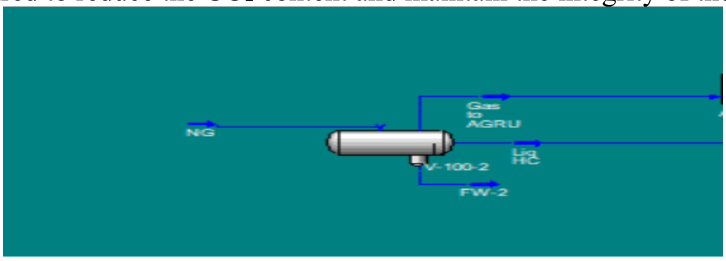

Figure 1 Unisim Separator

Table 3 Flow Conditions and Properties of Natural Gas to AGRU (Unisim Simulation)

Worksheet: Conditions | Properties | Composition | PF Specs

Name	NG	Gas to AGRU	Liq HC	FW-2
Vapour	1,0000	1,0000	0,0000	0,0000
Temperature [F]	122,0	122,0	122,0	122,0
Pressure [psia]	2750	2750	2750	2750
Molar Flow [MMSCFD]	750,0	750,0	0,0000	0,0000
Mass Flow [lb/day]	3,930e+007	3,930e+007	0,0000	0,0000
Std Ideal Liq Vol Flow [barrel/day]	3,156e+005	3,156e+005	0,0000	0,0000
Molar Enthalpy [Btu/lbmole]	-3,975e+004	-3,975e+004	-3,975e+004	-1,233e+005
Molar Entropy [Btu/lbmole-F]	33,37	33,37	33,37	14,82
Heat Flow [Btu/hr]	-3,274e+009	-3,274e+009	0,0000	0,0000

Table 4 Natural Gas Composition and Component Distribution to AGRU, Liquid Hydrocarbons, and Free Water (Unisim simulation)

Worksheet: Conditions | Properties | Composition | PF Specs

	NG	Gas to AGRU	Liq HC	FW-2
Nitrogen	0,0349	0,0349	0,0349	0,0012
CO2	0,0498	0,0498	0,0499	0,0284
Methane	0,8275	0,8275	0,8275	0,0001
Ethane	0,0485	0,0485	0,0485	0,0000
Propane	0,0234	0,0234	0,0234	0,0000
i-Butane	0,0062	0,0062	0,0062	0,0000
n-Butane	0,0092	0,0092	0,0092	0,0000
H2O	0,0001	0,0001	0,0001	0,9703
C5+*	0,0004	0,0004	0,0004	0,0000
EGlycol	0,0000	0,0000	0,0000	0,0000
TEGlycol	0,0000	0,0000	0,0000	0,0000

2. Acid Gas Removal Unit

Functions to remove acid gases, especially CO_2, from natural gas using an amine solution to produce sweet gas that is safe, non-corrosive, and meets pipeline gas specifications. Feed gas of 750 MMSCFD at 122 °F and 2750 psia is processed in the absorber column, so that the CO_2 content decreases from 0.05 to 0.02. The resulting gas (sweet gas) comes out at 724.4 MMSCFD and is flowed to the dehydration unit. The flow rate of the amine solution entering the absorber column (amine to contactor) is 450 MMSCFD. The amine solution that has absorbed the acid gas (rich amine) is then flowed to the flash drum to release the entrained light gas. Next, the rich amine is heated through a heat exchanger before entering the regenerator column (stripper) for the regeneration process. In the regenerator column, CO_2 is separated through a heating process using a reboiler. Acid gas of 114.4 MMSCFD is flowed to the flare system of the Sulfur Recovery Unit, while lean amine of approximately 359.4 MMSCFD is returned to the absorber column using an amine pump. This process requires approximately 196,500 hp of reboiler energy and approximately 5,506 hp of pump power.

Based on the simulation results, approximately 25.6 MMSCFD of acid gas was successfully absorbed by the amine solution. The decrease in the CO_2 mole fraction from 0.0499 to 0.0167, as well as the dominance of the methane (CH_4) fraction, indicates that the absorption process is effective and selective in improving gas quality.

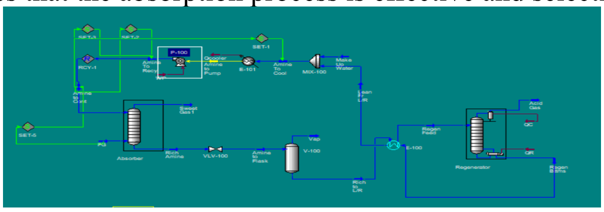

Figure 2 Unisim AGRU Absorber Process Flow

Table 5 Stream Simulation Results on Absorber Column (AGRU)

Worksheet: **Conditions** | Properties | Composition | PF Specs

Name	Amine to Cont	FG	Rich Amine	Sweet Gas1
Vapour	0,0000	1,0000	0,0000	1,0000
Temperature [F]	86,00	122,0	144,4	119,6
Pressure [psia]	2740	2750	2750	2740
Molar Flow [MMSCFD]	450,0	750,0	475,6	724,4
Mass Flow [lb/day]	3,207e+007	3,922e+007	3,502e+007	3,627e+007
Std Ideal Liq Vol Flow [barrel/day]	9,037e+004	3,154e+005	1,008e+005	3,049e+005
Molar Enthalpy [Btu/lbmole]	-1,049e+004	6327	-9644	6368
Molar Entropy [Btu/lbmole-F]	22,67	44,26	23,04	44,12
Heat Flow [Btu/hr]	-5,182e+008	5,210e+008	-5,037e+008	5,065e+008

Table 6 Results of Component Composition in the Acid Gas Absorption Process (AGRU)

Worksheet: Conditions | Properties | **Composition** | PF Specs

	Amine to Cont	FG	Rich Amine	Sweet Gas1
Nitrogen	0,0000	0,0349	0,0001	0,0361
CO2	0,0023	0,0499	0,0554	0,0167
Methane	0,0000	0,8278	0,0021	0,8558
Ethane	0,0000	0,0485	0,0001	0,0501
Propane	0,0000	0,0234	0,0000	0,0243
i-Butane	0,0000	0,0062	0,0000	0,0064
n-Butane	0,0000	0,0092	0,0000	0,0095
H2O	0,8964	0,0001	0,8465	0,0011
MEAmine	0,0221	0,0000	0,0209	0,0000
MDEAmine	0,0793	0,0000	0,0750	0,0000

3. Dew Point Control Unit

The DPCU process aims to lower the hydrocarbon dew point so that the gas remains in the gas phase during transportation, thereby preventing condensate formation and meeting sales specifications (Campbell, 1992-Volume 2). The feed gas of 724.8 MMSCFD still contains heavy hydrocarbons and produces condensate of approximately 11 barrels/day. To prevent hydrate formation, ethylene glycol (EG) of 114.9 barrels/day (0.398 MMSCFD equivalent) is injected as an inhibitor.

Initial cooling of E-101 reduces the temperature from 113.3 °F to 86.0 °F without significant phase changes. Subsequently, cooling of E-102 to 32.0 °F causes condensation of heavy hydrocarbons. In the Cold Separator, the fluids are separated into: Gas of 723.4 MMSCFD, condensate of 20 barrels/day, EG+water mixture of 236.2 barrels/day which is recirculated.

The condensate is then stabilized in COL-6, producing 0.0178 MMSCFD of light gas and 11 barrels/day of condensate. The condensate is then stored in Tank V-103 under stable conditions.

Based on the composition analysis, the performance of the DPCU unit shows effective results, namely: the water content decreased from 0.0016 to 0.0001, which indicates that the dehydration process is optimally carried out by EG. Heavy hydrocarbons (C_2-5+) were successfully separated by 0.0280 MMSCFD through the condensation process. The methane fraction (CH_4) increased from 0.8548 to 0.8563, which indicates that the loss of light hydrocarbons is very small. Overall, the DPCU unit works effectively and selectively in reducing the water dew point and hydrocarbon dew point, so that the gas meets the specifications for transportation and sales.

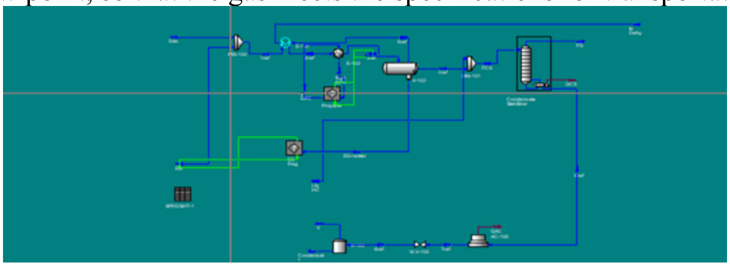

Figure 3 Unisim DPCU Flow Process

Table 7 Operating Conditions of Gas Flow in Heat Exchanger

Name	1ref	2ref	5ref	to ads
Vapour	0,9992	0,9984	1,0000	1,0000
Temperature [F]	113,3	57,31	32,00	86,00
Pressure [psia]	1160	1150	1140	1130
Molar Flow [MMSCFD]	724,8	724,8	723,4	723,4
Mass Flow [lb/day]	3,639e+007	3,639e+007	3,630e+007	3,630e+007
Std Ideal Liq Vol Flow [barrel/day]	3,052e+005	3,052e+005	3,049e+005	3,049e+005
Molar Enthalpy [Btu/lbmole]	-3,474e+004	-3,544e+004	-3,559e+004	-3,489e+004
Molar Entropy [Btu/lbmole-F]	35,68	34,41	33,82	35,19
Heat Flow [Btu/hr]	-2,765e+009	-2,821e+009	-2,827e+009	-2,772e+009

Table 8 Mole Fraction Composition of Gas in Heat Exchanger

E-101

	1ref	2ref	5ref	to ads
Nitrogen	0,0361	0,0361	0,0361	0,0361
CO2	0,0166	0,0166	0,0167	0,0167
Methane	0,8548	0,8548	0,8563	0,8563
Ethane	0,0501	0,0501	0,0502	0,0502
Propane	0,0242	0,0242	0,0243	0,0243
i-Butane	0,0064	0,0064	0,0064	0,0064
n-Butane	0,0095	0,0095	0,0095	0,0095
H2O	0,0016	0,0016	0,0001	0,0001
C5+*	0,0004	0,0004	0,0004	0,0004
EGlycol	0,0003	0,0003	0,0000	0,0000
TEGlycol	0,0000	0,0000	0,0000	0,0000

4. Adsorption Dehydration Unit

The gas from the DPCU then enters a molecular sieve-based adsorber dehydration system with a flow rate of 723.4 MMSCFD, at operating conditions of 1130 psia and 85 °F. This dehydration process aims to remove water vapor content from the gas stream through an adsorption mechanism using molecular sieve media.

During the process, the adsorbed water is separated and exits through the bottom flow of 0.077 MMSCFD, while dry gas is produced as the main product and exits through the top to the sales gas line with a flow rate of approximately 723.4 MMSCFD. Based on the simulation results, the gas at 86 °F and 1130 psia is completely in the vapor phase with a flow rate of 723.4 MMSCFD. This shows that the dehydration process is effective in reducing water content, so that the gas meets the conditions of sales gas and is safe for transmission without the risk of hydrate formation or corrosion.

The gas composition is dominated by methane (CH_4) at 85.6 mol%, with small contents of ethane, propane, and other heavy hydrocarbons. The content of CO_2 (1.67%) and nitrogen (3.61%) is also relatively low. After the dehydration process, the water content (H_2O) decreases from 0.0001 to 0, while other components do not experience significant changes. Overall, the molecular sieve-based dehydration unit works effectively and selectively in removing water content without affecting the main composition of the gas, so that the gas quality meets the specifications for the transmission and sales process.

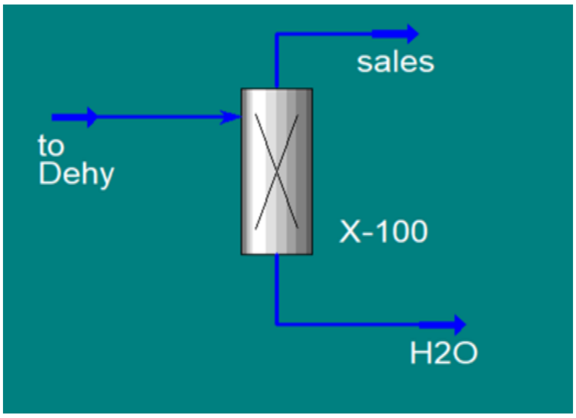

Figure 4 Dehydration Unit Adsorption

Table 9 Results of Stream Conditions and Properties in the Dehydration Unit Adsorption

X-100

Worksheet: Conditions | Properties | Composition | PF Specs

Name	to Dehy	sales	H20+AC
Vapour	1,0000	1,0000	0,7086
Temperature [F]	86,00	86,00	558,8
Pressure [psia]	1130	1130	1130
Molar Flow [MMSCFD]	723,4	723,4	7,748e-002
Mass Flow [lb/day]	3,630e+007	3,629e+007	3692
Std Ideal Liq Vol Flow [barrel/day]	3,049e+005	3,049e+005	10,55
Molar Enthalpy [Btu/lbmole]	-3,489e+004	-3,489e+004	-1,050e+005
Molar Entropy [Btu/lbmole-F]	35,19	35,18	33,68
Heat Flow [Btu/hr]	-2,772e+009	-2,771e+009	-8,935e+005

Table 10 Stream Composition in the Adsorption Dehydration Unit

Worksheet: Conditions | Properties | Composition | PF Specs

	to Dehy	sales
Nitrogen	0,0361	0,0361
CO2	0,0167	0,0167
Methane	0,8563	0,8564
Ethane	0,0502	0,0502
Propane	0,0243	0,0243
i-Butane	0,0064	0,0064
n-Butane	0,0095	0,0095
H20	0,0001	0,0000
C5+*	0,0004	0,0004
EGlycol	0,0000	0,0000
TEGlycol	0,0000	0,0000

5. Compressor

Compressor power requirements are determined by the compression requirements to maintain the gas flow rate and pressure in the pipeline, and serve as the basis for estimating investment and operating costs (Michael, 2021). UniSim simulation results show changes in gas flow conditions before and after the compressor. Pressure increases from 1130 psia to 2200 psia and temperature increase, while the flow rate remains constant. Furthermore, the gas composition (CH_4, CO_2, and other components) remains unchanged between the inlet and outlet.

Based on the simulation results using UniSim, a 20-inch diameter pipe requires 8 compressor units, with a power of 27,328 hp on the first unit and 50,328 hp on the next unit (Design 2–8), so that the total power requirement reaches 379,624 hp. For a 22-inch pipe, 6 compressor units are needed, with a power of 27,328 hp on the first unit and 25,541 hp on the next unit (Design 2–6), so that the total power is 155,035 hp. A 24-inch pipe only requires 5 compressor units, with a power of 27,328 hp on the first unit and 14,944 hp on the next unit (Design 2–5), so that the total power reaches 87,104 hp.

The power value indicates the compressor's energy requirements to achieve the pressure and gas flow rate according to system specifications and maintain pressure stability during the gas transportation process so that it is safe and sustainable.

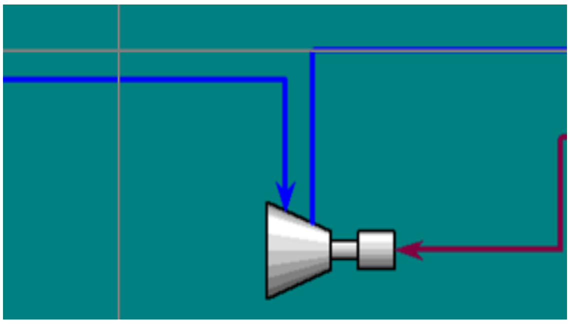

Figure 5 Unisim Compressor Design

Table 11 flow conditions in the compressor

Name	sales	Pipa
Vapour	1,0000	1,0000
Temperature [F]	86,00	194,3
Pressure [psia]	1130	2200
Molar Flow [MMSCFD]	723,4	723,4
Mass Flow [lb/day]	3,629e+007	3,629e+007
LiqVol Flow [barrel/day]	3,049e+005	3,049e+005
Molar Enthalpy [Btu/lbmole]	-3,489e+004	-3,401e+004
Molar Entropy [Btu/lbmole-F]	35,18	35,52
Heat Flow [Btu/hr]	-2,771e+009	-2,701e+009

Table 12 Flow Composition in Compressor

	sales	Pipa
Nitrogen	0,0361	0,0361
CO2	0,0167	0,0167
Methane	0,8564	0,8564
Ethane	0,0502	0,0502
Propane	0,0243	0,0243
i-Butane	0,0064	0,0064
n-Butane	0,0095	0,0095
H2O	0,0000	0,0000
C5+*	0,0004	0,0004
EGlycol	0,0000	0,0000
TEGlycol	0,0000	0,0000

6. Piping System

Piping systems function as a gas transport medium, designed based on pipe diameter, thickness, and material. Pipe sizes must be adjusted to the gas pressure and flow rate to optimize performance and minimize the risk of erosion (Ahamad et al., 2022).

Pipe diameter directly affects flow capacity, pressure stability, and investment costs (Association, 1998; Michael, 2021). Diameter determination was carried out through UniSim simulation, resulting in alternative sizes of 20 inches (508 mm), 22 inches (558.8 mm), 24 inches (609.6 mm), and 26 inches (660.4 mm). Pipe wall thickness was determined based on the ASME/ANSI B36.19 standard, taking into account safety aspects, material strength, and availability in the industry, so that pipe specifications meet applicable standards.

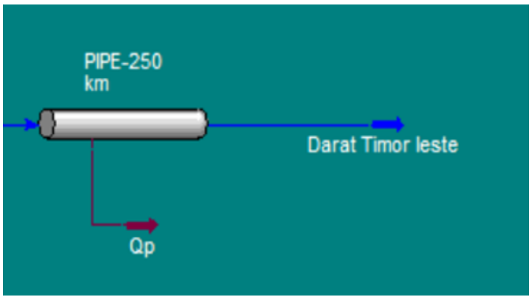

Figure 6 Pipe Diameter Size Design

Table 13 Flow Conditions in Gas Transmission Pipelines

Worksheet: **Conditions** | Properties | Composition | PF Specs

Name	Pipa	P70 KM1
Vapour	1,0000	1,0000
Temperature [F]	194,2900	71,8950
Pressure [psia]	2200	1452
Molar Flow [MMSCFD]	723,3645	723,3645
Mass Flow [lb/day]	36291599,1642	36291599,1642
LiqVol Flow [barrel/day]	304912,4760	304912,4760
Molar Enthalpy [Btu/lbmole]	-3,401e+004	-3,527e+004
Molar Entropy [Btu/lbmole-F]	35,52	34,07
Heat Flow [Btu/hr]	-2,70121e+09	-2,80167e+09

Table 14 Flow Composition in Gas Transmission Pipelines

Worksheet: Conditions | Properties | **Composition** | PF Specs

	Pipa	P70 KM1
Nitrogen	0,0361	0,0361
CO2	0,0167	0,0167
Methane	0,8564	0,8564
Ethane	0,0502	0,0502
Propane	0,0243	0,0243
i-Butane	0,0064	0,0064
n-Butane	0,0095	0,0095
H2O	0,0000	0,0000
C5+*	0,0004	0,0004
EGlycol	0,0000	0,0000
TEGlycol	0,0000	0,0000

Determination of pipe wall thickness for diameters of 20, 22, 24, and 26 inches was carried out based on UniSim simulation results to ensure the pipe is able to withstand operating pressure and prevent structural failure. Calculations refer to standard pipe design equations, thus meeting mechanical strength and operational safety requirements (Chipade & Patil, 2016; Pandapipe, 2026).

The pipe wall thickness is calculated using the equation,

$$t=\frac{P\ do}{2(S\ E\ W+P\ Y)} \qquad 1$$

tm = t+c

The inner diameter of the pipe is calculated using the equation

Di=D0-2t 2

Meanwhile, the weight of the pipe per unit length is calculated using the equation:

W(kg/m)=0.02466x(OD-t)xt 3

In calculating pipe wall thickness, the calculated thickness value is compared with the standard thickness based on ASME B36.10/B36.19 (Corp, 1987). Next, the closest and larger standard thickness is selected to meet the design safety factor.

In determining the pipe inner diameter (ID) and pipe weight (W), standard pipe thicknesses are used which have been selected based on ASME B36.10/B36.19 (Corp, 1987). The specifications aim to meet design safety requirements and consider the availability of pipe sizes commonly used in the industry.

Table 15 Calculation Results for Wall Thickness, Inner Diameter, Pipe Weight

Pipe Diameter (inch)	OD (mm)	t Count (mm)	t Standard (mm)	ID (mm)	ID (inch)	Weight (kg/m)
20	508	34.25	38.1	431.8	17	441
22	558.8	37.38	41.28	476.24	19	527
24	609.6	40.5	46.02	517.56	20.38	640

Based on the table, increasing the pipe diameter results in an increase in the inner diameter (ID) and pipe weight. 20-, 22-, and 24-inch pipes have IDs of 431.8 mm, 476.24 mm, and 517.56 mm, respectively, with weights of 441, 527, and 640 kg/m, respectively. Larger diameters provide higher flow capacities and reduce pressure losses, although they increase material requirements.

The 26 inch diameter was not used because it does not meet the ASME B36.10/B36.19 standard, so the analysis focused on the 20, 22, and 24 inch diameters.

7. Pipe CAPEX

One of the main components of the CAPEX of a gas transmission project is the pipeline investment cost, which is calculated using the equation (Michael, 2021).

Based on international pipe price data, API 5L Grade X65 Schedule 120 pipe with a diameter of 20 inches and a thickness of 38.1 mm has a price of around 3591.9 USD per ton. Pipe with a diameter of 22 inches and a thickness of 41.28 mm has a price of around 4563 USD per ton, while pipe with a diameter of 24 inches and a thickness of 46.02 mm has a price of around 3591.9 USD per ton (Steel, 2008).

CAPEX 24 inch Pipe= *ton/km X USD/ton*
4
Pipe CAPEX = Cost Per km X Pipe Length 5

Table 16 Results of Pipe CAPEX Calculation Based on Diameter

Pipe Diameter (inch)	CAPEX (Million USD/km)	long (km)	Total CAPEX (Million USD)
20	1.59	250	397
22	2.40	250	601
24	2.30	250	574

8. CAPEX for chartering submarine pipeline installation vessels.

The charter cost for the S-Lay pipe installation vessel is USD 350,000/day with a productivity of 5 km/day (Re, 2020). The length of the Greater Sunrise pipeline from the offshore area to Timor Leste is estimated at 250 km (Seggie et al., 2003).

The installation duration of the Greater Sunrise pipeline from sea to land in Timor Leste is 250 km and 50 days at a cost of around 17.5 million USD.

9. Compressor CAPEX

Compressor CAPEX is calculated based on power (horsepower) with the equation:
Compressor Price = 20000 X Per hp (Copco, 2025), assuming 20,000 USD/hp and the CAPEX cost of offshore platform installation is assumed to be 200 million USD per gas compressor unit.
Compressor Price = 20000 X Per hp 6
CAPEXplatform = Total Compressor x 200 million USD

Table 17 Results of Total CAPEX Calculation for Offshore Compressors and Platforms

Pipe Diameter (inch)	Number of Compressors (units)	Total Compressor Power (HP)	Compressor Price (Million USD)	Platform CAPEX (Million USD)
20	8	379,624	7,592	1,600
22	6	155,035	3.101	1,200
24	5	87,104	1,742	1,000

10. Total CAPEX

Total capex is the accumulated cost of pipelines, installation vessel rentals, compressors, and offshore platforms.

CCAPEX =Cpipe+ Ccomp+Cvessel+ Cplatform
8

Table 18 Results of Total Project CAPEX Calculation Based on Pipe Diameter

Pipe Diameter (inch)	Pipeline CAPEX (Million USD)	Compressor CAPEX (Million USD)	Ship CAPEX (Million USD)	Platform CAPEX (Million USD)	Total CAPEX (Million USD)
20	397	7,592	17.5	1,600	9,606
22	601	3.101	17.5	1,200	4,919
24	574	1,741	17.5	1,000	3,334

The results in Table 18 show that the 24-inch pipe has the lowest CAPEX, so it was selected as the most economical option.

11. CAPEX separator

The estimated cost of the separator is Rp. 3,449,436,000.

Assuming an exchange rate of 1 USD = Rp. 15,000, it becomes 229,962 USD (0.23 million USD) (Alibaba, 2026b).

12. CAPEX AGRU

Based on data from the CO_2 content separation industry from natural gas in Indonesia (Adhim & Nirmala, 2025), the estimated cost of AGRU is IDR 17,077,737,600.

Assuming an exchange rate of 1 USD = Rp. 15,000, it becomes 1,138,516 USD (1.14 million USD) (Alibaba, 2026d).

13. CAPEX DPCU

Estimated DPCU costs are Rp. 8,626,120,000.

Assume that the exchange rate is 1 USD = Rp. 15,000, which is 0.58 million USD (Alibaba, 2026a).

14. CAPEX of adsorption dehydration

Based on industrial data for separating water content in natural gas using the molecular sieve dehydration process (Batuthoh et al., 2024), the estimated unit cost of molecular sieve dehydration is IDR 12,075,168,000.

Assuming an exchange rate of 1 USD = Rp. 15,000, it becomes 805,011 USD (0.81 million USD) (Alibaba, 2026c).

Table 19 Results of Gas Purification Unit CAPEX Estimation

Purification Process Unit	CAPEX (Million USD)
Separator	0.230
AGRU	1.14
DPCU	0.575
Adsorption Dehydration	0.805

Total	3

The total system CAPEX is the sum of the transportation CAPEX and the refining CAPEX of the Eq. project.
Total CAPEX=Transportation CAPEX + Refinery CAPEX 9
Total CAPEX = 3334 + 3 = 3337 Million USD

15. OPEX

Operational costs (OPEX) include costs during the project's operating period (Ardiansyah, 2008).
Annual OPEX factors assumed: pipe 2%, separator 4%, AGRU 6%, DPCU 5%, and dehydration unit 12%
Pipe OPEX= 2% × 574 million = 11.5 million USD/year
Cost/day = Qmmbtu/dayxgas price.

Table 1 Gas Composition and Gross Heating Value (GHV) Calculation

Comp	mole fraction	GHV	GHVi
N2	0.0361	0	0
CO2	0.0167	0	0
C1	0.8564	1010	864,984
C2	0.0502	1769	88.7755
C3	0.0243	2518	61,1155
iC4	0.0064	3256	20,9931
nC4	0.0095	3264	31,0194
H2O	0	0	0
C5+	0.0004	0	1,5569
Total GHVI (BTU/SCF			1068

Based on UniSim simulations, the compressor's total heat flow of 221.63 million BTU/hr with an efficiency of 0.35 results in a fuel requirement of approximately 15,197 MMBTU/day. With a gas price of $7/MMBTU, the fuel cost is approximately $106,382/day, or approximately $39 million/year.
Refining OPEX Calculation
OPEX = OPEX factor (% per year) X CAPEX

Table 21 OPEX Calculation for Gas Purification Unit

Process Unit	Percentage OPEX (% CAPEX)	OPEX (Million USD/year)
Separator	4%	0.009
AGRU	6%	0.068
DPCU	5%	0.029
Adsorption Dehydration	12%	0.097
Total Purification		0.20

Total OPEX=pipe OPEX+compressor OPEX+refining OPEX 11
OPEXtotal=11.5+39+0.20= 51 Million USD

16. Economic Parameters

Net Present Value (NPV) is an indicator of project feasibility based on the present value of net cash flows. A positive NPV indicates a profitable project, while a negative NPV indicates it is not feasible.(Ardiansyah, 2008; Wang, 2025).

$$NPV= []-\sum_{t=1}^{n} \frac{CF_t}{(1+r)^t} I_0 \quad 12$$

The Internal Rate of Return calculation is carried out using IRR in Microsoft Excel, namely = IRR (range-cashflow), which is based on the project's cash flow.(Patrick & French, 2016).

Industrial standard natural gas project Gas 8-12% assume r1 10%.
The flow rate of the natural gas purification process is 723 mmscfd, thus providingwith 264028 mmbtu/year.
Projected Production DeclineExponential Decline Methodand Analysis Revenue (R_t), Cash Flow, IRR, and Payback Period, with gas price (4–10 USD/MMBTU) (Ardiansyah, 2008; Rahaman et al., 2025), then the income (*revenue):*
Revenue(Rt) = Qt × P 14
Net Cash Flow (CFt)= Rt – OPEXt −A-depreciation -Tax 15
Payback Period (PP) is calculated by identifying the time when the initial investment (-3337 million USD) is covered by the accumulated cash flow, namely when the value changes from negative to positive as an indicator of achieving a return on investment.Based on a project life of 25 years and an investment return period (installments) of 10 years, using the discrete compounding method at an interest rate of 6%, the Capital Recovery Factor (CRF) value is 0.1359. Annual cost (A) can be calculated using the equation(Bontadelli, 1999.).
Annual cost (A)=0.1359XCAPEX=453.45 Million USD

Depreciation Calculation
the decline in value of an asset during its useful life calculated annually for economic analysis purposes(Nikijuluw et al., 2023).

D=CS/n
D=3337 /25=133.5 Million USD
Tax Calculation = - 0.025X (Revenue - Depreciation)
Exposure calculation results for 25 years see the table below.

The analysis of production decline and economic evaluation of the project in the gas price scenario of 4–10 USD/MMBTU is shown in Table 22.

Table 22 Gas Price Sensitivity to NPV, IRR, and Payback Period"

Gas Prices	NPV10	IRR	PP	Depreciation	Panjak
4	-1491	2.00%	17	133.5	-164
5	-225	9.00%	10	133.5	-225
6	1040	16.00%	4	133.5	-286
7	2306	24.00%	3	133.5	-347
8	3571	32.00%	2	133.5	-408
9	4837	40.00%	2	133.5	-469
10	6102	47.00%	1	133.5	-530

Based on the calculation results in the gas price sensitivity table, it is clear that increasing gas prices significantly impact the project's economic feasibility. At a price of 4–5 USD/MMBTU, the project is not yet financially viable, as indicated by a negative NPV, an IRR below the discount rate, and a relatively long Payback Period (9–17 years). Starting at a price of 6 USD/MMBTU, the project demonstrates feasibility with a positive NPV of 1,040 million USD and an IRR of 16%. At a price of 7–10 USD/MMBTU, economic performance improves significantly. At a price of 7 USD/MMBTU, the NPV reaches 2,306 million USD, an IRR of 24%, and a Payback Period of approximately 3 years. At a price of 10 USD/MMBTU, the NPV increases to 6,102 million USD, an IRR of 47%, and a Payback Period of approximately 1–3 years.

With current gas prices around 7 USD/MMBTU, the project is deemed feasible and offers attractive economic returns.

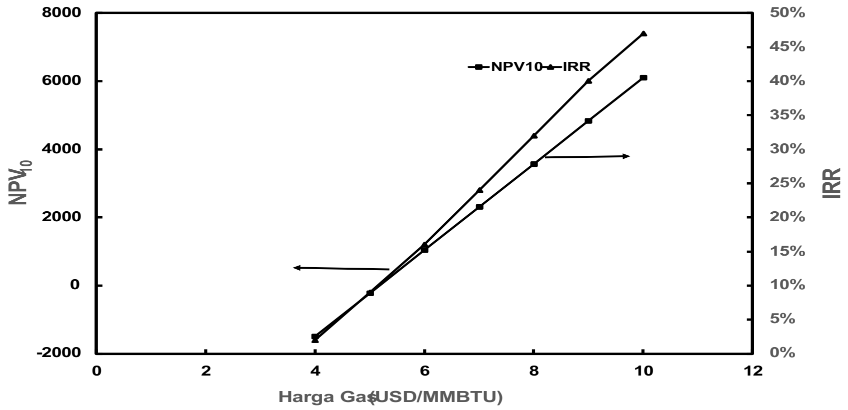

Figure 7. Gas price sensitivity (4–10 USD/MMBTU). against NPV (discount rate 10%) and IRR

Based on the analysis of the relationship between gas prices and economic parameters (NPV and IRR), it is also clear that at a price range of 4–5 USD/MMBTU, the project is still not economically feasible. This is indicated by the NPV value which is still negative and the IRR which is below the discount rate (10%). At a price of 4 USD/MMBTU, the NPV is -1491 million USD with an IRR of 2%, while at a price of 5 USD/MMBTU the NPV increases to -225 million USD with an IRR of 9%, but still does not meet the feasibility criteria. As gas prices increase, the NPV and IRR show a significant upward trend.
At a gas price of around 7 USD/MMBTU, the project is declared economically viable because the NPV has been positive and the IRR has exceeded the discount rate, thus providing an attractive return.

RESULTS AND DISCUSSION

The simulation results show that the production fluid is dominated by gas of 750 MMSCFD with a methane content of 82.7 mol% and very low water content of 0.008 mol%, so that no hydrocarbon condensate is formed. In the AGRU, CO_2 is reduced from 0.05 to 0.02 mol fraction, resulting in 724.4 MMSCFD of sweet gas. Furthermore, in the DPCU, a small condensate (11 barrels/day) and 723.4 MMSCFD of gas are formed, which are then dehydrated using a molecular sieve to form 723 MMSCFD of dry gas. 0.077 MMSCFD of adsorbed water is formed.
For a 250 km transmission, the compressor requirement decreases with increasing pipe diameter: 20 inch (8 units; 379,624 hp), 22 inch (6 units; 155,035 hp), and 24 inch (5 units; 87,104 hp). The CAPEX analysis shows that the 24 inch pipe is the most optimal with the lowest cost of 3,334 million USD. The total CAPEX of the project is estimated at 3,337 million USD, with operational costs (OPEX) of 51 million USD. The investment includes the construction of a natural gas purification facility and the installation of a 24 inch diameter transmission pipe with a length of approximately 250 km. This value includes the CAPEX of the pipe of 574 million USD, the rental cost of the subsea pipe installation vessel of 17.5 million USD, the cost of the compressor of 1,742 million USD, and the installation cost of the offshore compressor platform of 1,000 million USD for 5 compressor units with an estimated 200 million USD per unit. In addition, there are gas purification facility costs which include a separator of USD 0.230 million, an Acid Gas Removal Unit of USD 1.14 million, a Dew Point Control Unit of USD 0.58 million, and a dehydration unit of USD 0.805 million.

Based on a gas price sensitivity analysis, project feasibility is significantly influenced by the natural gas selling price. At a price of 4–5 USD/MMBTU, the project is not economically feasible because the NPV is still negative, amounting to -1,491 million USD (at 4 USD/MMBTU) and -225 million USD at 5 USD/MMBTU. The IRRs are only 2% and 9%, respectively, still below the discount rate (10%), with payback periods of approximately 17 and 10 years, respectively, making the investment payback period relatively long.

The project began to demonstrate economic feasibility at a price of 6 USD/MMBTU with an NPV of 1040 million USD, an IRR of 16%, and a Payback Period of approximately 4 years, which has met the investment criteria. At a price of 7-10 USD/MMBTU, the project's economic performance improves significantly. At a price of 7 USD/MMBTU, the NPV is obtained at 2306 million USD, an IRR of 24%, and a Payback Period of approximately 3 years. At a higher price, the NPV increases to 6102 million USD, the IRR reaches 47%, and the Payback Period is faster, namely around 1-3 years. Considering that the current natural gas price is around 7 USD/MMBTU, this project can be declared economically feasible and has excellent development prospects, with the potential for an attractive rate of return under current market conditions.

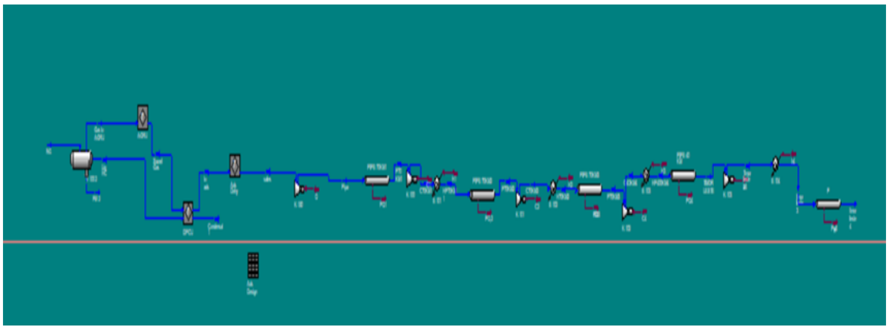

Figure 8 Design Flowchart for Technical Analysis of Natural Gas Purification and Pipeline Construction Process

CONCLUSION

Techno-economic analysis of the construction of a gas purification system and pipeline from the Sea to Natarbora shows high feasibility, with an integrated study covering process design, compressor system, pipe diameter optimization, as well as CAPEX, NPV, IRR, and Payback Period evaluations. Technically, the purification process series consisting of a three-phase separator, Acid Gas Removal Unit Dew Point Control Unit and molecular sieve adsorption dehydration unit is capable of producing gas according to specifications. Feed gas of 750 MMSCFD with a methane content of 82.7 mol% was successfully purified with a reduction in CO_2 from 0.05 to 0.02 mole fraction, and produced condensate of 11 barrels/day. The outlet gas flow rate was 724.4 MMSCFD and after dehydration became 723.4 MMSCFD, thus meeting industry standards and preventing hydrate formation. The results of the transportation analysis showed that increasing the pipe diameter reduces the need for a compressor. The 24-inch pipe was chosen as the optimum design because it is more energy efficient, has a lower pressure drop, and more economical investment costs. Economically, the project's total CAPEX is USD 3,337 million with OPEX of approximately USD 51 million per year. At gas prices of USD 4-6, the project is still not feasible. At a gas price of USD 7/MMBTU, the project is feasible with an NPV of USD 2,306 million, an IRR of 24%, and a Payback Period of 3 years. At prices up to USD 10/MMBTU, the NPV reaches USD 6,102 million, an IRR of 47%, and a Payback Period of 1-3 years. With the current global gas price of approximately USD 7/MMBTU, the project is declared feasible and profitable. The integration of gas refining and pipeline systems has proven to be technically and economically feasible, and has the potential to increase the added value of Greater Sunrise gas and support Timor-Leste's energy security.

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