Chair of Petroleum and Geothermal Energy Recovery

Master's Thesis

Challenges in production and distribution of low-tonnage LNG

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Kurzfassung

Laut der globalen Marktanalyse besteht ein dringender Bedarf, auf grüne Energie umzusteigen, insbesondere im Kraftstoff- und Energiesektor (Versorgungsunternehmen, Produktionsbedarf, Spezialgeräte, LKWs usw.). Daher hat LNG in den letzten Jahren bei Unternehmen an Popularität gewonnen. Die LNG-Produktion im kleinen Maßstab kann die meisten dieser Anforderungen erfüllen, beispielsweise, sie erfordert nicht viel Zeit für den Bau der Anlage, hohe Investitionskosten, viel Platz für den Bau, usw.). Das Grundschema der LNG-Produktion basiert auf dem Drosseleffekt im Reduktionsaggregat, der mit Hilfe eines Drosselventils realisiert wird, aber es gibt auch andere Möglichkeiten, diesen Effekt zu realisieren.

Das Ziel dieser Forschungsarbeit ist die Optimierung und Verbesserung einer Flüssiggasanlage mit geringer Tonnage, die auf dem Stickstoffkreislauf basiert. Dies geschieht durch die Auswahl der optimalen Leistung der Betriebsmodi der LNG-Anlage sowie durch die Analyse der Effizienz der Reduktionseinheiten und den Vorschlag möglicher Optionen zur Verbesserung des Kühlprozesses in dieser Anlage, basierend auf vielversprechenden Technologien (Turboexpander und Überschallseparator).

Der globale Markt für die Produktion von Flüssigerdgas mit niedriger Tonnage hat gezeigt, dass die wichtigsten Faktoren für die Wahl der LNG-Technologie die technologische Effizienz, die Zuverlässigkeit der Konstruktionslösungen, die einfache Wartung der Anlage, die Modularität und die niedrigen Investitionskosten sind. Darüber hinaus sollte die Notwendigkeit, die Betriebskosten zu minimieren, berücksichtigt werden. Als Ergebnis dieser Analyse wurde ein Schema, das auf einem externen isolierten Stickstoffkreislauf basiert, als Grundlage für die weitere Untersuchung gewählt.

Basierend auf der Analyse der Leistungsindikatoren der Betriebsarten der LNG-Anlage wurde festgestellt, dass bei Durchflussraten über 5.500 Nm3/h das Gaskühlsystem der Belastung nicht gewachsen ist, d.h. es arbeitet nicht stabil und mit einer großen Fehlerspanne bei den erforderlichen Werten. Die Arbeit bewertete die technologische Effizienz einer Reihe von möglichen Modi der Steuerung der LNG-Anlage in Abhängigkeit von der Anforderung der potenziellen Kunden. Es wurde festgestellt, dass die LNG-Anlage, die aus 6 Linien von 5500 Nm3/h besteht, die größte Flexibilität hat, die sowohl die minimal mögliche Belastung der Anlage, als auch die maximale, die die geforderte um 30% überschreitet, gewährleisten kann.

Es wurde auch festgestellt, dass der Turboexpander und der Überschallabscheider ungefähr das gleiche Potential haben. Diese Werte liegen jedoch innerhalb der Fehlermarge. In Anbetracht der für den 3S-Abscheider erforderlichen Materialien werden spezielle nicht spröde und kältebeständige Werkstoffe benötigt. Es wurde auch festgestellt, dass Turboexpander und Überschallabscheider in etwa das gleiche Potenzial haben. Diese Werte liegen jedoch innerhalb der Fehlermarge. In Anbetracht der für den 3S-Abscheider erforderlichen Materialien werden spezielle nicht spröde und kältebeständige Werkstoffe benötigt.

Abstract

According to the global market analysis, there is an urgent need to switch to green energy, especially in the fuel and energy sector (utilities, production needs, special equipment, trucks, etc.). Therefore, LNG has been gaining popularity among companies in recent years. Small-scale LNG production can cope with most of these needs for example, it does not require much time for construction of the plant, high capital expenditures, large amount of space for construction, etc.). The basic scheme of LNG production is based on the throttle effect in the reduction unit realized by means of a throttle valve, but there are other ways of realizing this effect.

The objective of this research work is to optimize and improve a low-tonnage liquefied natural gas plant based on the nitrogen cycle. This is done by selecting the optimum performance of the LNG plant operating modes, as well as analysing the efficiency of the reduction units and proposing possible options for improving the cooling process in it based on promising technologies (turbo expander and supersonic separator).

The global market for low-tonnage liquefied natural gas production has revealed that the main factors for the choice of LNG technology are technological efficiency, reliability of design solutions, ease of plant maintenance, modularity and low investment costs. Moreover, the need to minimise operating costs should be considered. As a result of this analysis, a scheme based on an external isolated nitrogen cycle was chosen as the basic one for further study.

Based on the analysis of performance indicators of LNG plant operating modes, based on the results of the calculation experiment, it was found that at flow rates above 5,500 Nm3/h, the gas cooling system cannot cope with the load, i.e., it does not work stably and with a large margin of error in the required values. The work assessed the technological efficiency of a number of possible modes of LNG plant control depending on the request of potential customers. It was found that the LNG plant, consisting of 6 lines of 5500 Nm3/h, has the greatest flexibility, which can provide both the minimum possible load of the plant, and the maximum, exceeding the required one by 30%.

It has also been observed that turbo expander and supersonic separator have about the same potential. However, these values are within the margin of error. Considering the materials required for the 3S separator, special non-brittle and cold resistant materials will be required.
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Introduction

The growth of energy resource consumption in the world leads to increased production of gas fields and the development of ways to transport energy resources to the consumer. Thus, according to the Russian Ministry of Energy for 2018, total gas production in Russia increased by 5.0% (+34.3 bcm by 2017) and reached a new record level for the whole period of Russian gas production - 725.4 bcm.

At the same time, according to the strategic development plan of PJSC “Gazprom” until 2025, as well as according to the strategy of PJSC “Gazprom Neft” until 2030 it is planned to introduce LNG as a marine fuel[1].

The world market share of LNG was 70 million tons per year in 1995, 111 million tons in 2002, 293 million tons in 2017, and 357 million tons in 2020. The growth is driven by technology development, resulting in lower unit costs for LNG production, i.e., lower LNG costs. It should be noted that as of 2017[2], Russia’s share of global LNG was only 4.0% (11.76 million tons).

Low-tonnage LNG provides greater flexibility and speed: with access to multiple sources of feedstock and no need to include production and deep gas treatment in one project.[3] The low-tonnage LNG project is implemented within six months and the payback period of the project itself takes up to four years depending on the planned profit volumes, which is much shorter than the time required to implement a large-scale LNG project[4].

Low-tonnage LNG is focused on the local natural gas consumption market. Thus, LNG plants will be located close to the customer, which will be equipped with a regasification point. At the same time, the construction of the production facilities is not time-consuming as the modular design is expected. Full installation of all modules takes about two weeks. Minor deviations from the design schedule may occur depending on the complexity of the project and the availability of the production site.

Delivery of low-tonnage LNG to the consumer can be organized on all types of transport, with virtually no restrictions on the distance and location of the end-user. It will also allow LNG

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[1] Gazprom Innovative Development Program Passport until 2025. Moscow: Gazprom, 2016. - 77 c.
[2] FSRH CEU response to the decision of NICE not to update https://www.fsrh.org/standards-and-guidance/documents/fsrh-ceu-statement-regarding-the-decision-of-nice-not-to-update/fsrh-ceu-response-to-the-decision-of-nice-not-to-update-larc-guidelines-11dec.pdf
[3] Arkharov A.M. Why the exergic variant of thermodynamic analysis is not rational for the study of basic low-temperature systems / A.M. Arkharov // Cooling Engineering. 2011. - No 10. C. 8-12.
[4] BusinessWorld | Energy department lists sites for new https://www.bworldonline.com/content.php?id=146630

producers to use local pricing models, thus encouraging consumers to save money by switching to LNG.

The additional use of LNG as a motor fuel can be a decisive factor in the success of the LNG industry. However, for LNG plants with low tonnage capacity, the ability to produce high methane LNG for the NGV segment will enable them to compete with LPG producers.

Thus, low-tonnage liquefied natural gas (LNG) production is an industry with great potential. Considering the government program of gasification of rural population centers until 2025. It is being implemented in the Russian Federation since 2001, up to 85% from 68.6% as of 2018. The growth of sparsely populated settlements at a relative distance from the existing gas distribution or trunk networks. Also, the issue of developing this alternative method of gasification of population centers becomes urgent.

In accordance with the above, the purpose of the thesis work was to study the operating modes of the existing basic scheme of a low-tonnage liquefying natural gas plant.

LNG relevance

The liquefied natural gas (LNG) market will play an important role in tomorrow's energy market as both gas and renewables become the fastest-growing energy segments in the coming years.

Although fossil fuels in the aggregate are set to decline in the coming decades, significant environmental benefits can be achieved by replacing coal with gas. Natural gas is the cleanest of fossil fuels - it has a high energy density, but emits the least CO2. Natural gas is an excellent alternative for reducing greenhouse gas emissions and tackling global warming.

The role of pipeline gas has declined over the past two decades, from 73 percent of the total gas shipped in 2000 to around 65 percent in 2020. China and Russia are expected to significantly increase LNG imports and exports over the next five years because pipeline capacity will be insufficient.

According to McKinsey Global Gas & LNG Outlook 2035, LNG demand could grow by 3.6% a year between 2018 and 35. S&P Global Platts Analytics also expects global LNG demand to grow by 2% in 2020 to around 362 million tonnes despite the pandemic and another 3% in 2021. This is much slower than the 11% market growth in 2019 and double-digit percentage growth in previous years, as demand growth in Asia, led by China, compensates for the decline in Europe.

This experience with liquefied petroleum gas production in Russia, plans are also in full swing to develop an LNG network in small- and medium-tonnage capacity.

In the current global pandemic situation, it is reported that a South Korean company is planning to use LNG cooled warehouses for mass storage of COVID-19 vaccines, as conventional cooled ones cannot cope.

New issue how to use LNG found in Asia, where gas exporters are looking to the industrial, petrochemical, urban gas and transportation sectors to become significant sources of demand, complementing the traditional use of LNG in the energy sector.[5]

Objective

This research aims to optimize and improve a low-tonnage liquefied natural gas plant based on the nitrogen cycle. This is done by selecting the optimum performance of the LNG plant operating modes, as well as analysing the efficiency of the reduction units and proposing possible options for improving the cooling process in it based on promising technologies (turbo expander and supersonic separator).

[5] Kadiev, V.A. Gasification of motor transport - the problem of ecological and energy security, the state / V.A. Kadiev. 2003. URL http://www.mosavtogaz.ru/pressa3.html

Chapter 1: Literature review

This chapter consist of the main equipment used in a natural gas liquefaction facility and the different gas treatment methods and liquefaction technologies.

In most cases, a natural gas liquefaction complex consists of gas treatment units, a dehydration and a CO2 treatment units, with heat-exchange equipment, compressor units, separators, and a gas reduction unit. This is the basic set of equipment required for the operation of the plant and further on the technology of each type of equipment in more detail.

1.1. Crude gas treatment unit

The natural gas treatment unit in this scheme is based on adsorption technology. It consists of adsorbers in which the raw gas is purified from carbon dioxide. It should be noted that there are other methods of gas purification, which are discussed in the following paragraphs. The main methods for purification are absorption, adsorption, and membrane purification of the raw gas.[6]

1.1.1. The absorption method

The absorption process consists of the sorption of the individual components of the gas mixture with a liquid absorbent. The absorbent is chosen according to the solubility of the substance to be removed from the gas mixture. For example, water is used as an absorbent for cleaning gas from ammonia, hydrogen chloride, hydrogen fluoride, sulphuric acid for absorbing water vapor, and heavy fractions of alkane hydrocarbons for absorbing aromatic hydrocarbons.

Absorption involves the convective diffusion of vapor and gas components into the liquid absorbers. High-quality purification of the raw gas from undesirable components (carbon dioxide) can only be achieved by ensuring a complete offset of the pure gas and the absorbent. This purification is achieved by the design of the internal structural elements of the mass exchange apparatus (absorption columns).

Regeneration of the absorbent is reached by reducing the total (or partial) pressure of the unwanted component (carbon dioxide), by increasing the temperature or by combining both methods.

[6] Izotov, N.I. Liquefied natural gas. Technologies and equipment / N.I. Izotov. Moscow: Gazprom VNIIGAZ, 2013. 306 c.

1.1.2. The adsorption method

The adsorption method is based on the sorption of unwanted gas impurities through the surface of solid particles. Adsorption is used for deep purification of raw gas from vapor and mainly gaseous unwanted impurities (vapor solvent, ether, acetone, various hydrocarbons). Adsorption processes take place on the surface of a solid absorbent, which interacts with the extracted components and traps them due to intermolecular interaction.

For example, activated carbon is the best known and most widely used adsorbent. It is used to purify water vapor (or water, in the case of liquid adsorption) from organic vapors and some other impurities. Activated alumina and silica gel are also used. The process of adsorption gas purification is carried out by solid static layers of adsorbents and moving layers.

To sorb unwanted components from natural gas and inert them (if possible) into harmless substances, the chemical and physical properties of the adsorbent must be correctly selected. Thermal destruction and oxidation of substances that neutralize flammable components from the gas are the most frequently used[7].

Adsorbents must meet the following requirements:

- high adsorption capacity for the adsorption of components;
- high selectivity;
- high mechanical strength;
- recoverability;
- low cost.

The following adsorbents are currently used:

- activated carbon;
- silica gels;
- aluminum oxide;
- zeolites.

Silica gels are used to dry gases. Silica gels are usually manufactured as granules ranging in size from 0.2 to 7 mm in diameter with a bulk density of 400 to 900 kg/m.

Aluminum gels (or aluminum oxide) are used in drying gases and capturing other polar organic substances (e.g., methanol) from them. Today, granular cylindrical aluminogels with a granule diameter of 2.5 to 5 mm as well as spherical shapes with a granule diameter of up to 3.4 mm are produced.

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[7] Visual Encyclopedia of Chemical Engineering. http://encyclopedia.che.engin.umich.edu/Pages/SeparationsChemical/Adsorbers/Adsorbers.html

Zeolites are aluminosilicates that contain alkali and alkali earth metal oxides. They are divided into two groups:

- Natural;
- Synthetic.

Synthetic zeolites are produced in spherical granules with a particle diameter of about 2.5 mm and cylindrical form with particle diameter and length of 2.4 and 2.4 mm respectively. Synthetic zeolites of types A, X, and Y are characterized by high regularity of crystalline structures like natural zeolites and other adsorbents. Nowadays, many varieties of zeolites are produced, and their characteristics are constantly being improved. Zeolites are used for the separation of hydrocarbon mixtures; numerous effective multi-base catalysts are produced on their basis. Zeolites being water adsorbers for the deep drying of gases of types NaA and NaX have bulk density of 680-730 and 590-620 kg/m3 at a proper density of 2700-2830 kg/m3 and adsorption limit volume of 0.25 cm3/g; they are produced in the form of powder, granules, and balls of 1.6-3.2 mm. Five types of zeolites are distinguished by the domestic (KA, NaA, CaA, CaX, and NaX) and American (3A, 4A, 5A, 10X, and 13X) classifications, with inlet pore sizes of 0.3, 0.4, 0.5, 0.8 and 1.0 nm, respectively.

Figure 1: Zeolite NaX granular sorbent (source: silikagel.ru)

Figure 2: NAA sorbent (source: silikagel.ru)

1.1.3. Membrane purification

This is a technology for purifying natural and petroleum gases, which may also contain H2S. The technology includes, as one of the main steps, the purification of hydrocarbon components from the so-called acid gas mixture of CO2 and H2S, the content of which in some cases can be up to 40-45% vol. The technology is an alternative method for the isolation of acid gases. The choice of the membrane CO2 extraction method depends on its volume fraction in the raw material. There are several reasons for the prevalence of the membrane method of gas treatment:

- The outgoing gas is pressurized, so there is no need to install compressors;
- The stream enriched with recoverable acidic components can be used in case of oil and gas condensate field (OGCF) raw gas treatment, then directly in the field to enhance oil recovery from reservoirs and depleted wells;
- The use of membrane technology allows for purified and dry gas to be produced to the required standards without additional post-treatment.

It is generally assumed that membrane processes effectively treat natural gas and associated petroleum gases to remove the bulk of the impurities. However, for finer further purification, absorption or adsorption processes need to be applied. When comparing the two-stage membrane method with diethanolamine absorption (DEA), the costs of the membrane process are lower, even if the membrane plant is operated under unfavorable conditions. The membrane process's investment and operating costs are 25 or even 60 % lower than the absorption with DEA water solutions. This method is particularly efficient for CO2 retention. Nevertheless, when deep CO2 abstraction of the feed gas is needed, the membrane process is less efficient both technologically and economically.[8]

1.2. Crude gas drying unit

The natural gas drying unit in the considered scheme is also based on adsorption technology; it consists of adsorbers in which water is sorbed from the raw gas stream. Nevertheless, there are other methods of gas dehydration that are considered in the following paragraphs. The main methods for drying are absorption and adsorption.

1.2.1. Absorption method

One of the most crucial raw gas drying methods is glycol drying based on the absorption method.

[8] Batuner, L.M. Mathematical methods in chemical engineering / L.M. Batuner, M.E. Pozin. L.: Chemistry, 1968. 823 c.

Glycol drying of gases is the selective absorption of water vapor using a special liquid absorber. Today, mono-ethylene glycols, diethylene glycols, and Triethylene glycol are used as absorbents in practice, capable of absorbing moisture up to 40 g/l and acid gases up to 99 % of the initial values.

Generally, the raw natural gas produced from Russian fields is saturated with moisture and sometimes waterlogged. Moisture in the gas can cause problems during extraction, treatment, and onward transmission. For example, natural gas hydrates form at low temperatures and frozen water in the pipeline, or solid hydrates form with other gas components.

Such compounds can also occur at high temperatures under the high pressure. The most significant risk is from possible hydrate deposits in the pipelines and process equipment. Glycol drying of gas can remove water and alter the thermodynamic equilibrium by lowering the hydrate formation temperature.

Glycol drying uses the partial pressure difference between the water vapor in the absorbent and the gas. The amount of moisture that can be removed from the gas by the absorbent is determined by the absorbent properties of the desiccant, the temperature and pressure, and the effective contact area[9].

1.2.2. Adsorption method

Deep drying of gases is achieved up to the dew point temperature (down to minus 20-40 °C on silica gels, up to minus 40-50 °C on alumina gels, and up to minus 60-70 °C and below on zeolites). The adsorption efficiency, like absorption, increases with decreasing temperature and increasing pressure, and vice versa. Regeneration of adsorbents is carried out at reduced pressure and high temperature (180-220 °C and higher, depending on the nature of the adsorbent). Regeneration of the adsorbent can also only be carried out by a stepwise pressure reduction to 0.350-0.007 MPa without changing the temperature in the adsorber.

1.2.3. Heat exchange equipment unit

The heat exchange equipment unit consists of two cooling stages, i.e., two multi-pass heat exchangers.

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[9] Replacing Glycol Dehydrators With Desiccant Dehydrators. https://www.epa.gov/sites/production/files/2016-06/documents/ll_desde.pdf

1.2.3.1. Purpose of the heat exchange apparatuses

The purpose of heat exchangers is to transfer heat from one medium to another or from the environment to the heating or cooling agent. It should be considered that heat exchange equipment transfers energy from a hot medium (raw material gas) to a cold medium (refrigerant - nitrogen)[10]. Thus, without additional equipment (compressor-expander and compressor-throttle systems) ensuring the required refrigerant temperature, heat exchange equipment does not fulfill its function.

When designing and constructing heat exchangers, it is necessary to satisfy as many and in most cases, conflicting requirements for heat exchangers as possible. The main of them are compliance with technological process conditions; the highest possible heat transfer coefficient; low hydraulic resistance of the apparatus; corrosion resistance of heat transfer surfaces; availability of heat transfer surfaces for cleaning; manufacturability of design; economical use of materials.

First of all, heat exchangers are classified according to the method of heat transfer from one medium to another (from one heat carrier to another):

- recuperative (surface);
- regenerative;
- mixing (contact);
- electrically heated;
- radial-spiral type.

1.2.3.2. Multi-pass heat exchanger

Surface heat exchangers are the most common and most extensive group of heat exchangers in the industry. In surface heat exchangers, the heat transfer medium is separated by a wall and the heat is transferred through the surface of that wall. If the heat exchange surface in such heat exchangers consists of tubes, this type of heat exchanger is called a tube heat exchanger.

In the chemical and petrochemical industry, surface tube heat exchangers in horizontal design are most used. They are easy to manufacture, have a large heat exchange surface in a single unit casing, and are reliable in operation.

This type of heat exchangers is the most common due to its simple design and production technology. The following technological standards produce tubular heat exchangers:

- with fixed tube sheets and rigid casing;

[10] Compact Heat Exchangers. https://heatexchanger.ae/compact-heat-exchangers/

- with fixed tube sheets and temperature compensation on the casing;
- with fixed tube sheets and U-shaped heat exchange pipes;
- with floating head;
- with gland on the floating head.

Heat exchangers depending on their purpose can be:
- preheaters;
- coolers;
- condensers;
- evaporators.

To increase the velocity of flow of coolants, tube heat exchangers are designed as two-, four-, six- or twelve-way units.

1.2.4. Dynamic equipment unit

Dynamic equipment in the proposed technological scheme includes compressor units, as well as turboexpander.

1.2.4.1. Multistage axial compressors

Figure 3 is a schematic representation of the main components of an axial multistage compressor.

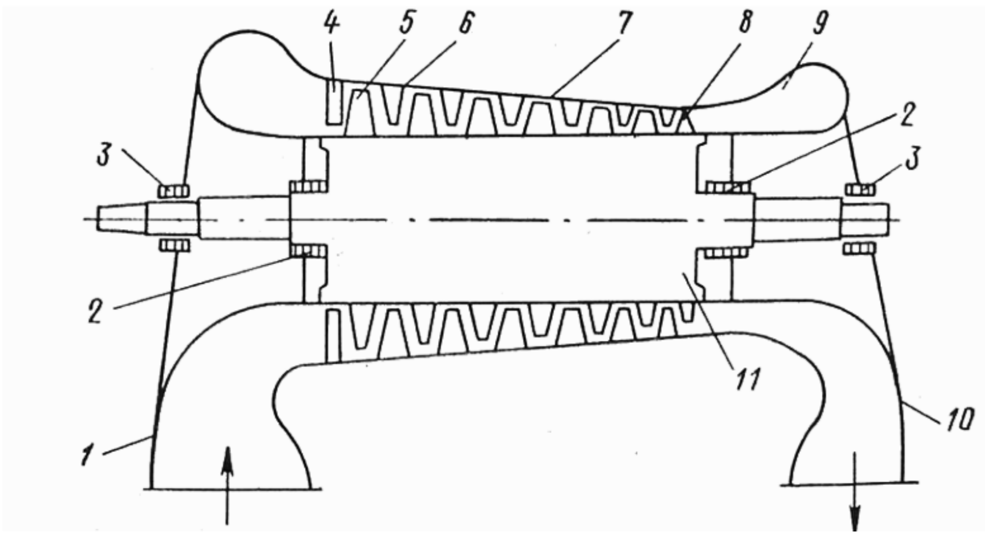


Figure 3: Schematic diagram of an axial multistage compressor

This type of compressors consists of a group of guide vanes 6, rigidly fixed to the inner surface of casing 7, and impeller 5, fixed to the rotor 11. The height of the guide vanes and working vanes is reduced as the volume of raw gas decreases during the compression process.

The raw gas is fed to the compressor through nozzle 1 and a guide device 4, which provides the necessary flow direction in the unit. The raw gas is then given kinetic energy by rotating the working blades. As the raw gas flows through the space between the blades, which increases in the course of gas movement, the relative flow velocity decreases, so that the kinetic energy is converted to potential energy, i.e. the gas pressure increases. This is ensured by controlling the amount of energy transferred to the unit.

The gas velocity decreases with an accompanying increase in pressure due to the larger flow area in the channels between the guide vanes. The compressed gas then enters the straightener 8 to combine and stabilize the flow and, via the diffuser 9 and outlet 10, is discharged into the discharge line.

The number of compression stages in such compressors is provided by a necessary number of working blades 5.

1.2.4.2. Multistage piston compressors

For reception of compressed gas of higher pressure (0,1-2 MPa and above), multistage compressors with intermediate cooling of gas after each stage are applied.[11]

The essence of multistage compression can be explained on the example of a two-stage compressor, the scheme of which and the ideal (at $V_0 = 0$) indicator diagram is presented in figure 4:

[11] PB 03-581-03 Rules for the installation and safe operation of stationary compressor units, air pipelines and gas pipelines. Moscow: Ministry of Justice of RF, 2003. - 12 p.

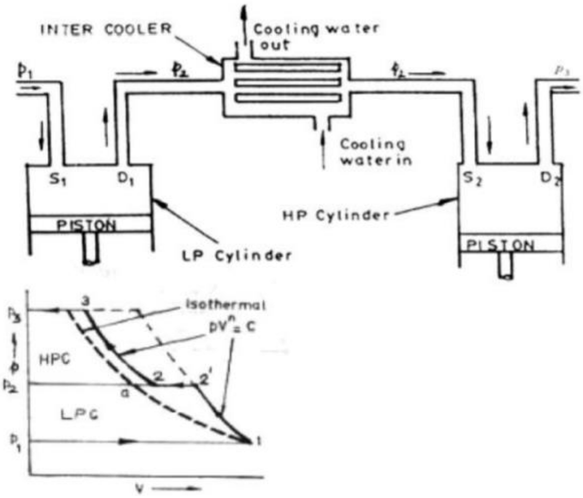


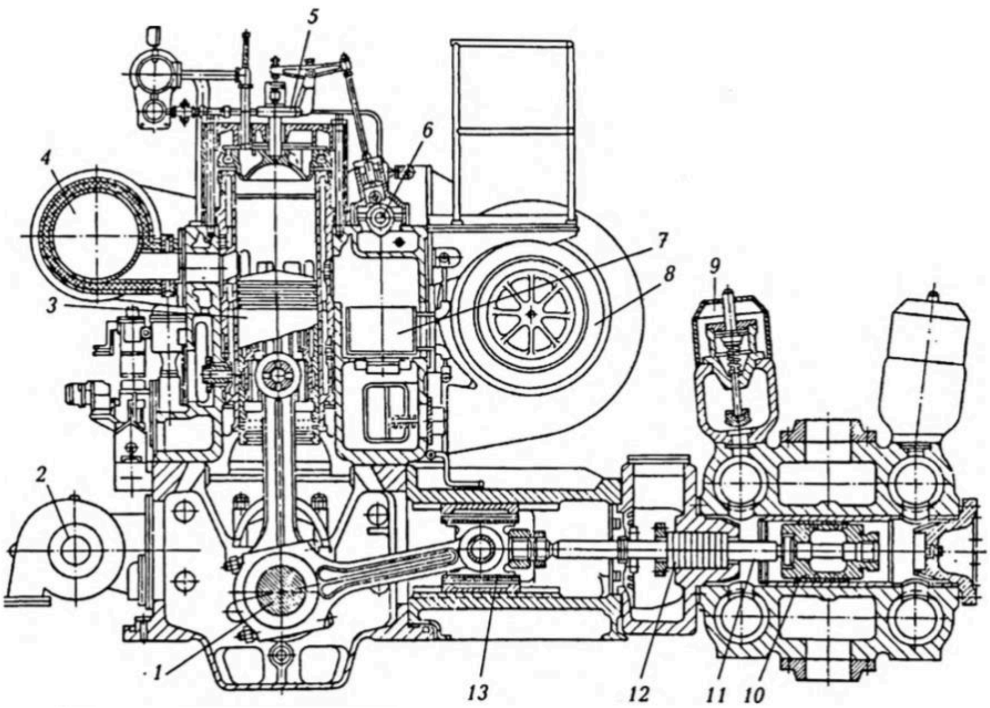
Figure 4: Display diagram of the compression stages of a block of reciprocating compressors

In the first stage 1, the gas is compressed along with a polytrope 1-2’ to pressure P2, and then it enters the intermediate cooler 2, where it is cooled to an initial temperature T1. The hydraulic resistance of the refrigerator along the air duct is made small. This allows considering the isothermal cooling process 2-3.[12]

After the refrigerator, gas enters the second stage 2, where it is compressed by polytrope 3-3’ to the pressure P3. If compression to pressure P3 were performed in an ideal single-stage compressor (line 1-2’).

In two-stage compression with intercooling, this work is numerically equal to area 12’233’a. The shaded area corresponds to the saving of cycle work in two-stage compression.

[12] Plastinin, P.I. Piston compressors. In 2 vols. Vol.1 Theory and calculation: Textbook for universities / P.I. Plastinin. Moscow: KolosS, 2006. 456 p.



1 - crankshaft; 2 - water pump; 3 - engine piston; 4 - exhaust manifold; 5 - exhaust valve; 6 - camshaft; 7 - charge air cooler 8 - turbocharger; 9 - capacity regulator; 10 - compressor piston 11 - piston rod; 12 - oil seal; 13 - crosshead

Figure 5: General view of piston compressor

1.3. Process equipment

The process equipment in the selected layout includes separators.

Separators are the most common type of process equipment and they are required for cleaning and phase separation of oil and gas. In the oil and gas industry, there are two main types of separation plants: oil and gas separators and gas separators. The former degasify oil to produce degassed crude oil and associated petroleum gas. Gas separators are designed to separate the liquid and gas phase of natural gas, as well as its purification, i.e., the process of gas preparation for storage and transportation.

A distinction should also be made between different types of separation plants and the most common in natural gas liquefaction is low-temperature separation. This type of separator is used in the gas liquefaction process to separate heavy hydrocarbons from the natural gas and remove residual moisture from it. This takes place at low temperatures and the water vapor phase condenses into crystalline hydrates, which are extracted together with the heavy hydrocarbons in the separation plant.

In low-temperature gas condensation, cooling is maintained only to a certain degree of condensation of the raw gas. This is determined by the required depth of extraction of necessary components from the gas and is achieved by a precisely defined end temperature of the cooling process (depending on the composition of the feed gas and the pressure in the system). Low-temperature condensation ensures deep recovery and high purity of the final product and is the most economical of all processes currently used.

1.3.1. Nitrogen cycle

The natural gas liquefaction technology considered in this paper is based on a nitrogen gas cooling circuit. It is mainly used in low-tonnage plants, particularly in the power generation sector, to provide load balancing. It is used for energy supply, fuel supply for almost all modes of transport from onshore to offshore, as well as for gasification of rural communities.

A closed nitrogen cycle, i.e., with a fixed volume in the system without external communication, is mainly used to liquefy natural gas. The nitrogen passes through three stages in such a scheme: pre-compression, main compression, and expansion (throttling). The resulting cold is transferred to the feed gas via multi-pass heat exchangers. This is the process by which the feed gas is liquefied using a nitrogen cooling cycle.

1.3.2. Low-temperature gas separation

Low-temperature separation (LTS) is the process of field treatment of natural gas to extract gas condensate and remove moisture from it, carried out at temperatures between 0 and -30 degrees Celsius, while for natural gas liquefaction this temperature can be as low as -165 degrees Celsius.

Gas cooling in field plants is done using excess pressure when the gas pressure is triggered at chokes or turbo expander units. When gas is throttled, depending on its composition, it is cooled by 3-4.5ºC at a pressure drop of 1 MPa. With the cold recovery of the gas flow after the throttle in gas-to-gas heat exchangers, it is possible to reduce the gas temperature to minus 25°C or more if large amounts of cold energy from other gases or liquids are used.

The entire LTS process is reduced to the cooling of natural gas, followed by separating the gas condensate mixture in a separator into liquid and gas phases.

A refrigeration unit is included in the circuit when the gas pressure drops and therefore, the specified separation temperature cannot be achieved.

The technological model of the LTS plant is determined by the pressure potential, gas and condensate composition, and field product requirements. The pressure of the last separation stage is taken based on pipeline pressure and the temperature based on the amount of moisture and heavy hydrocarbons release.

Low-temperature separation technology is suitable for any climatic zone, allows for non-hydrocarbon components in gas, provides for condensate C_{5+} recovery rate of up to 97%, and a dew-point temperature that prevents precipitation of water and heavy hydrocarbons in natural gas transportation.

The advantage of the LTS unit is low capital and operating costs (in the presence of a free differential pressure). The disadvantages are a low recovery of condensate-forming components from lean gases, a continuous decrease of gas recovery efficiency during operation, and the necessity of radical reconstruction during depletion of the throttle effect.

Speaking about the efficiency of LTS units, the feed gas composition, temperature, pressure, equipment efficiency, and a number of separation stages have had and still have a major influence. Regarding gas composition - the heavier is the feed mixture composition (the higher is the average molecular weight of gas), the higher is the degree of liquid hydrocarbons recovery. However, starting from a certain gas composition (average molar boiling point about minus 133ºC, molecular weight about 22), the heavier the initial mixture composition has almost no effect on the degree of recovery of C_{3+} components.

The temperature in low-temperature separation plants is selected based on the required dew point for transporting gas through the pipeline in a single-phase state. For light gases (average molecular weight not exceeding 22, average molecular boiling point minus 156 to minus 133ºC), a decrease in separation temperature from 0 to minus 40ºC ensures a considerable increase in condensate-forming components' recovery as C_{3+}, $C_{4+,}$ and C_{5+}. For gases with high gasoline vapor content, however, the influence of temperature on the recovery of liquid hydrocarbon fractions is small.

The pressure for the separation is determined by the pressure at the booster compressor station located immediately downstream of the LTS unit (within the range of commonly used pressures of 5-7.5 MPa). The initial pressure generally has little influence on the degree of recovery of C_{5+} components. More important is the free pressure difference which allows reaching low separation temperatures.

Turning to equipment efficiency, more attention is paid to the cooling sources used or, for more extended wells, to the refrigerants selected, namely its flow rate in the evaporator and heat exchange surface.

Regarding the number of separation stages, it can be noted that usually LTS schemes are presented in one, two, or three stages, where the first stage is the most effective. It is in the first separator that most qualitative separation of gas and liquid flows takes place.

All these factors even now play a decisive role when choosing a scheme of low-temperature separation of “blue fuel.” The efficiency of both the plant and the separation process as a whole has been the subject of lengthy debate. Given the vast geography of gas fields and the growing trend towards offshore and offshore development of gas-rich locations, this is not surprising.

By 2021, low-temperature separation plants have undergone several changes. Thanks to the latest technology, sophisticated equipment combination schemes and different thermal design of the reduction unit, deeper separation, and separation of C_{2+} fractions is now possible. Based on numerous sources from the last 10 years, comparative analyses of possible LTS schemes are carried out, the first landmark improvement in gas separation and gas condensate separation was the Joule-Thomson effect, which will be discussed below.

1.3.3. The Joule-Thomson effect

The British scientists Joule (1818 - 1889) and Thomson (1824 - 1907) studied the temperature change of natural gas during adiabatic expansion. In other words, they observed the effects of changes in potential energy of repulsive and attractive forces of gas molecules during its expansion process.[13][14]

This experiment (Fig. 6) the gas through a thermally insulated bamboo tube passed through a porous cotton swab from the high-pressure region P_1 to the low-pressure region P_2 and expanded.[15][16]

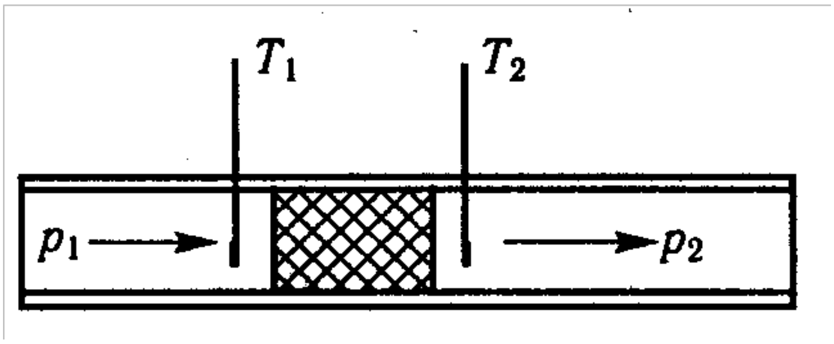


Figure 6: Schematic of a Joule-Thomson experiment set-up

[13] B. N. Roy (2002). Fundamentals of Classical and Statistical Thermodynamics. John Wiley & Sons
[14] F. Reif (1965). "Chapter 5 – Simple applications of macroscopic thermodynamics". Fundamentals of Statistical and Thermal Physics. McGraw-Hill. ISBN 978-0-07-051800-1
[15] W. C. Edmister, B. I. Lee (1984). Applied Hydrocarbon Thermodynamics. Vol. 1 (2nd ed.). Gulf Publishing. ISBN 978-0-87201-855-6
[16] W.R. Salzman. "Joule Expansion". Department of Chemistry, University of Arizona, 2012

The left side of the unit was pressurized with a positive displacement pump and the right side was pressurized with a vacuum pump. The thermometer readings in both parts were used to determine the temperature change with gas expansion.[17]

Experiments at room temperature showed that almost all gases, expanding, cool down. The exceptions were hydrogen and helium, which warmed slightly as they expanded.

It was also found that the intensity of the effect depended on the gas composition, pressure drop, and initial gas temperature. For example, the temperature drop per 100 kPa of pressure drop was 0.25 K for air, 1.3 K for carbon dioxide, and for hydrogen, the temperature rise was 0.3 K.[18]

Later, more accurate research has shown that there is a functional relationship between P and T in the Joule-Thomson effect. At some values of P and T, this effect was not observed at all. Obviously, in this case, conditions were created when the molecular forces of repulsion and attraction in the gas expansion were mutually compensated. At high pressures, the repulsive forces of the molecules were stronger, so the observed effect was assumed to be harmful, and at low pressures, the forces of attraction were stronger, so the observed effect became positive.

Later on, the positive Joule-Thomson effect became the basis for a method of liquefying gases. This method was reduced when the compressed gas was expanded through a porous partition or slot, the work of gas expansion was carried out due to its internal energy, so the gas was cooled. This effect was initially used in an industrial air liquefaction machine created by German scientist K. Linde in 1895 (Fig. 7).[19]

Compressor K sucks atmospheric air through valve S and pressurizes it into tank A up to pressure 20 MPa. The tank is cooled with water; the compressed and partially cooled air enters through coil B into vessel C, expanding and further cooled. The cold air from vessel C enters through pipes b1, b2 enclosing the coil, and b3 to the compressor and is pressurized again in the tank.

[17] Atkins, Peter (1997). Physical Chemistry (6th ed.). New York: W.H. Freeman and Co. p. 930. ISBN 978-0-7167-2871-9
[18] Atkins, Peter (1997). Physical Chemistry (6th ed.). New York: W.H. Freeman and Co. pp. 86–90. ISBN 978-0-7167-2871-9
[19] DE 88824, Linde, Carl, "Verfahren zur Verflüssigung atmosphärischer Luft oder anderer Gase", issued 29 September 1896

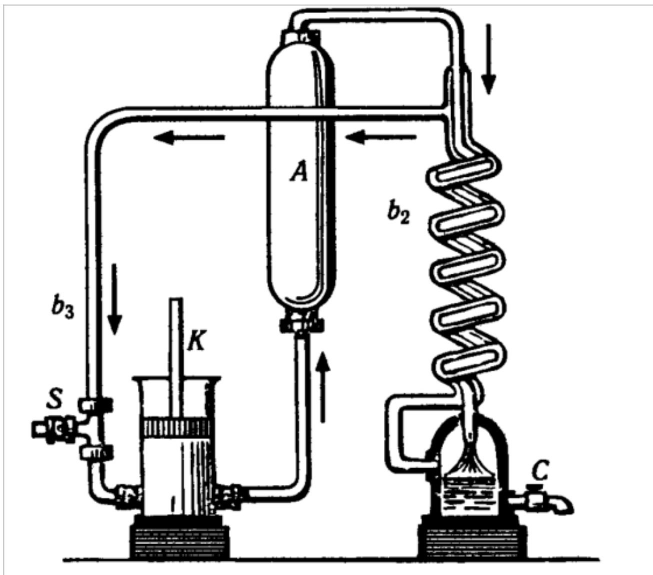


Figure 7: C. Linde industrial air liquefaction machine

This ensured that the air was circulated. Each subsequent expansion of the air in vessel C caused a further drop in temperature. The air exiting vessel C was cooling the incoming air, which was flowing through the coil. After some time, the air began to condense.[20]

As noted above, the behavior of hydrogen and helium is different from other gases. The Linde machine could not be used to liquefy these gases due to very low inversion temperatures for these gas representatives.[21]

Today, the Joule-Thomson effect in the oil and gas industry is better known as the throttle effect, and the value of the change in temperature of a gas when its pressure drops by 0.1 MPa is referred to as the Joule-Thomson factor. For an ideal dry gas, this coefficient is about 0.3ºC. However, real gases always contain moisture and heavy hydrocarbons, which, when lowered, turn into a liquid state, releasing the latent heat of condensation. Therefore, under real conditions, the Joule-Thomson coefficient is 0.15-0.25ºC.

The Joule-Thomson effect can often be observed on gas pipes and equipment, which become covered by a white deposit in the form of frost or snow. This fouling is formed from ambient air moisture condensing on metal surfaces cooled by gas due to pressure reduction on connectors, gate valves, expansion in apparatus, changes in the diameter of the pipeline, etc.

[20] de Waele, A. T. A. M. (2017). "Basics of Joule–Thomson Liquefaction and JT Cooling" (PDF). Journal of Low Temperature Physics. 186 (5–6): 385–403

[21] Linde, Carl, "Process of producing low temperatures, the liquefaction of gases, and the separation of the constituents of gaseous mixtures", 1903

1.4. Equipment in the reduction unit

1.4.1. Turbo expander units

The turbo expander is a continuous blade turbine machine. The turbo expander expands the gas to cool it down. The energy released allows useful external work to be carried out. The turbo expander is directly involved in liquefying the gas and separating the target components.[22][23]

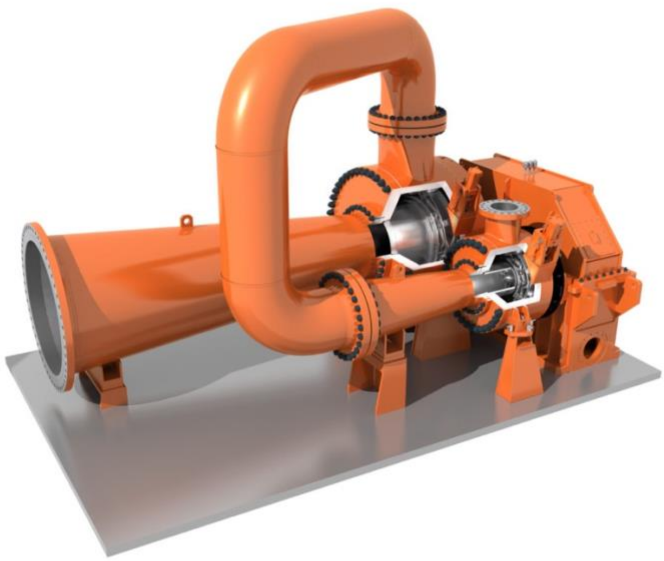


Figure 8: Turboexpander unit (source: akyudesign.com)

The turbo expander (Fig. 8) consists of a casing, rotor, adjustable nozzle apparatus, and guide apparatus equipped with rotating mechanisms. The unit is completely sealed and requires no electrical power.[24] The direction of the moving gas flow determines its design. Therefore, turboexpanders can be centrifugal, centripetal, and radial (axial). There are different degrees of gas expansion in the nozzles. Therefore, turboexpanders are divided into active and reactive. In the first case, the pressure is reduced only in the stationary guide channels, while

[22] Arkharov A.M. Development of liquefied natural gas technology in Moscow region / A.M. Arkharov // Bulletin of N.E. Bauman Moscow State Technical University. Mashinostroenie. 2010. Cryogenic and Refrigeration Technique: Special Issue. C. 214-229.
[23] Semenov, V.Yu. The results of experimental studies of the cryogenic expander-compressor / V.Yu. Semenov // Chemical and Oil and Gas Engineering. 2009. No. 4. C. 23-25.
[24] Wetherston, R. Energy exchanger - a new concept in the theory of high-efficiency gas-turbine cycles / R. Wetherston, A. Herzeberg // Power Machines. 1966. T. 4., No 2. C. 48-62.

in the second case, the pressure is also reduced in the rotating channels of the rotor. The units can be designed as single or multistage units.[25]

Gas or liquefied gaseous mixtures pass through the openings of fixed guide channels which act as nozzles. At this point, the potential energy of the gas is partially converted into kinetic energy, which drives the rotating vane channels of the rotor. The rapid expansion of the gas causes the pressure to drop, which in turn causes the rotor to work mechanically, cooling the gas flow at the same time. Simultaneously with the rotor, the compressor wheel attached to it rotates.

When handling liquefied gas mixtures, it is important to consider the possibility of droplet erosion - wear of the surface of the vanes due to droplets of liquid hitting the surface at high speed. The consequences of this phenomenon are considerable. The main one is the reduction of blade reliability due to:

- Higher bending and tensile stresses due to the reduced cross-sectional area of the blade;
- Decreasing the fatigue strength because of worsening surface quality and increasing stress concentration.[26]

As a rule, constant pressure is maintained at the turbine inlet in accordance with the design level. In this case, the pressure regulation is carried out by special valves which are not quite rational. Turbines with variable pressure with fully open inlet valves are considered to be more efficient. The valves used must be as large as possible, making it possible to achieve the necessary throttling at differential pressures of only 5-10%. For traditional valves, this figure is 25-50% due to the small dimensions. The same applies to pumps that create gas pressure. They are selected according to the specific operating conditions.[27]

LTS with the application of turboexpander units (TEU) has been widely spread at the Russian gas condensate fields of the Far North. This treatment method makes it possible to ensure water and hydrocarbon dew-point requirements to the transported gas and cool gas before it is supplied to the trunk pipeline. This eliminates a separate facility such as a gas cooling station. Gas cooling with TEU allows the available differential pressure to be used as efficiently as possible.

[25] Kozlov A.V. Increase of operation efficiency of expanders in the installations of low-temperature treatment of hydrocarbon gas: Ph.D. in Technical Sciences: 05.02.13 / Kozlov Alexey Valerievich. M., 2003. 161 c.
[26] Studizba. Drip erosion of working blades. URL: https://studizba.com/lectures/129-inzhenerija/1941-nadezhnost-raboty-turbinnogo-oborudovanija/37952-6-kapelnaja-jerozija-rabochih-lopatok.html
[27] The principle of turboexpander action. ELECTRIC-220.RU. URL: https://electric-220.ru/news/princip_dejstvija_turbodetandera/2016-06-06-972

Units with a vertical design are the most promising due to several advantages: smaller footprint; technological design allows quick (10-15 minutes) change of the flow part; availability of bearing relief effect during TEU operation (axial compressor force acts in the direction opposite to the gravity force).

The choice of turbine type is also based on the specifics of the plant: the capacity of the process lines, the moisture content of the gas and the heavy hydrocarbons that condense. Here, axial turbines are less subject to erosion wear at liquid condensation because due to the axial direction of flow, the condensing liquid is carried away from the flow part and it comes into contact with structural elements of the flow part for much less time.

The turbo expander unit can also be subdivided according to the values of operation indicators: capacity, pressure, and temperature, a ratio of pressures in the turbine πT, ratio of pressures in the compressor πK, etc. The existing TEUs are operated at working pressures 12,2-5,6 MPa, πT up to 1,1-2,1 units, and productivity 5-20 mln. m^3/day. At the same time, the units provide cooling of gas at 30ºC and more, which in schemes with cold recovery allows reaching a temperature of minus 30ºC in summer and up to minus 40ºC in the autumn-winter-spring period.[28]

In spite of the impressive list of advantages of this technology, turbo-expander units no longer meet the challenges of reducing capital and operating costs during the construction and development of fields, not to mention the requirements to the quality of gas separation. To address the new gas treatment issues, supersonic separators have moved on.

1.4.2. Supersonic separators

The 3S technology is based on the expansion of a pre-curved gas stream in a Laval nozzle. In this case, due to the transfer of part of the potential energy of the gas flow into kinetic energy at its acceleration to supersonic speeds, there is a sharp drop in static pressure in the flow, which is accompanied by its strong cooling.

In 3S-separator, it is possible to regulate the temperature of flow in the nozzle by calculating nozzle geometry and pressure drop ratio (usually characterized by Mach number M). The number of M is selected[29] so the “target” (separable) components of the gas mixture will pass into the liquid phase. The resulting droplets actively coagulate, helped by the turbulent nature

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[28] Vorontsov M.A., Fedulov D.M., Grachev A.S., Prokopov A.V., Glazunov V.Yu. Methodical approach to the calculated study of field preparation of natural gas for transportation by low-temperature separation technology using turbo expander units // Vesti gazovoy nauki. 2016. №2 (26). Pp. 105-111;

[29] S.Z. Imaev, M.I. Safyannikov. Control of supersonic separators // Territory Neftegaz, No. 9, 2016. p. 98-104

of the swirling flow. The degree of flow twist[30] is usually selected to ensure centrifugal forces of the order of 10^5 g in the nozzle operating section.[31]

Centrifugal forces force liquid droplets into a boundary layer, where a gas-liquid mixture enriched with “target” components is created. The “central” part of the flow is cleared of the “target” components.[32]

In the extraction unit, the gas stream is further divided into two: a gas-liquid stream to the gas-liquid separator and a prepared (marketable) gas stream to the consumer.

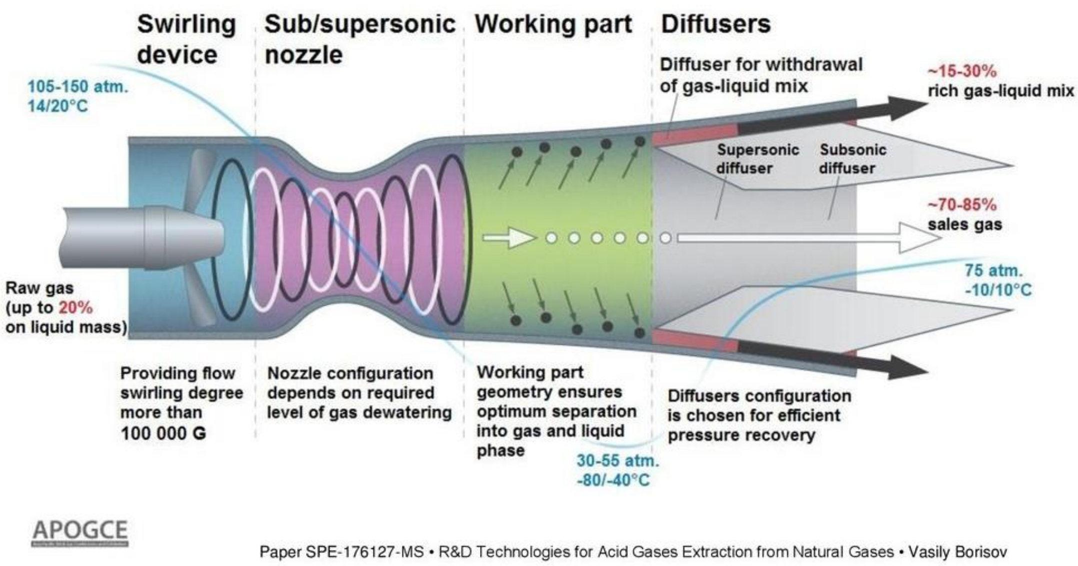

Figure 9: The basic scheme of the Supersonic separator

Using diffusers at the outlet of the working part of the 3S-separator allows converting part of the kinetic energy of the flow into potential energy by deceleration. It provides significantly higher gas pressure at the diffuser outlet than the static gas pressure in the Laval nozzle, where the condensation of the target components occurs (Figure 9).[33]

Among the advantages of 3S technology are the following:
- Small plant dimensions;

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[30] S.Z. Imaev, E.A. Nikolaev. The study of gas flow in the channel of a supersonic separator // Modern Science, № 2 (10), 2012. p. 290-294
[31] Khetagurov V.A., Slugin P.P., Vorontsov M.A., Kubanov A.N. Experience and prospects for the use of turbo expander units at the field technological facilities of the Russian gas industry // Gas Industry. 2018. №11 (777). Pp. 14-22
[32] A.M. Shammazov, A.G. Gumerov, L.I. Bykov, A.V. Kolchin, V.R. Gallyamov. Geometry optimization of 3S-separator as a reducing device for low-temperature technology // Problemy sbora, podgotovki I transporta nefti I nefteproduktov. 2019. Vol. 122(6). P. 133-146
[33] S.Z. Imaev, E.V. Voitenkov, E.A. Nikolaev. Possibilities of using supersonic gas processing technologies in subsea production complexes // Oil and Gas. Innovations, № 9 (176), 2013. p. 52-57

- no moving parts;
- lower capital and operating costs;
- higher efficiency, but for a reduced period compared to the classical variant;
- Possibility to provide enhanced recovery of propane-butane and ethane;
- Effective recovery of CO_2 and H_2S from acidic natural gases;
- Extension of the no-compression period of the fields by ensuring high efficiency at low gas pressure drops in the plant;
- environmental safety.

The application of 3S technology has significant economic advantages: Capex savings achieved, as compared to previous-generation technologies, ranging from 30 to 70% depending on currently used equipment and characteristics of treated gas; possibility to develop gas and oil fields with high content of sour components, which are difficult or impossible to develop using existing technologies; minimum operating costs.[34]

1.4.3. Vortex tubes

The principle of operation of the vortex tube is based on the vortex effect. The essence of the vortex effect is to reduce the temperature in the central layers of the swirling gas flow (free vortex) and increase the temperature in the peripheral layers. With an appropriate device design, the gas vortex can be divided into two flows: one with reduced and increased temperature.

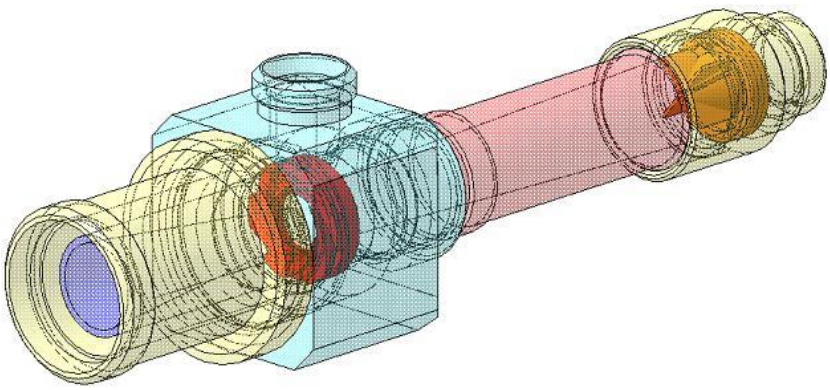

Figure 10: Vortex tube drawing (source: chkz-yugson.ru)

Heating and cooling systems based on steam-compression chillers are quite common today.

Vortex tube systems have several advantages over chillers:

- The main advantage of vortex tube systems is that there are no refrigerants or heat transfer fluids;

[34] V.E. Borisov, G.A. Tarasov. Supersonic technologies of natural and associated gas treatment // Oil and Gas, № 1 (103), 2018. p. 136-151

- The vortex tube is of simple construction, which reduces the time required for fabrication, installation, and maintenance;
- No moving parts in the vortex tube design considerably increase the reliability of the overall refrigeration and heating system;
- The convenience of the layout. All the equipment is sufficiently compact and lightweight. The whole system consists of separate units, which can be placed in different places and any position;
- Possibility to cool and heat gas with one system;
- A vortex tube is a low-inertia unit. The vortex tube has only a few seconds to reach nominal operation after supplying compressed gas to the inlet. This fact makes it possible to regulate the thermal operation of any system with high precision and almost instantaneously.

Currently, all substances used as a refrigerant in refrigeration machines have an increased fluidity. For example, the normalized leakage rate of Freon per year is around 6-8 % of the total. Minor connection faults, micro-cracks, and significant fluctuations in ambient temperature lead to additional refrigerant leaks. Leakages of these substances have a significant impact on human health and the planet's ecology. The substances used as coolants are generally poisonous and as a result, also constitute a danger to health.

The range of possible applications for vortex units is quite broad and includes almost all industry and national economies. Vortex tube-based devices leave practically no alternative if a compressed gas source is already installed. Some applications for vortex tubes are listed below.

- Industrial refrigeration and heating systems;
- Refrigeration systems for the food industry and trade;
- Space heating and air conditioning;
- Heat pumps;
- Cryotechnology.

Although vortex tubes were developed initially as refrigeration and heating devices, they can also be used in several other applications, such as cleaning liquids and gases, separating liquids and gases into fractions, etc.

Chapter 2. Selection and description of a basic LNG production scheme

2.1. Modern level of development of low-tonnage liquefied natural gas production technologies (Methods)

2.1.1. High-pressure systems with throttling and pre-cooling

One of the first plants in the Russian Federation to produce small volume LNG with 1 ton/hour capacity was built at CNG filling station No 8 in Peterhof, Leningrad region. This plant is a classical high-pressure throttle cycle with pre-cooling using a freon circuit (Linde Hampson cycle) (Figure 11). This technology allows for the liquefaction of incoming natural gas by about 40 %. Later this technology was modernized and used to construct natural gas liquefaction plants with the capacity of 1 t/h in Kingisepp, Leningrad region, and two plants were built in Kaliningrad with the capacity of 1.5 t/h.[35][36]

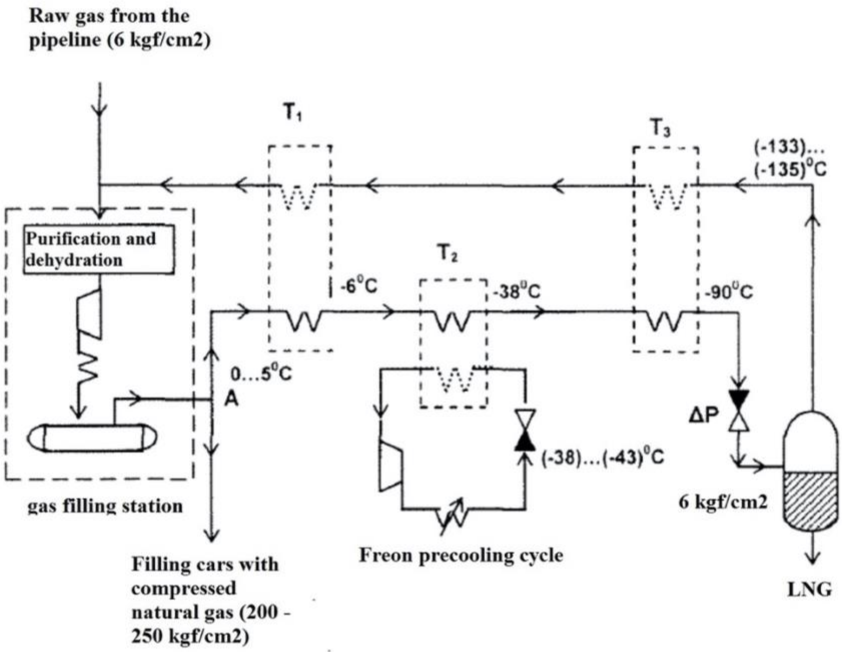

Figure 11: High-Pressure Throttle Cycle Scheme with Pre-Cooling by the Refrigerator at 233K[37]

[35] Bezrukov, K.V. Box installation of natural gas liquefaction with capacity of 1,5 t/h / K.V. Bezrukov, A.L. Dovbish, V.A. Peredelsky // Technical gases. 2008. No 3. C. 67-70.
[36] GOST R 55892 - 2013 Facilities for small-scale production and consumption of liquefied natural gas. General technical requirements. Moscow: Standartinform, 2014. - 44 c.
[37] Pat. 5036671 A United States, IPC5 F 25 J 3/06. Method of liquefying natural gas [Text] / Nelson W.L., Garcia L.; applicant and patentee Liquid Air Engineering Company. Filed 06.02.90; pub. 06.08.91; No 07/475,908 (USA). 7 p.

High-pressure pre-cooling cycles for liquefying natural gas were further improved. The main technological innovation of this scheme was that the throttle valve was replaced by an ejector, which used the kinetic energy of gas expansion. To remove non-condensable impurities such as hydrogen, helium, and nitrogen, the return line is divided into two unequal parts: most of the return line flow is fed through the heat exchanger to the compressor discharge line, and a smaller part is fed to the separator cooling and throttling with subsequent separation of the gas fraction containing mainly nitrogen. A high-pressure throttle cycle variant with 233 K pre-cooling with two ejectors is shown in Figure 12.[38]

The desire to reduce investment in significant equipment has led to the development of a system that takes advantage of the joint location of the LNG plant using high pressure, pre-cooled circuit (Figure 13). In this case, the scheme does not use expensive cooling equipment, and the liquefaction factor is 35-40 %.

The Canadian company Cleanair Combustion Systems has created a condensation technology called Anker Gram Liquefier. This technology is a high-pressure natural gas pre-cooling system using a double throttle circuit with a partial return to the compressor inlet at the appropriate compression stage. The peculiarity of this unit is the integration of heat and mass flows of the natural gas treatment system and the compressor drive (Figure 14). This technology was created to produce liquefied natural gas for its further use as motor fuel.[39]

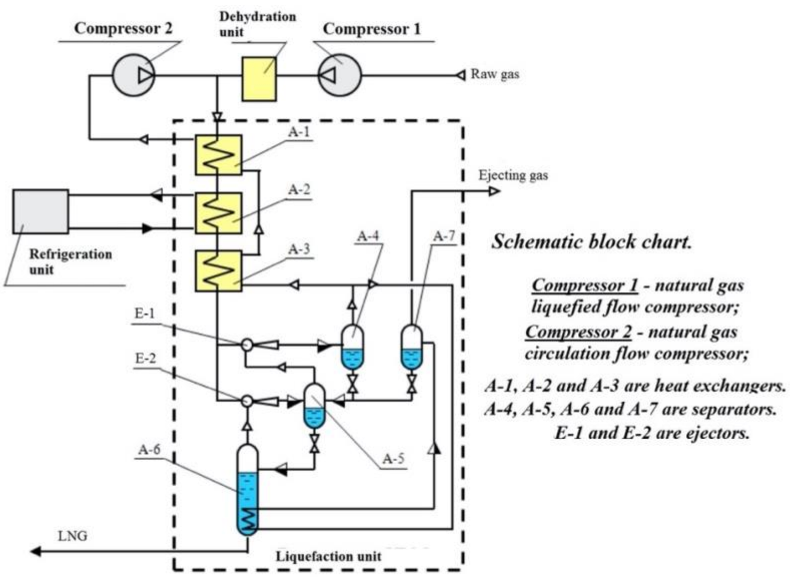

Figure 12: High-Pressure Throttle Cycle Diagram with Pre-Cooling by Refrigeration Machine at 233 K with Two Throttle Ejectors

[38] Krakovsky, B.D. Modern technologies of natural gas liquefaction in installations of small capacity / B.D. Krakovsky // Technical Gases. 2008. No 6. pp. 26-30.
[39] Khodorkov, I.L. Low-tonnage LNG plant on the basis of the combined complex / I.L. Khodorkov // AGZK+AT. 2004. No 3(15). pp. 50- 51.

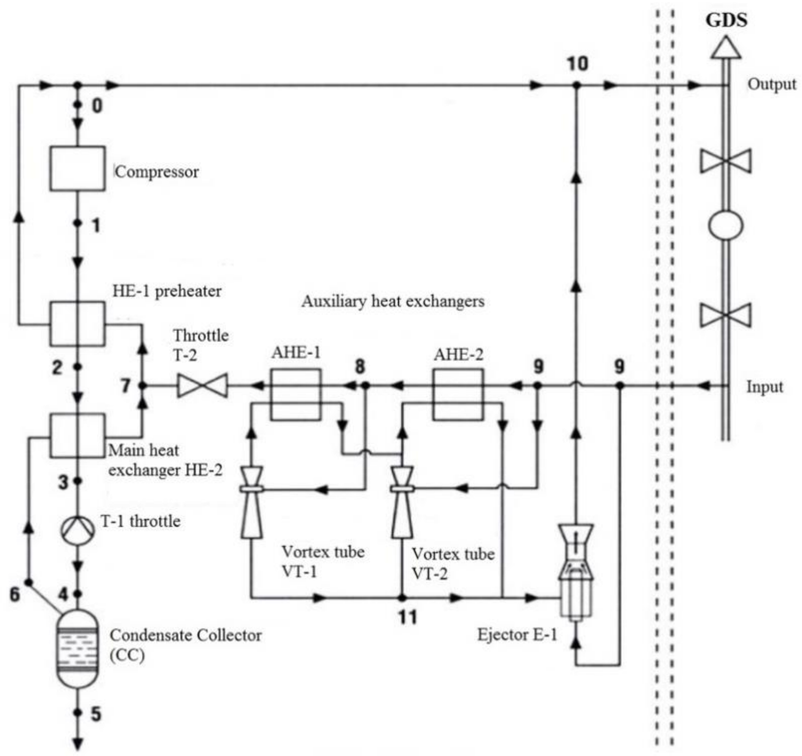

Figure 13: High-Pressure Cycle Diagram with Pre-Cooling by Rank Hills Vortex Tubes[40]

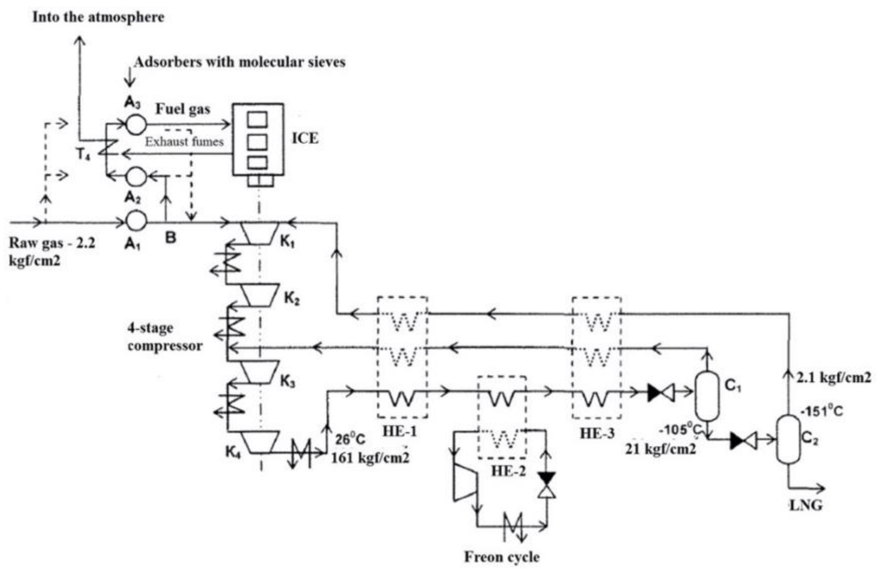

Figure 14: High-pressure cycle diagram with double throttling, pre-cooling, and use of exhaust gas heat from compressor drive for the regeneration process of the natural gas cleaning system[41]

40 Pat. 7594414 B2 United States, IPC F 25 J 1/00. Apparatus for the liquefaction of natural gas and methods relating to same [Text] / Wilding B.M. [et al.]; applicant and patentee Battelle Energy Alliance, Llc. Filed 05.05.06; pub. 29.09.09; prior. 28.09.06, No 11/381,904 (USA). 53 p.
41 Pat. 4033735 A United States, IPC2 F 25 J 1/00. Single mixed refrigerant, closed loop process for liquefying natural gas [Text] / Swenson L.K.; applicant and patentee J. F. Pritchard And Company. Filed 08.11.76; pub. 05.07.77; No 05/739,793 (USA). 12 p.

The desire to reduce the cost of the primary process equipment has led Linde to develop a scheme of the new sample (Figure 14), in which pre-cooling is organized by regulating a direct high-pressure flow of 200 bar to a pressure of 40 and 6 bar with the subsequent return of expanded gas to the appropriate stages of the compressor.[42]

The rest of the flow is throttled to atmospheric pressure and partially liquefied after passing the third cooling stage. This flow is then separated and the gas phase passing through the regenerative heat exchangers is returned to the first stage of the compressor. This technology was used in Stuttgart (Germany).

2.1.2. Medium pressure settings

Lightweight LNG units operating in a medium pressure cycle can be classified according to the type of medium used for cooling.

Therefore, two types of equipment are considered below: equipment with external nitrogen cooling and equipment with a cooling circuit using raw gas. In both cases, the turboexpander is a source of cold.

2.1.2.1. Medium pressure units with a cooling circuit on raw gas

American company Chicago Bridge & Iron Co patented the LNG-Pro technology. The technology is designed to liquefy gas in remote wells with low flow rates. It is a classic Claude cycle with a turboexpander, throttle valve and precooling to 233 K by using a propane cycle refrigeration machine. The claimed characteristics of this technology are liquefaction factor of 0.67, value of specific power consumption of 0.305-0.324 kWh/kg LNG at inlet pressure of raw natural gas of 65-70 bar.[43][44]

2.1.2.2. External nitrogen-cooled systems

German company Tractebel Gas Engineering (TGE) serially produces a reverse condensation plant for liquefied natural gas tankers with a reversal cycle for the external nitrogen cooling circuit and a simple throttle cycle for the natural gas flow. This technology is also used in power plants with peak winter loads in Belgium (capacity 10 t/h) and England (8 t/h).

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[42] Arkharov A.M. Entropic-statistical analysis of low tonnage plants for liquefaction of natural gas with methane content of 92 % / A.M. Arkharov // Chemical and Oil and Gas Engineering. 2012. No 4. - C. 19-27.
[43] Pat.20090100844 A1 United States, IPC F 25 J 1/02, F 17 C 13/08, F 17 C 6/00. Apparatus and method for controlling temperature in a boil-off gas [Text] / Rummelhoff C.J.; Hamworthy Gas Systems As. Filed 11.11.04; pub. 23.04.09; prior. 13.11.03, No PCT/NO2004/000342. 9 p.
[44] Wunsch, A. Zum Stand der Entwicklung von gasdynamischen Druckwellennaschinen fur die Aufladung von Dieselmotoren / A. Wunsch // Brown Boveri Mitteilungen. 1968. No8, Vol. 55. P. 440-447.

The external nitrogen cooling scheme is widely used by one of the world market leaders in the cryogenic industry, the American company APCI (Air Products and Chemicals Inc.). In the APCI scheme, external nitrogen cooling is provided by the Claude cycle: the nitrogen flow is also divided into expander and throttle flows (Figure 15).

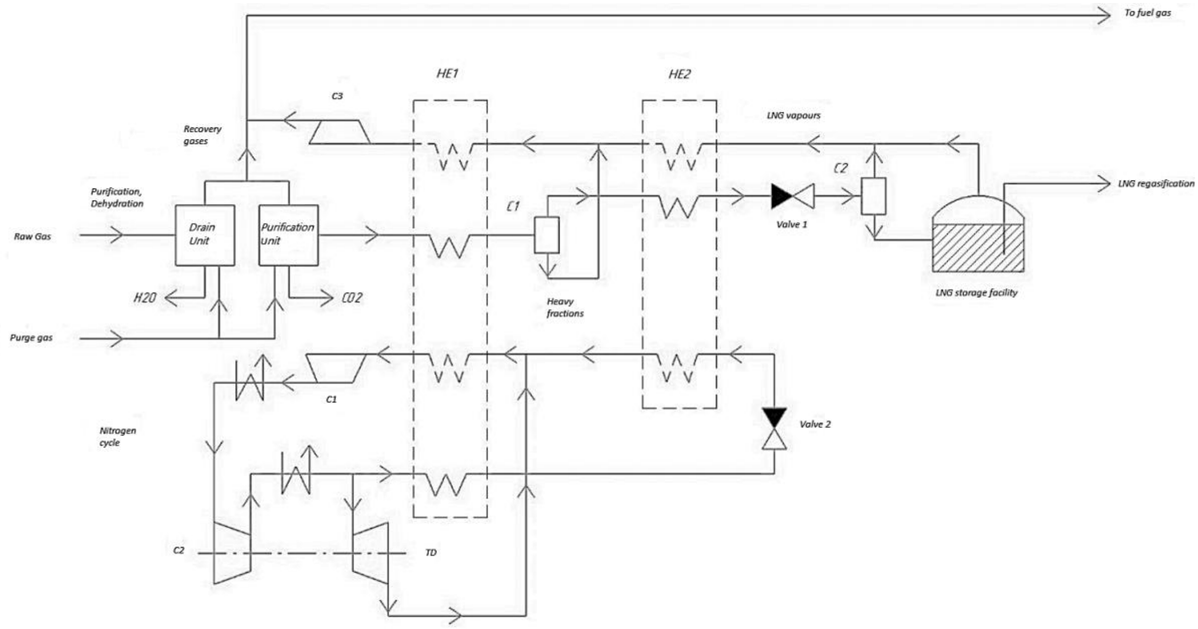

Figure 15: Cycle diagram with external nitrogen cooling circuit, nitrogen expansion in turboexpander, throttling, and separation of heavy hydrocarbon fraction in feed gas circuit

The scheme of the liquefaction process is based on an external nitrogen circuit with a cascaded arrangement of dynamic expanders[45] (Collins circuit), patented by Kosmodin LLC under the name "Nitrogen Expansion Cycle." It is used to provide filling stations selling liquefied natural gas as motor fuel. After dehydration and absorption of carbon dioxide, the raw gas is cooled using a typical freon cycle refrigeration machine. After the first heat exchanger, only the liquid hydrocarbon phase is separated. The natural gas is condensed and subcooled in the second heat exchanger. After expansion in the throttle, unit LNG is directed to the tank farm.[46]

[45] Barnes, J.A. The pressure exchanger / J.A. Barnes, D.B. Spalding // The oil Engine and Gas Turbine. 1958. No 294, Vol. 25. 364-366.
[46] Berchtold, M. A new small power output gas turbine concept / M. Berchtold, T.W. Lutz // ASME Paper. 1974. – No 74-GT-111. 48 p.

2.1.2.3. Blended refrigerant-based cycles

At the congress of the International Institute for Refrigeration in Copenhagen, a cycle for liquefying natural gas using a specially synthesized working mixture (refrigerant) was proposed. The working mixture is a mixture of nitrogen and hydrocarbons. Use of the mixed refrigerant allows arranging multiple throttling cycles at different temperature regimes, which is regulated by the content of hydrocarbon components in the blend.[47]

It should be noted that all world market leaders in the cryogenic industry have unique patented technologies to produce liquefied natural gas, based on the Kleemenko cycle[48] in various variations. One of the first plants designed to produce liquefied natural gas for export was PRICO (Poly Holerant Integrated-cycle operations[49]) - a process of throttling cycle on the use of mixed refrigerant with a specific value of expansion pressure and adjustable refrigerant temperature level.

2.2. Requirements for LNG

According to GOST R 56021-2014 liquefied natural gas must be produced in accordance with the requirements of technical regulations, approved in the prescribed manner at the production plant.[50]

According to this standard, marketable LNG is divided into 3 quality grades: A, B, and C. Indicators are regulated depending on the type of LNG, the main of them are:

- Range of values of Wobbe number under standard conditions;
- The lower heating value at standard conditions;
- The molar fraction of methane;
- The molar fraction of nitrogen;
- The molar fraction of carbon dioxide.

A complete list of the physicochemical quality indicators for LNG is given in Table 1.

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[47] Lanchakov G.A., Kulkov A.N., Siebert G.K. Technological processes of natural gas treatment and methods of equipment calculation. - Moscow: Nedra - Business Center, 2000. - 279 p.: ill. - ISBN 5-8365-0047-9;

[48] A.L. Dovbish, R.V. Darbinyan, 2003 - Efficient plant for natural gas liquefaction with application of 'Klimenko's open cycle'

[49] Silvia Pérez, Rocío Díez, 2014 - Opportunities of monetising natural gas reserves using small to medium scale lng technologies

[50] GOST R 56021 - 2014 Liquefied combustible natural gas. Fuel for internal combustion engines and power installations. Moscow: Standardinform, 2014. 13 p.

However, regasified LNG of grade B must meet the requirements of GOST 27577[51], and LNG of grade B must meet the requirements of GOST 5542, the only exception being odor intensity of this grade.[52]

Table 1: Quality indicators§

Parameter name	Value for the brand			Method of analysis or measurement
	A	B	C	
Component composition, molar fraction, %	The determination is compulsory			According to GOST 31371.1-GOST 31371.7
Range of values for Wobbe number (highest) under standard conditions, MJ/m^3	from 47,2 to 49,2	Not regulated	From 41,2 to 54,5	According to GOST 31369
Lower heating value under standard conditions, MJ/m^3	Not regulated	From 31,8 to 36,8	At least 31.8	According to GOST 31369
Methane molar fraction, %, min.	99,0	80,0	75,0	According to GOST 31371.1-GOST 31371.7
Molar fraction of nitrogen, %, max.	Not regulated	5,0	5,0	According to GOST 31371.1-GOST 31371.7
Mole fraction of carbon dioxide, %, max.	0,005	0,015	0,030	
Molar fraction of oxygen, %, max.	0,020			
Mass concentration of hydrogen sulfide, g/m^3, max.	0,020			According to 8.4 GOST R 56021-2014
Mass concentration of mercaptan sulfur, g/m^3, max.	0,036			According to 8.4 GOST R 56021-2014
Calculated octane number (motor method), at least	Not regulated	105	Not regulated	According to GOST 27577

[51] GOST 27577 – 2000. Compressed natural fuel gas for internal-combustion engines. Specifications
[52] GOST 5542 - 2014 Natural combustible gases for industrial and municipal purposes. Technical conditions. Moscow: Standardinform, 2015. - 9 p.

2.3. Description of the Aspen Hysys software

Aspen HYSYS (or simply HYSYS) is a dynamic simulator, designed for the mathematical simulation of chemical and physical-chemical processes, ranging from the simplest ones implemented in Gas and Gas Condensate Processing Units (GCPU) to the most complex chemical plants in oil refineries. HYSYS can perform many key process calculations in the oil and gas industry including mass balance, energy balance, phase equilibrium, heat transfer, mass transfer, chemical kinetics, and hydraulic calculations. HYSYS is widely used in industry[53] and scientific research for stationary and dynamic simulation of technological processes, design, and optimization of engineered plants.[54]

2.4. Methodology

The objective of this study is modeling a natural gas liquefaction plant with a raw gas flow rate in the range of 5.5 to 30 thousand Nm^3 per hour. The composition of the raw gas[55] is given in Table 2.

Table 2: Composition of raw gas

N_2	CO_2	H_2S	H_2O	CH_4	C_2H_6	C_3H_8	C_4H_{10}	C_{5+}
6,01%	5,56%	0,01%	0,01%	83,29%	3,03%	0,79%	0,32%	1,00%

The modeling of the process flow diagram of a small-scale LNG plant was performed in the Aspen HYSYS software package, based on the Peng-Robinson thermodynamic package, in static mode. However, the following assumptions were made in the calculations:

1) Modeling was carried out without considering heat exchange with the environment. Also, the hydraulic resistance of pipelines was not taken into account;
2) Water content in raw gas was assumed to be in saturation condition at given thermobaric conditions at the inlet to the plant;
3) Water content in gas after drying unit was neglected;
4) Temperature head in heat-exchangers was assumed: for heat-exchangers “gas-gas” 10 °C, for heat-exchangers “gas-condensate” 15 °C.[56]

Due to imperfect separation and heat-exchange equipment, another correction was accepted in the calculation scheme to reduce commercial LNG temperature by 3-7 °C below the minimum storage range limit by GOST R 56021 2014.

[53] University of Calgary. "Hyprotech: Simulation software for industry", 2014
[54] Wilcox, William R. "HYSYS and UniSim", 2018
[55] Natural gas. Chemical composition. Deposits: URL:http://biofile.ru/geo/3286.html
[56] Zagoruchenko, V.A. Thermal physical properties of gaseous and liquid methane / V.A. Zagoruchenko, A.M. Zhuravlev. M.: Publishing house of the Committee on Standards, Measures, and Measuring Instruments of the USSR Council of Ministers, 1969. 236 p.

The quality requirements for the final LNG were determined in accordance with GOST R 56021-2014. As the raw gas comes mainly from the main gas pipeline, the incoming marketable gas meets the requirements of STO Gazprom 089, 2010.[57]

Variable parameters of the study were:

- the raw gas flow rate in a given range for four values: 5500, 10000, 20000, and 30000 Nm^3/h;
- outlet temperatures from heat exchangers;
- volume distribution in the nitrogen cycle;
- total system nitrogen volume (30,000, 40,000 and 50,000 kg).

2.5. Calculation of operating modes

Calculation of operating modes was performed in the software HYSYS (version 8.8, license 30.0.0.8433) based on the Peng-Robinson thermodynamic package in the static mode. The general view of the calculated scheme of the installation is shown in Figure 16.

The operating modes were investigated according to the following algorithm:

- Range of possible combinations on flow splitter H1;
- Dependence of the amount of gas in the recycle and feed gas on the ratio of nitrogen flows in the cooling system;
- Possible volumes of gas recycled in the liquefaction system and gas taken for own use (feed gas);
- Dependencies of the feed gas mass for own needs on the power requirement of the plant.

The gas inlet temperature to the liquefaction plant was assumed to be 20°C.

[57] STO Gazprom 089-2010. Natural combustible gases supplied and transported by main gas pipelines [Text]. Replacement of OST 51- 40-93; introduced. 08.08.11. м: JSC Gazprom, 2010. 15 p.

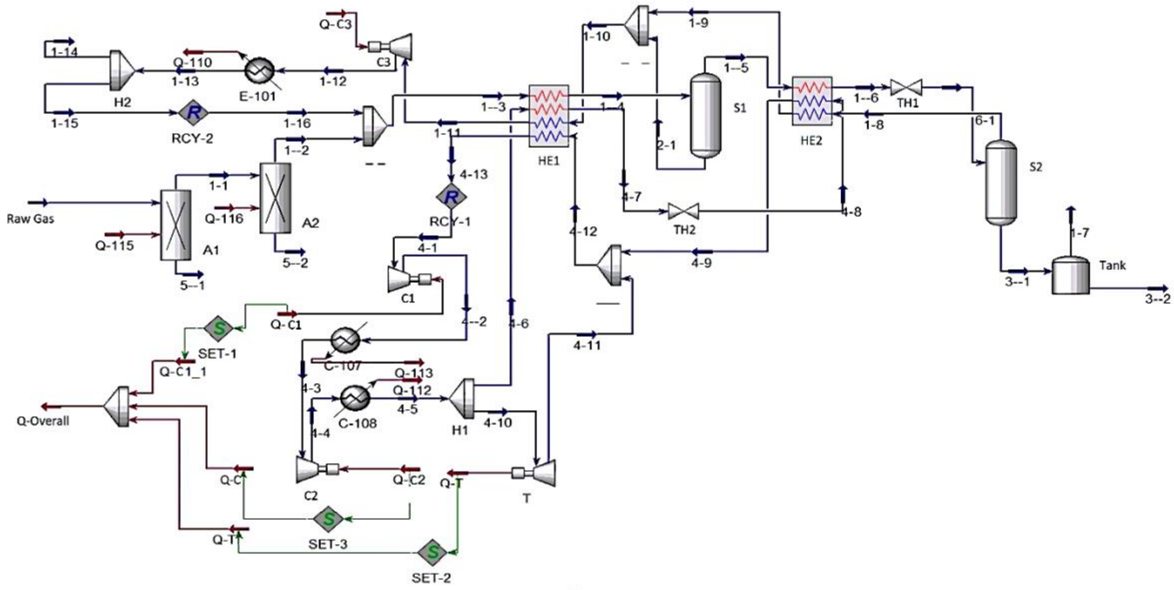
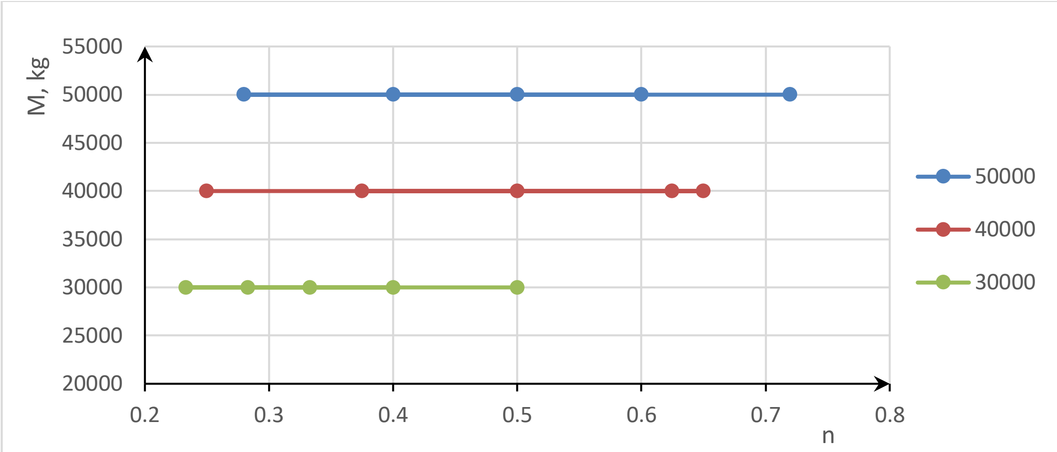


Figure 16: Calculated installation diagram

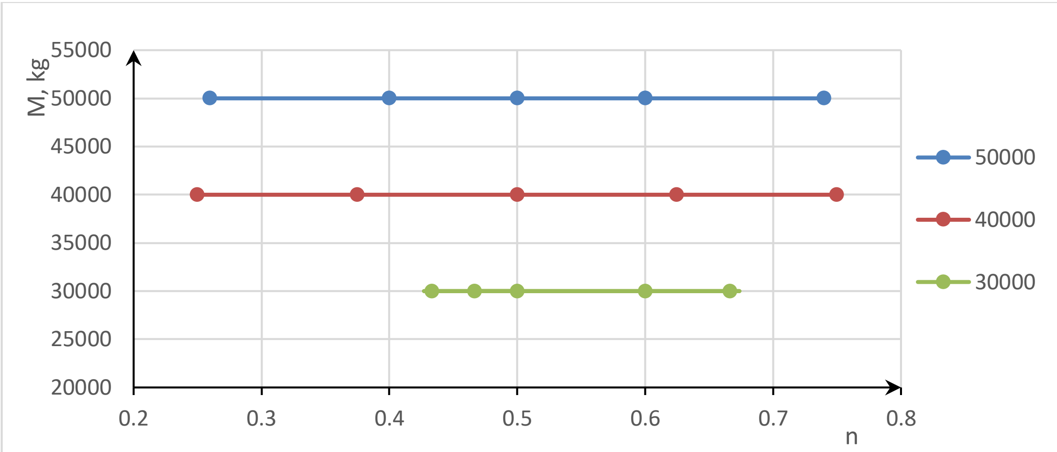
This unit's basic principle of operation is based primarily on the pretreatment of gas from the trunk pipeline. The gas enters the dehydration unit and the purification unit. The dehydration unit is required to remove all moisture from the gas to avoid hydrate formation. The purification unit is required to remove CO2 from the gas. After that, prepared gas enters the heat exchanger, where gas cooling takes place. This cooling is performed by cold energy from the nitrogen cycle. Gas after the first heat exchanger enters the separator. In the separator the heavy hydrocarbon liquid phase is separated from the light gaseous phase. Heavy hydrocarbons escape into the return flow, where it mixes with the tank's gas phase (from storage), forming a recycling flow of the system. After the separator, the gas enters the second heat exchanger, where it is cooled by the cold flow of the nitrogen cycle with a temperature of -183.3 degrees Celsius before entering the throttle. Then partially liquefied gas with high pressure from the main pipeline enters the throttle. The gas is liquefied due to sharp pressure change and, consequently, fast cooling (about -161.5 ºC). Then this LNG enters storage in the tank farm for further transportation.

2.5.1. Range of possible combinations on flow splitter H1.

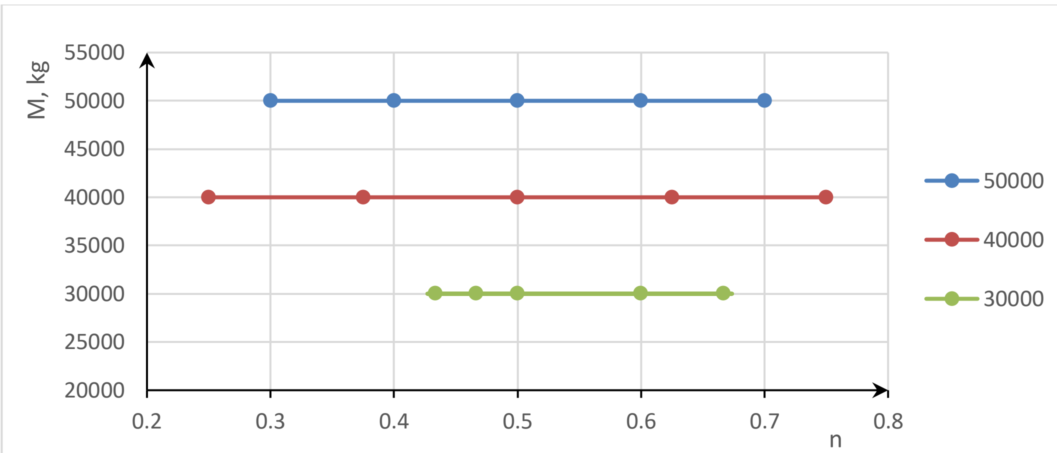
According to the calculated scheme, the range of stable operation of the natural gas liquefaction scheme for three volumes of nitrogen cycle was determined: 30000, 40000, 50000 kg. The calculation results are shown in Figure 17, where axis ‘y’ is Nitrogen cycle capacity (M, kg) and axis ‘x’ is Ratio of two nitrogen cycle streams 4-10 and 4-6 (n).



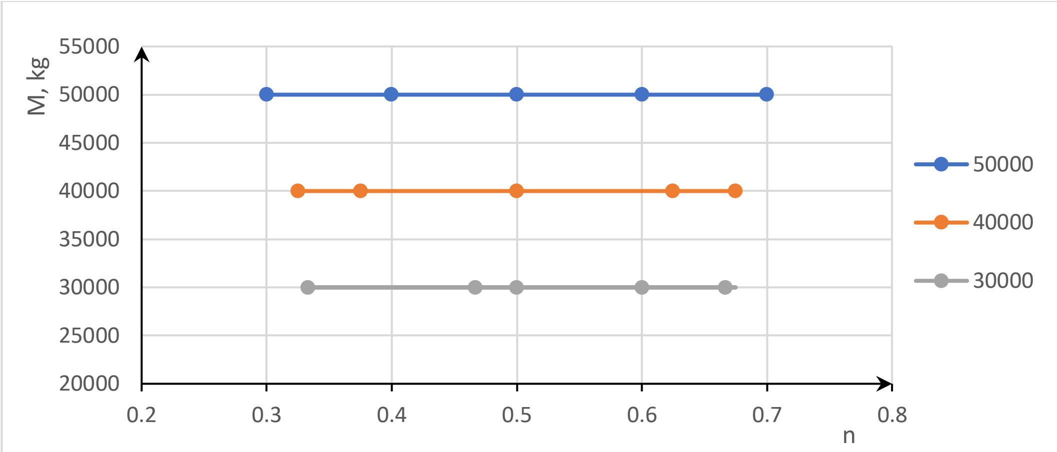
a) Natural gas flow rate 5500 Nm^3/h



b) Natural gas flow rate 10000 Nm^3/h



c) Natural gas flow rate 20000 Nm^3/h



d) Natural gas flow rate 30000 Nm^3/h

Figure 17: Range of possible combinations on flow splitter H1

It was found that the increase in the volume of the nitrogen cycle leads to an increase in the range of stable operation of the technology by 10% cumulatively. In other words, the number of nitrogen flow separation combinations can be increased when the volume of nitrogen is increased, which is directed to the HE1 and TEU. This provides greater flexibility of the system for regulation. However, an increase in the volume of the nitrogen cycle leads to a greater specific quantity of metal of the system and, consequently, higher capital costs. The optimal volume of the nitrogen cycle should be determined by the total capital costs of design, construction, supervision, and commissioning, and technological feasibility.

As can be seen from the graphs, the boundary conditions of the technological scheme under study change depending on the raw gas consumption. By increasing the capacity of the plant, the load on all equipment of the system increases. In addition, the cold potential of the system is not enough to ensure the necessary operating conditions and the quality of liquefied natural gas (or its production). Consequently, there is a decrease in the range of boundary conditions.

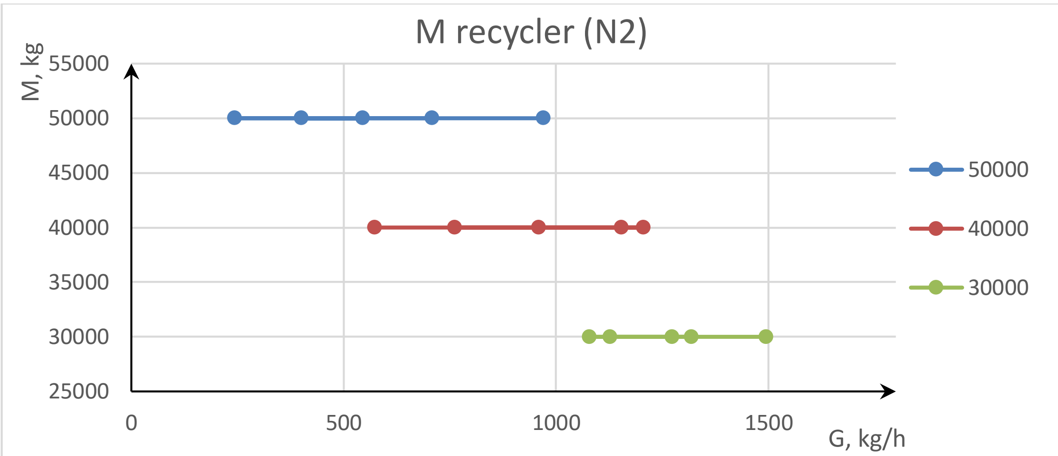
When simulating, the temperature of commercial LNG was set at 2-3 °C. This assumption compensated for the given estimated software error, additionally providing for obtaining a smaller volume of LNG stabilization gases.

However, the possibility of increasing the operating range of the plant for the volume of nitrogen cycle of 40000 m3 has been investigated (Figure 17, b). With the increase in gas consumption, the quality of gas slightly deteriorated, which led to an increase in LNG outlet temperature to minus 162.5 °C. The choice of this scheme mode at the nitrogen cycle of 40000 kg is due to lower economic costs by compressor units compared to the volume of the nitrogen cycle of 50000 kg.

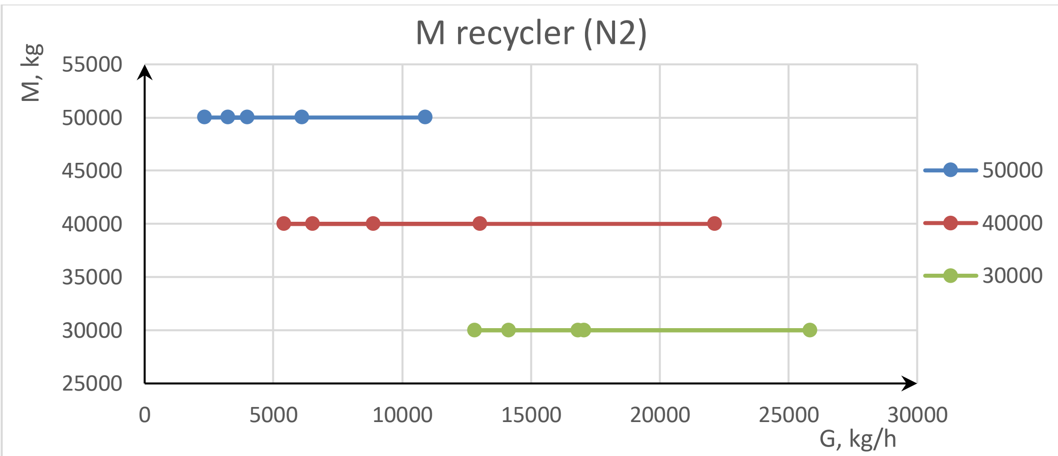
After analyzing the stable operation range of the circuit for the whole range of the plant's productivity, it is reasonable to consider further the operation of the plant at the productivity of 5500 Nm^3.

2.5.2. Possible volumes of recirculating gas in the liquefaction system and gas have been taken away for own needs.

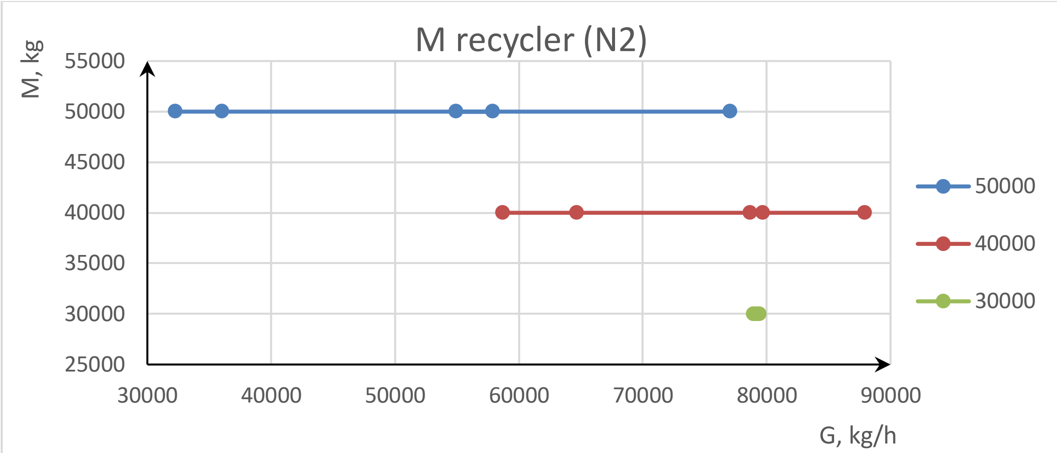
Further, within a certain range of stable operation of the natural gas liquefaction scheme, possible volumes of gas recirculating in the liquefaction system on Figure 18 (Nitrogen cycle capacity, M (kg) and Mass flow of Recycle, G (kg/h) and gas withdrawn for own needs were calculated on Figure 19 (Nitrogen cycle capacity, M (kg) and Mass flow rate of the gas for plant's own needs, G (kg/h). Calculations were made in the studied range of feed gas capacities for three volumes of the nitrogen cycle: 30000, 40000, 50000 m^3.



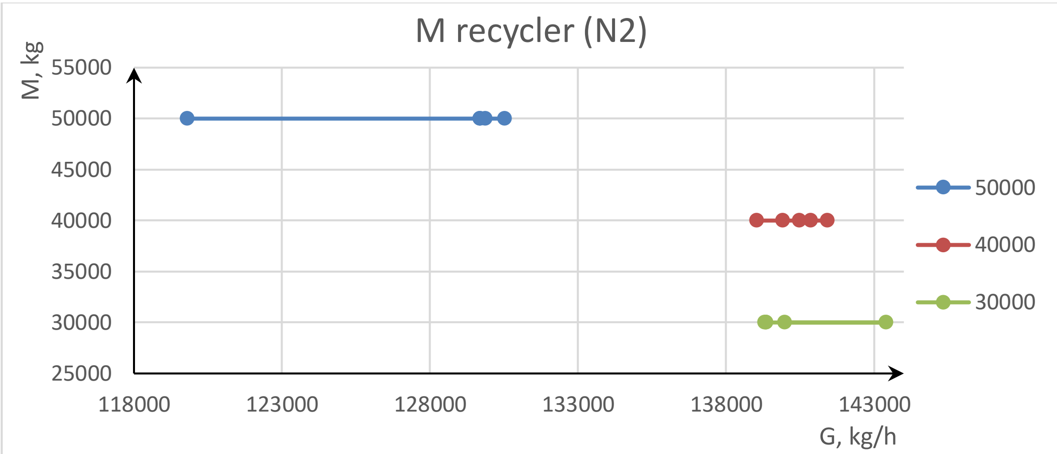
a) Natural gas flow rate 5500 Nm^3/h



b) Natural gas flow rate 10000 Nm^3/h

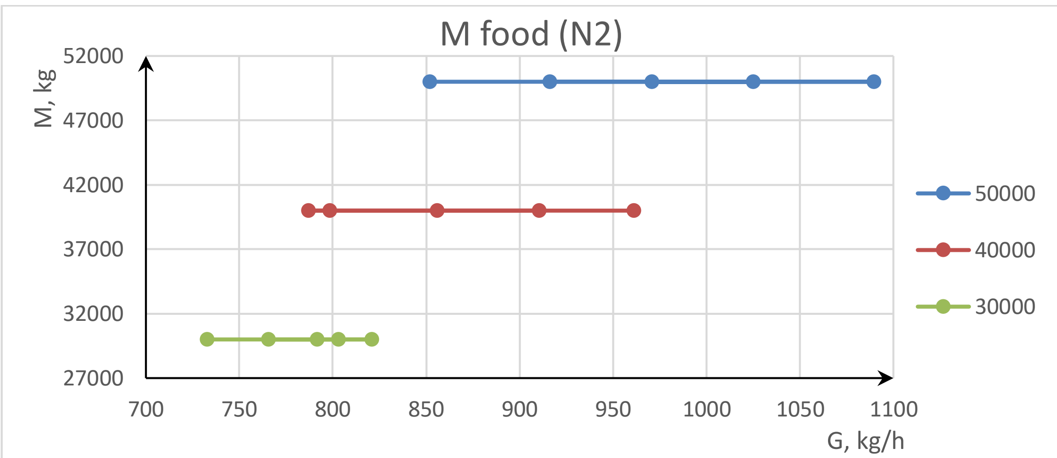


c) Natural gas flow rate 20000 Nm^3/h

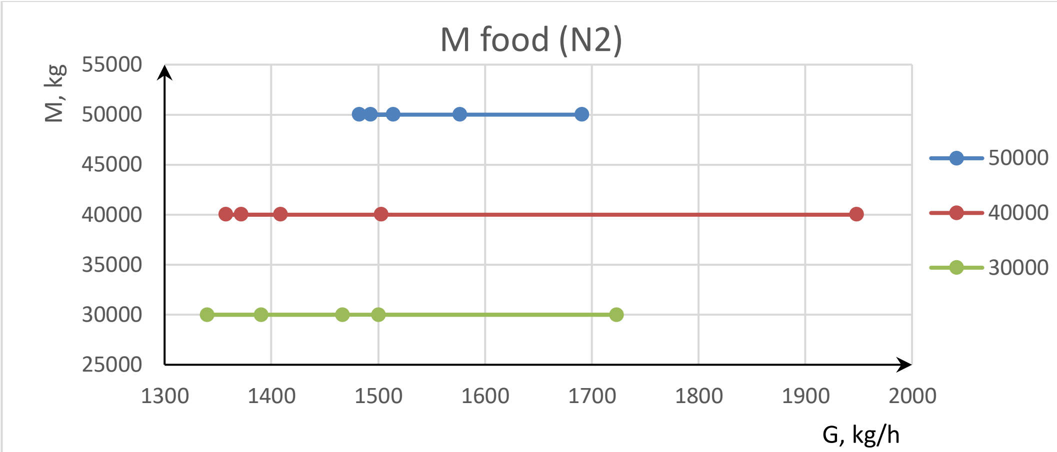


d) Natural gas flow rate 30000 Nm^3/h

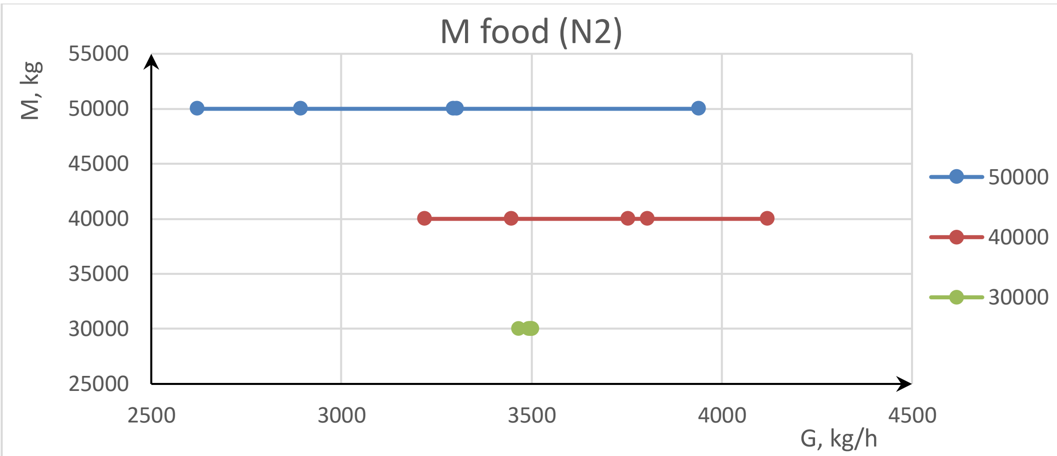
Figure 18: Graphical representation of the volume of recycled gas as a function of the volume of nitrogen in the system



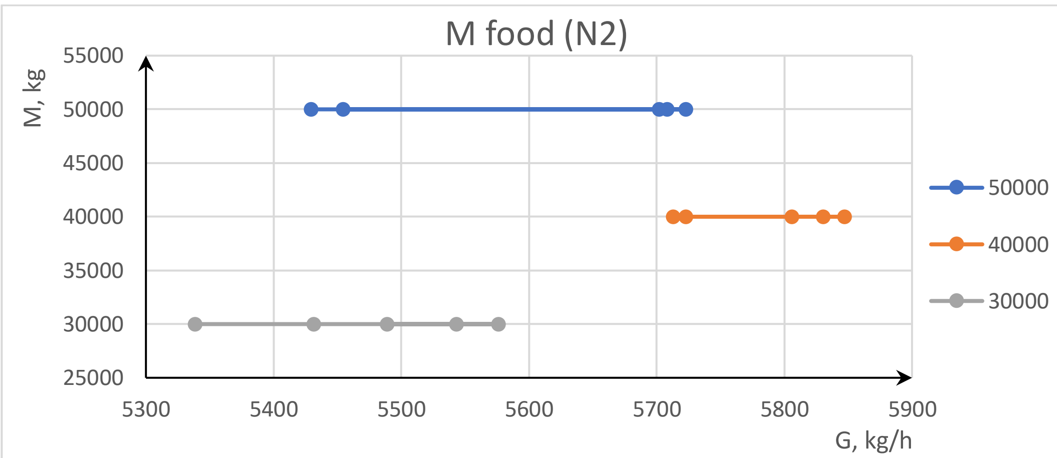
a) Natural gas flow rate 5500 Nm^3/h



b) Natural gas flow rate 10000 Nm^3/h



c) Natural gas flow rate 20000 Nm^3/h



d) Natural gas flow rate 30000 Nm^3/h

Figure 19: Graphical representation of the volume of gas required for the own needs of the plant depending on the volume of nitrogen in the system

Graphical interpretation in Figures 18-19 shows that as the volume of the nitrogen cycle increases, the gas volume of the cycle decreases due to a decrease in the gas temperature after throttling. However, at the same time, the energy costs to ensure the operation of this scheme increase. This is because with increasing volume of the cooling system, at the same time, there is an increase in load on the compressor units of the nitrogen cycle, and therefore an increase in their energy consumption.

Nevertheless, from the diagrams of Figures 18c, d it is seen that there are multiple increases in the volume of gas in the recycle compared with the plant's operating modes at feed gas capacity of 5500 and 10000 Nm^3/h (diagrams a and b). This can be explained by the fact that at the fixed upper limit of the nitrogen cycle volume and high feed gas capacity, the internal cold potential of the nitrogen cycle of the studied scheme is not enough to ensure the operation mode in the range of feed gas consumption of 20000 and 30000 Nm^3/h to ensure the normative quality of LNG.

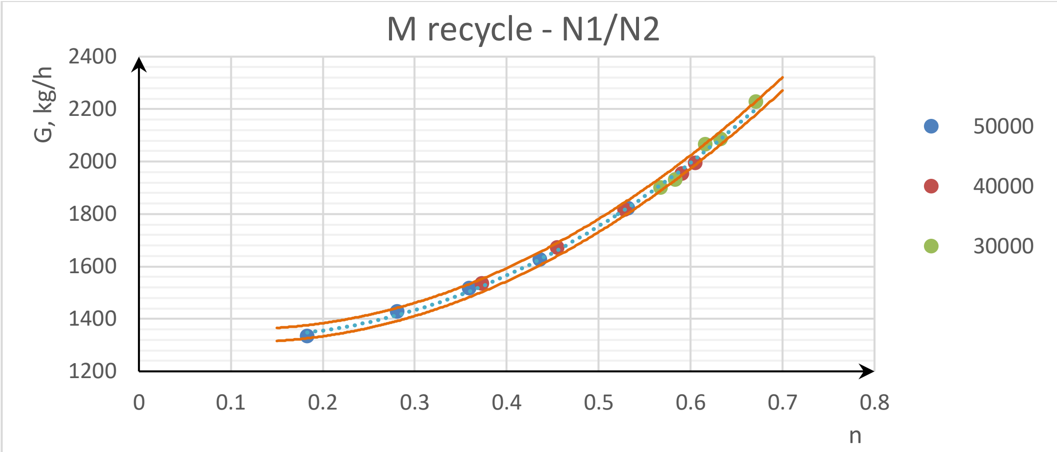
Based on the analysis of graphical interpretations of the gas volume for own needs, shown in Figure 19, it can be concluded that the optimal mode of operation of the unit corresponds to the feed gas capacities of 5500 and 20000 Nm^3/h (Figure 2.4a, c). Other possible modes of plant loading can be implemented but will lead to increased fuel gas consumption.

From the diagrams in Figure 19, one can see an evident dependence, that an increase in the control range leads to an increase in gas consumption for the plant's own needs. In terms of gas usage by the plant itself for operation. This can be explained by the fact that with increasing consumption of raw gas, the volume of recycling of the main line of the system increases, and then the load on the compressor of stabilization gases.

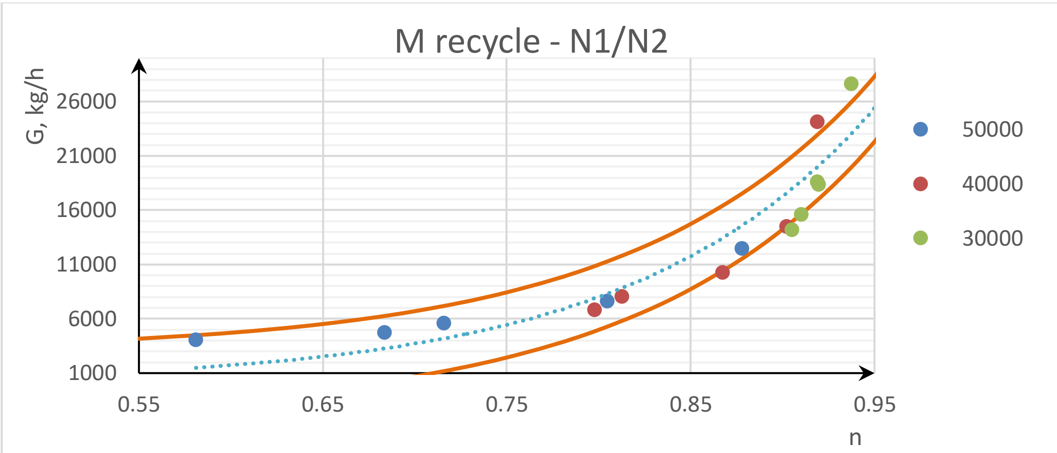
Thus, when the load on the heat-exchange equipment, separators and dynamic equipment of the liquefaction line of the plant is reduced, the energy balances of the system and the load of the dynamic equipment of the nitrogen cycle are redistributed simultaneously. This increases the technical efficiency and the efficiency of the whole complex for natural gas liquefaction. This effect is typical for the operating modes of the plant at the feed gas capacity of 5500 and 20000 Nm^3/h.

2.5.3. Dependence of the amount of gas in the recycle and gas for own needs on the ratio of nitrogen flows in the cooling system

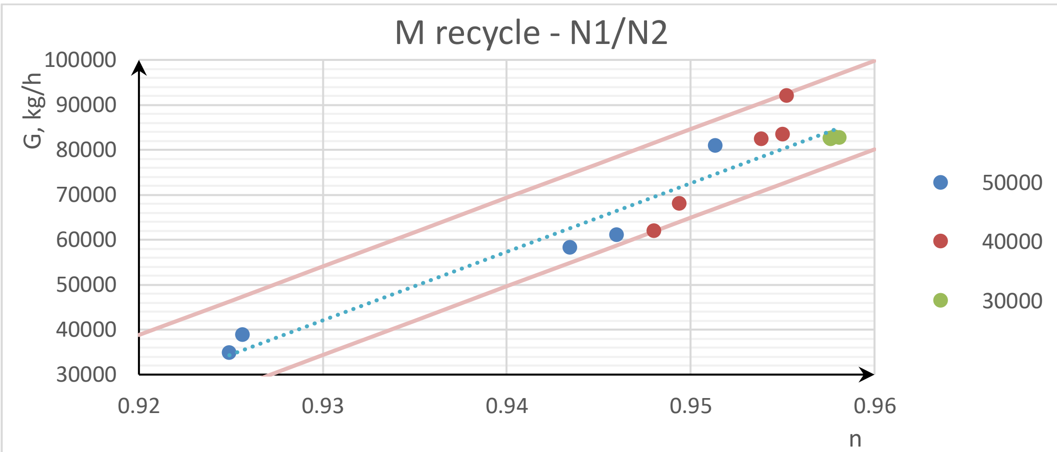
Dependence of the amount of gas in the recycler (Figure 20) and gas for own needs (Figure 21) on the ratio of nitrogen flow in the feed gas cooling system for three volumes of the nitrogen cycle: 30000, 40000, 50000 m^3 are shown in Figures 20-21, where the blue line shows the approximating data for all volumes of the nitrogen cycle, and the orange lines are the error of the data.



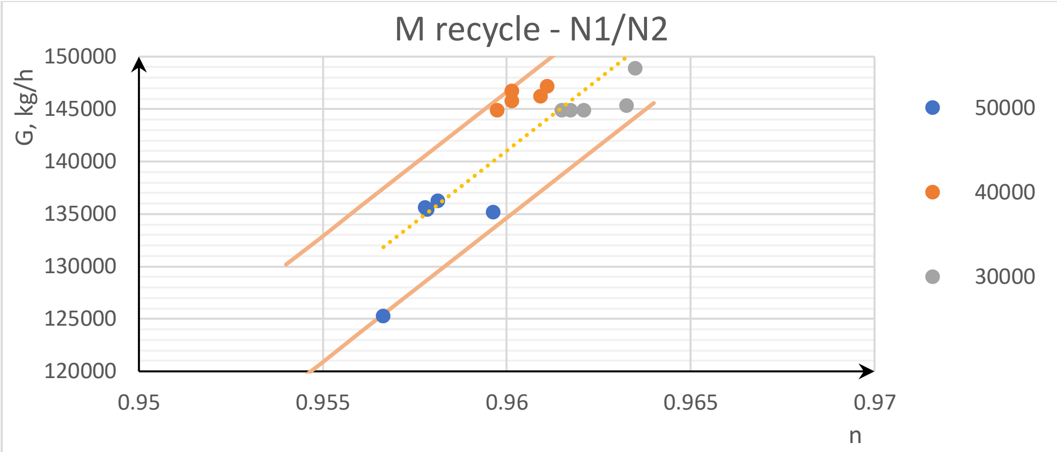
a) Natural gas flow rate 5500 Nm^3/h



b) Natural gas flow rate 10000 Nm^3/h

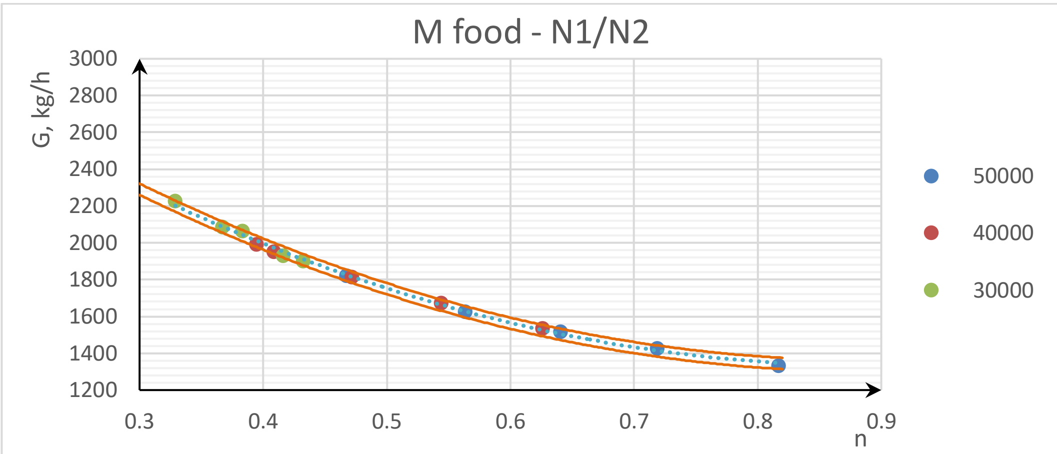


c) Natural gas flow rate 20000 Nm^3/h

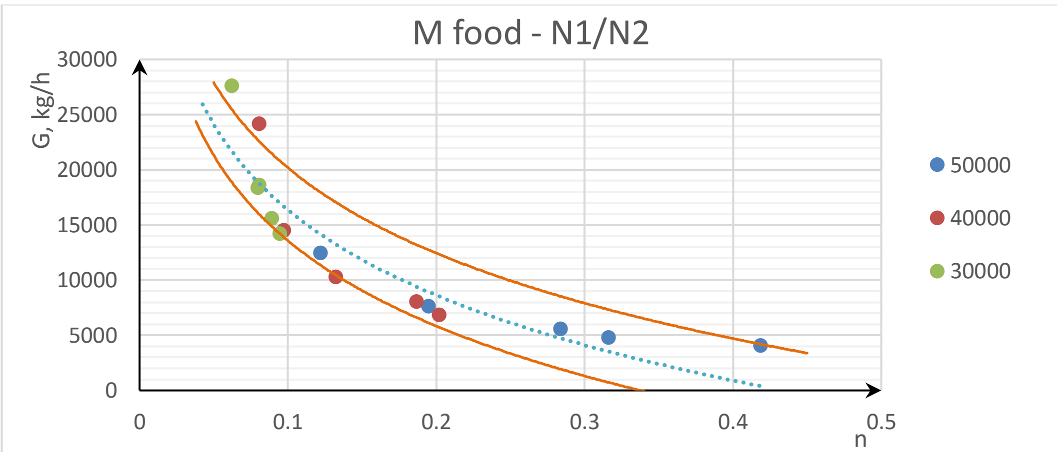


d) Natural gas flow rate 30000 Nm³/h

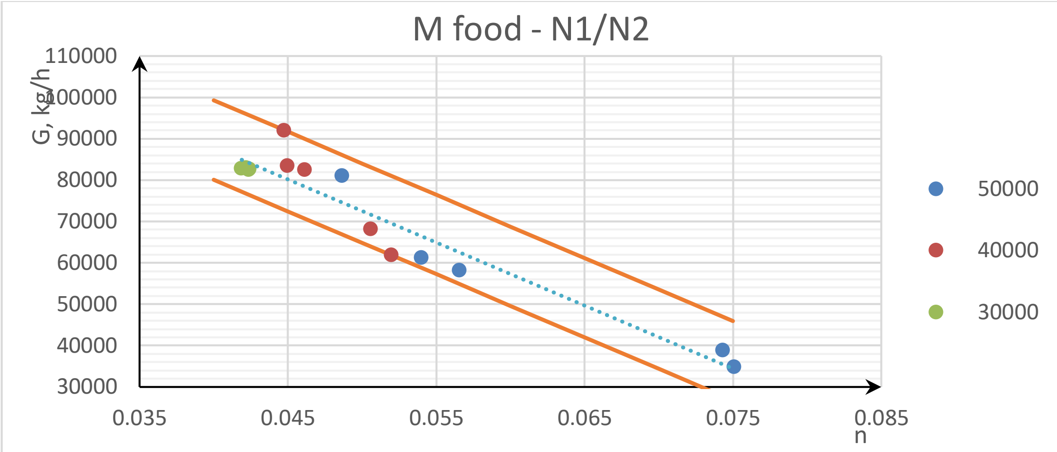
Figure 20: Dependence of gas quantity in the recycling on the ratio of nitrogen flows in the cooling system



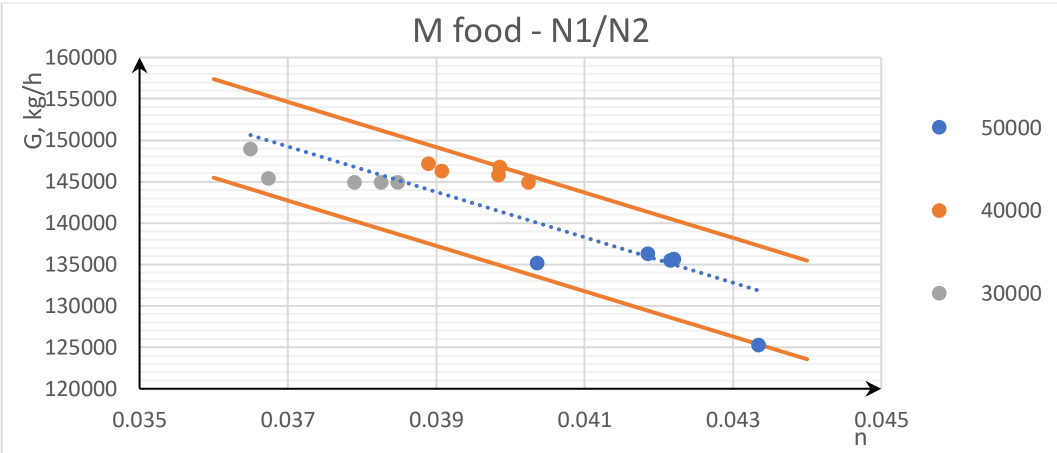
a) Natural gas flow rate 5500 Nm³/h



b) Natural gas flow rate 10000 Nm³/h



c) Natural gas flow rate 20000 Nm^3/h



d) Natural gas flow rate 30000 Nm^3/h

Figure 21: Dependence of gas mass for own needs on the ratio of nitrogen flows in the cooling system

As can be seen from the graph in Figure 20, the dependence for all three volumes of the nitrogen cycle has a clear tendency to decrease gas recycle. This recirculation evaporates from the reservoir and stabilizes the natural gas liquefaction process when the nitrogen cycle volume increases. In this case, there is an inverse symmetrical dependence on the increase of gas, which is necessary for the plant's own needs to ensure the operation of compressor units (Figure 21).

As can be seen from the diagrams in Figures 20-21, there is a different relationship between different dependencies of raw gas consumption. These trends are described by different regularities, as a limiting volume of 50,000 m^3 of nitrogen cycle was initially set. For this reason, as the flow rate increases, so does the divergence of the plotted points in the considered range of the nitrogen cycle volume. Thus, a universal functional dependence cannot be found based on isolated analysis of modes of operation at one capacity value.

This phenomenon is explained by the fact that with increasing consumption of raw gas, a fixed limit of the volume of the nitrogen cycle does not correctly cope with this load and there are dropping points with the high error of its values.

To minimize data errors, individual functional dependencies were applied for each given raw gas plant performance. To describe the universal functional dependence, the surfaces were constructed in the “Statistica” software (Figure 22-23).

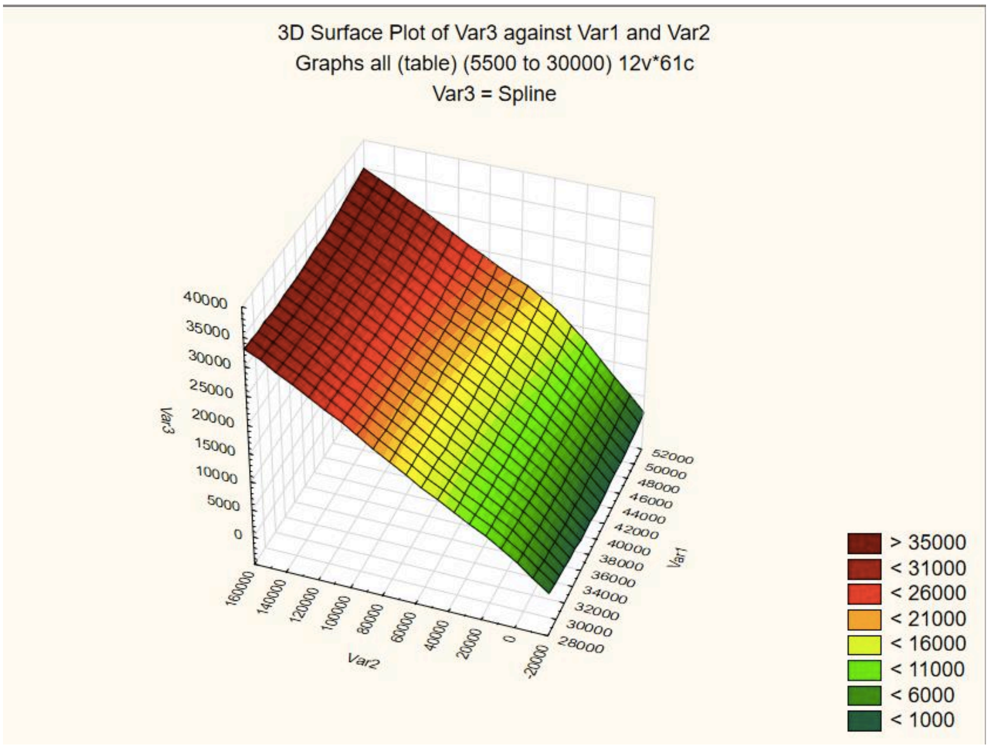


Figure 22: Dependence of gas quantity in the recycling on the ratio of nitrogen flows in the cooling system for the whole range of plant capacities

Table 3: Variables for the graph in Figure 22

Variable	Var 1	Var 2	Var 3
Name	Capacity of nitrogen cycle, kg	Gas volume in recycle, m^3	Natural gas flow rate, Nm^3/h

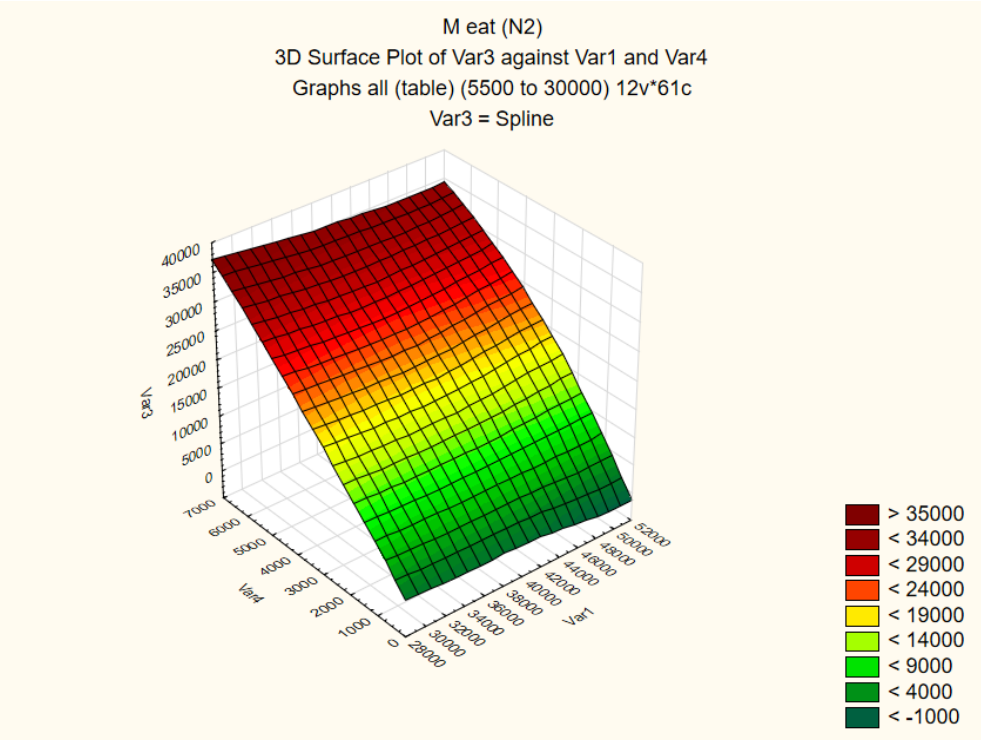


Figure 23: Dependence of gas mass on own needs on the ratio of nitrogen flows in the cooling system for the whole range of the plant productivity

Table 4: Variables for the graph in Figure 23

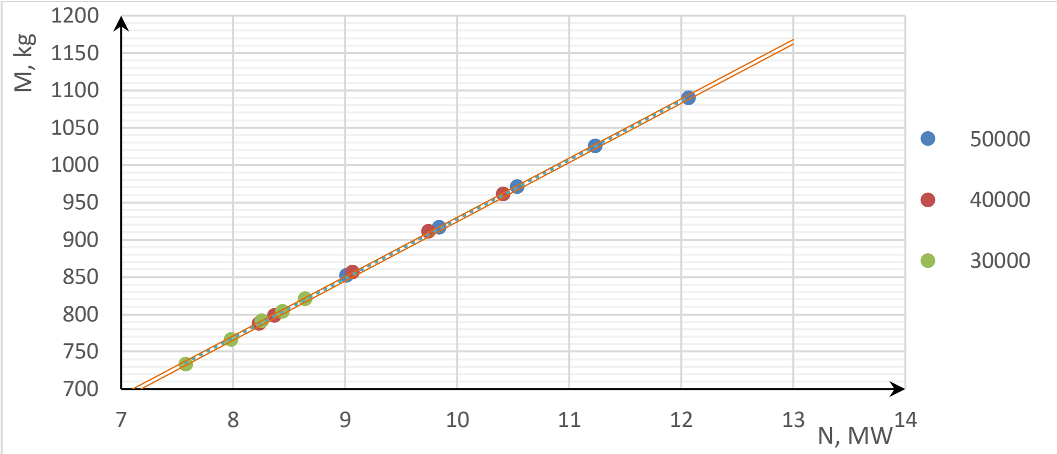
Variable	Var 1	Var 3	Var 4
Name	Capacity of nitrogen cycle, kg	Natural gas flow rate, Nm^3/h	Gas for own needs of the plant, m^3

Based on the analysis of data on the dependences of gas mass for own needs, as well as gas in the recycle on the ratio of nitrogen flows in the cooling system for the whole range of plant performance and therefore it is reasonable to further consider the operation of the plant at the performance 5500 Nm^3.

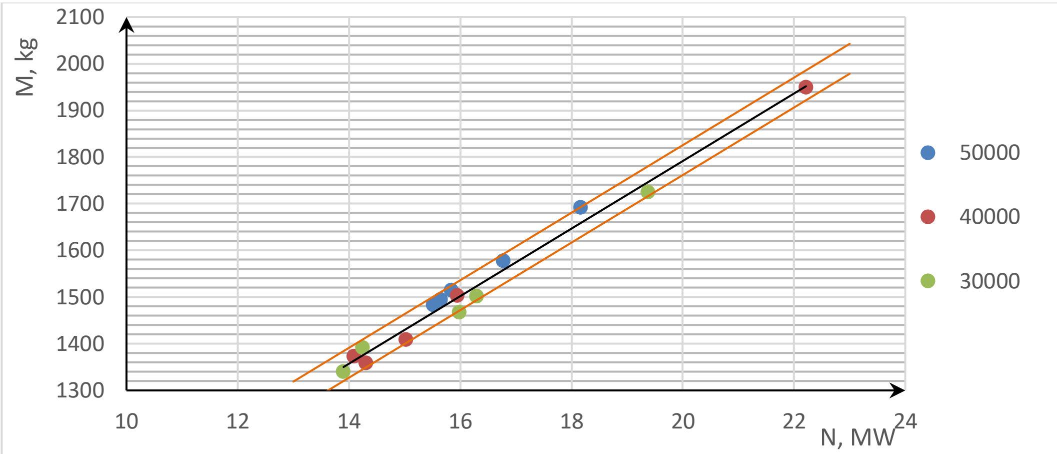
In addition, the intersection of different volume ranges at the selected feed gas capacity provides the possibility of using different variations of the nitrogen cycle volume, which allows greater flexibility in regulation and optimization of the plant.

2.5.4. Dependence of the mass of gas for own needs on the plant capacity

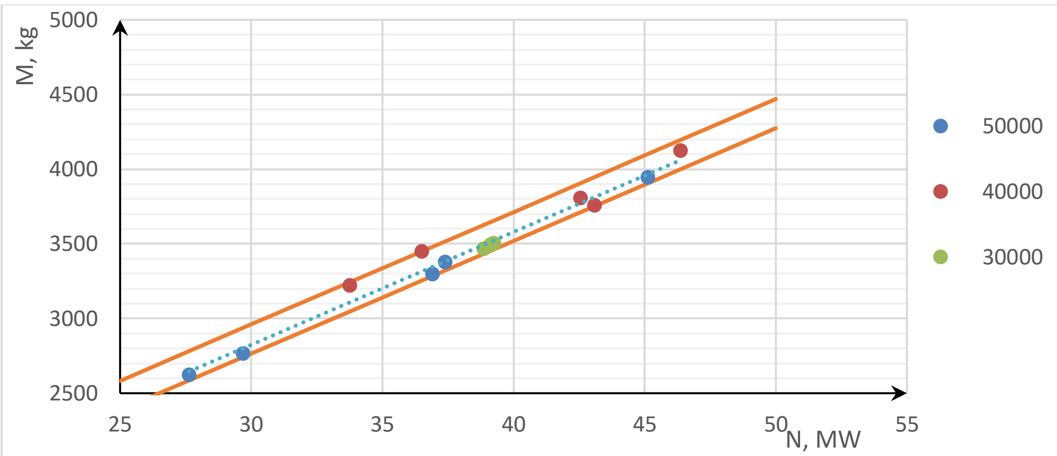
Figure 24 shows the dependence of the amount of gas required for own needs on the power consumption of the studied natural gas liquefaction unit for three volumes of the nitrogen cycle: 30000, 40000, 50000 m^3 for the whole range of feed gas capacities.



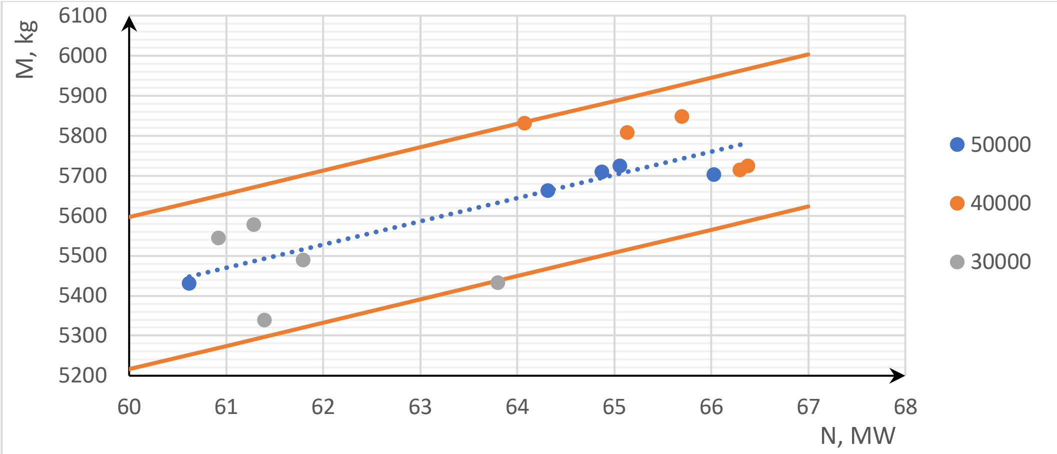
a) Natural gas flow rate 5500 Nm^3/h



b) Natural gas flow rate 10000 Nm^3/h



c) Natural gas flow rate 20000 Nm^3/h



d) Natural gas flow rate 30000 Nm^3/h

Figure 24: Private dependencies of gas mass for own needs on the plant power consumption for different plant capacity options

As can be seen from the graph, there is an expected linear dependence of the increase in the amount of gas to cover own needs with the increase in the power consumption of dynamic equipment. It should be noted that there is a general linear dependence of fuel gas volume on the total power consumption of equipment for different volumes of the nitrogen cycle.
With the increase in the raw gas capacity of the unit, there is a proportional increase in the total power consumption of compressor units, which leads to a corresponding increase in gas withdrawal for own needs as a fuel.

However, with increasing consumption, there is an increase in the error between the calculated values of fuel gas and determined by the approximated average, which is well illustrated in Figure 24.

As private regularities for the isolated values of the plant's feed gas performance are different, the universal functional dependence was determined, and its graphic interpretation was built (Figure 25).

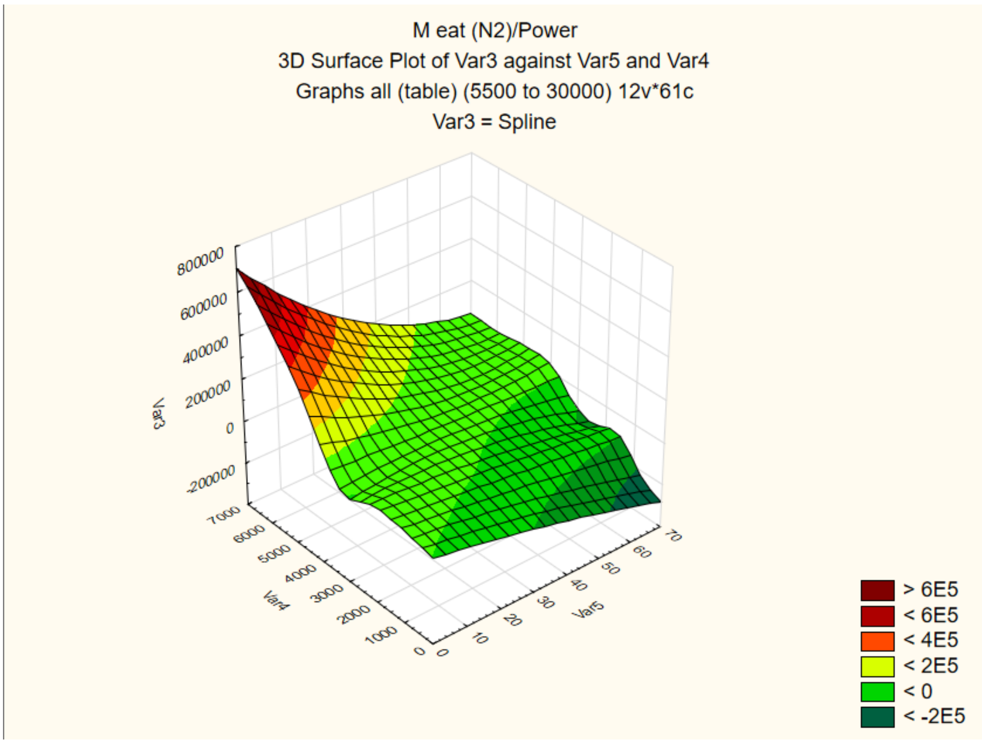


Figure 25: Universal dependence of the mass of gas for own needs on the required capacity of the plant for different variants of the plant productivity

Table 5: Variables for the graph in Figure 25

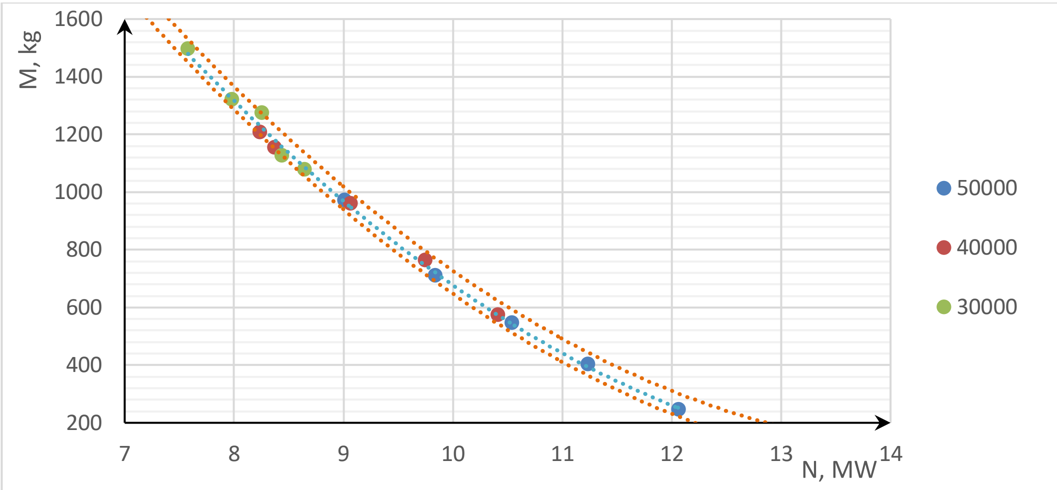
Variable	Var 3	Var 4	Var 5
Name	Gas volume in the system, m^3	Gas for own needs of the plant, m^3	Power consumptions of the plant, MW

Considering the smallest error in the partial approximation function and the lowest values of the required capacities it is recommended only mode of operation at raw gas performance of 5500 Nm3/h.

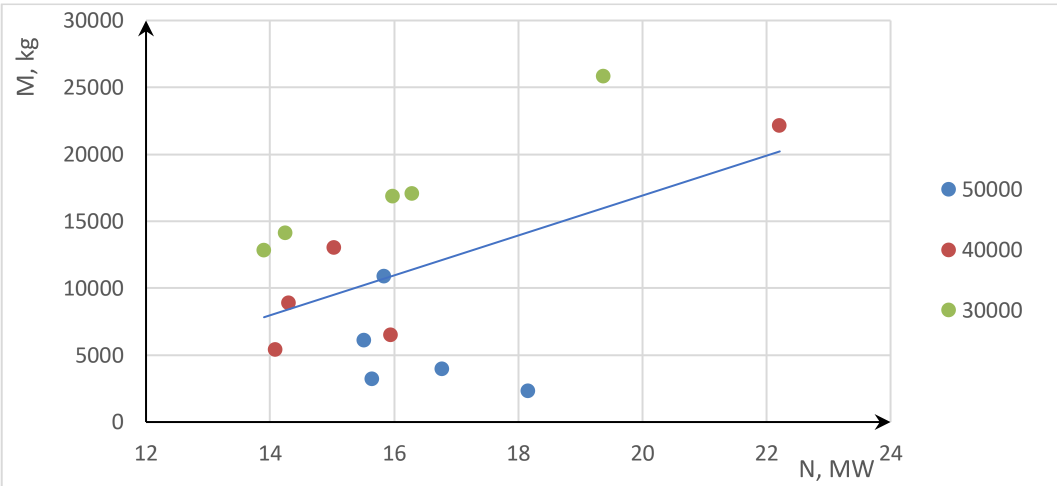
At the same time, overlapping of power values of different volumes of the cooling system allows working in one range of power consumption values (8-10 MW) in all three volumes of the nitrogen cycle, which provides a versatility of operation of the three cooling systems. This trend persists for all the calculated capacity modes, but at different fuel gas consumption levels. Thus, variations in raw plant load levels can be compensated for at any volume in the nitrogen cycle.

2.5.5. Dependence of power consumption on the recycle volume in the system of the plant

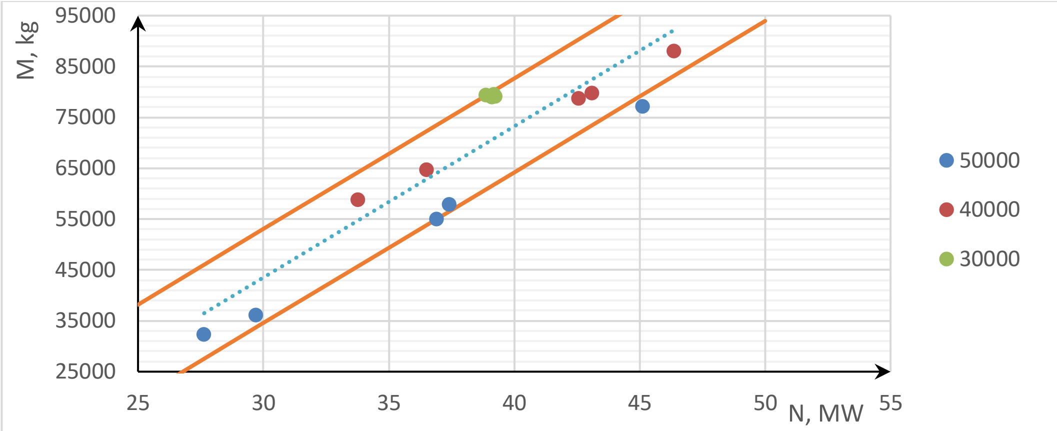
Figure 26 shows the dependencies of power consumption on the volume of the recycler in the system of the studied natural gas liquefaction plant for three volumes of the nitrogen cycle: 30000, 40000, 50000 m^3 for the whole range of raw gas capacities.



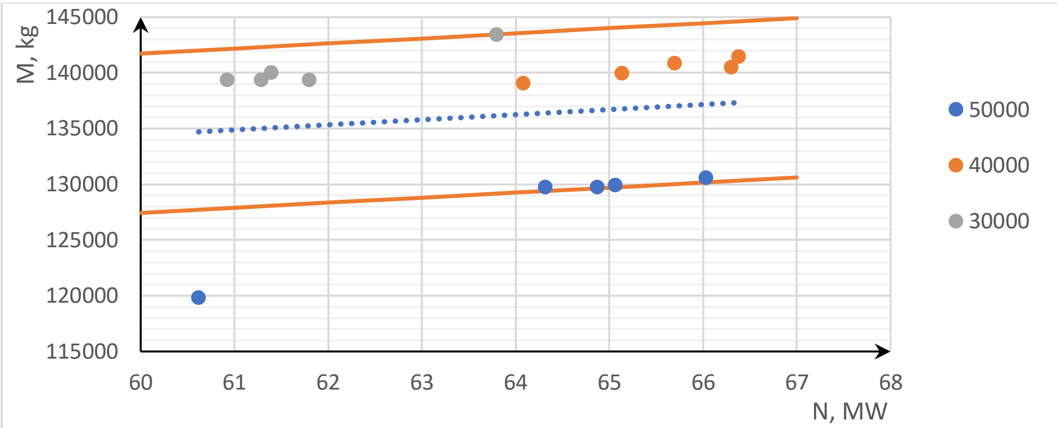
a) Natural gas flow rate 5500 Nm^3/h



b) Natural gas flow rate 10000 Nm^3/h



c) Natural gas flow rate 20000 Nm^3/h



d) Natural gas flow rate 30000 Nm^3/h

Figure 26: Private dependencies of power consumption on the volume of recycling in the system

According to the graphs, the distribution of design data points can only be described by specific patterns for each level of raw gas productivity. However, only the design scheme with the raw gas productivity of 5500 Nm^3/h can be used for the full mathematical analysis to find the approximating average with minimal errors. Other modes of operation (Fig. 26 b, c, d) with consumption of raw gas in 10000, 20000, and 30000 Nm^3/h have considerable error (Table 6) and are analyzed differently. For this purpose, a universal dependence was determined, the graphic interpretation of which is shown in Fig. 27.

Table 6: Error by Value

V \ N	$M_{recycle}$ (N2)	$M_{own\ needs}$(N2)	$M_{own\ needs}$ (N)	$M_{recycle}$ (N)
5500	4,41	4,49	0,65	9
10000	18,87	19,61	3,64	
20000	38,46	30,51	5,56	40,91
30000	9,23	8,89	6,95	10,37

Table 7: Main flow rates of the plant depending on the design model

Level	Unit capacity, thousand Nm^3/h						
	5,5		10,0		20,0		30,0
	One-line	Six-lines	One-line	Three-lines	One-line	Two-lines	One-line
Power consumptions, MW							
Max	12,07	72,40	22,21	66,64	46,37	92,75 (69,56)	66,38
Min	7,58	45,48	13,90	41,71	27,64	55,27 (41,45)	60,62
Gas for own needs, Nm^3/h.							
Max	1089,3	6535,6	1948,5	5845,5	4120,1	8240,2 (6180,1)	5847,3
Min	732,8	4397,0	1340,2	4020,6	2621,4	5242,8 (3932,1)	5338,8

Table 7 is a comparative analysis of different LNG plant capacities. For a clearer analysis, a "lines" distribution was made. This was necessary to compare low production rates (e.g., 5,500 Nm^3/h) with higher ones (30,000 Nm^3/h). Thus, based on the analysis of the plant operation criteria (Figures 17-27, Table 6-7), the optimal mode of operation is to load the plant with raw materials of 5500 Nm^3/h.

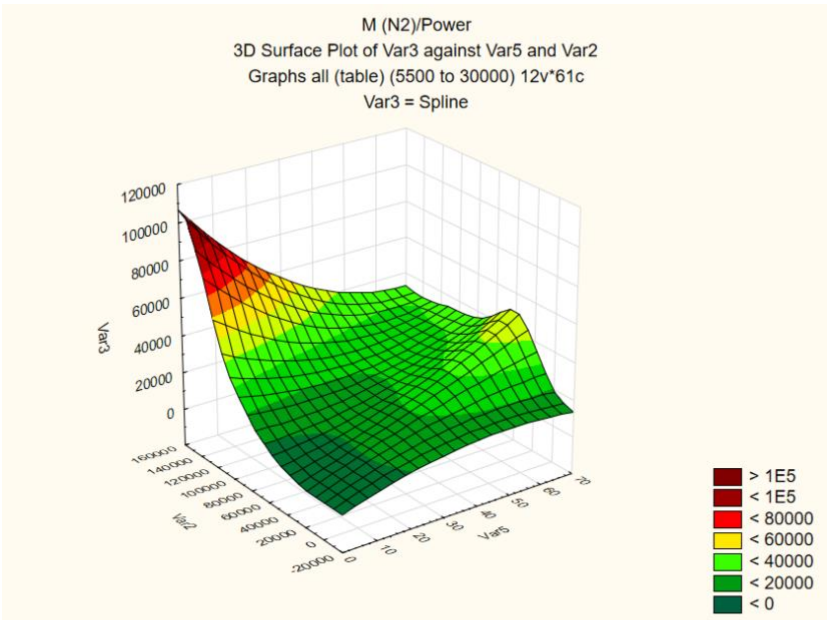


Figure 27: Universal dependence of power consumption on the volume of the recycler

Table 8: Variables for the graph in Figure 27

Variable	Var 2	Var 3	Var 5
Name	Gas volume in recycle, m^3	Gas volume in the system, m^3	Power consumptions of the plant, MW

Since the volume of the recycle is highly dependent on the raw gas capacity of the plant, the optimal mode of operation of the plant at the raw gas capacity of 5,500 Nm^3/h is optimal in terms of power consumption and flexibility of the approximating average.

2.6. Specification of modes of the design installation

In this section we made calculated specifications of the operating modes in the designed natural gas liquefaction plant.

During this analysis, nomograms of gas dependence were built (Figures 28-29). These dependencies were built for possible volumes of nitrogen cycle from 30000 to 50000 kg. These dependencies were plotted based on the flow separation in the H1 flow splitter. That is, the percentage of separation into two flows: O-6 and O-7.

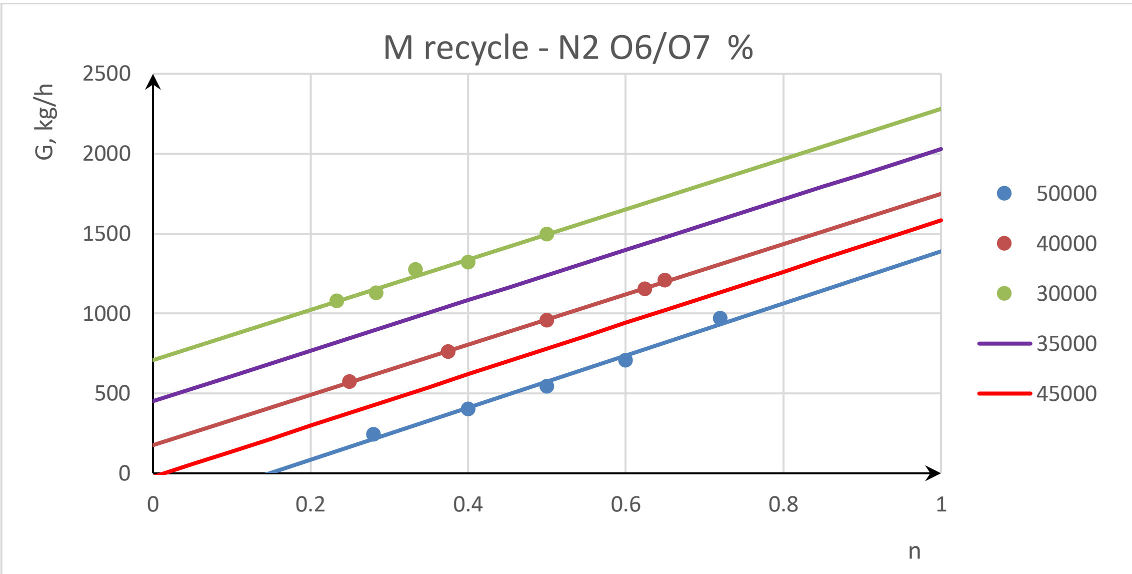


Figure 28: Nomogram of dependence of the gas quantity in the recycle on nitrogen separation in a small and large cycle of the cooling system

The subsequent calculation of the mathematical relationship of these curves from the nomograms in Figures 28-29 are shown in Table 9-10, as well as equations 1-7. It was from these mathematical equations that the curves for 35000 and 45000 kg were obtained. These curves have been added to Figures 28-29 to clearly demonstrate the smooth control of this unit, specifically for the nitrogen cycle and the gas used.

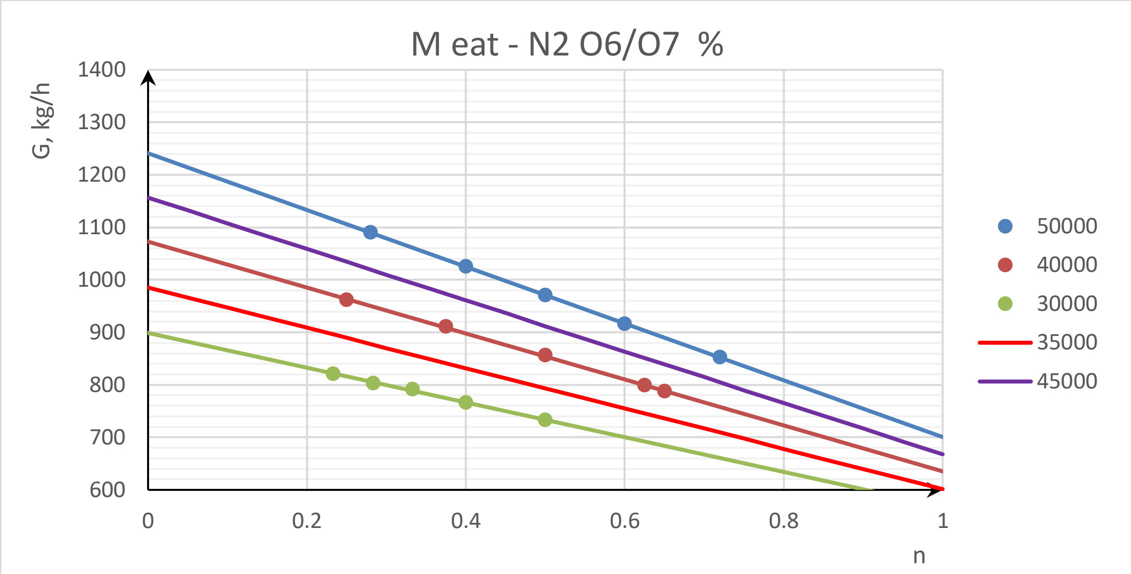


Figure 29: Nomogram of dependence of gas quantity on own needs on nitrogen separation in a small and large cycle of cooling system

Table 9: Input data for finding the coefficients of dependence of gas quantity in the recycle on nitrogen separation in a small and large cycle of the cooling system

		A_{N_2}		B_{N_2}
y	=	1573,1	+	707,81
		1573,0		175,82
		1630,7		-241,75

$$\frac{1630{,}7-1573{,}1}{50000-30000}=\frac{A-1573{,}1}{N_2-30000} \quad \text{(Eq. 1)}$$

$$A_{N_2}=\frac{(1630{,}7-1573{,}1)\cdot(N_2-30000)}{20000}+1573{,}1 \quad \text{(Eq. 2)}$$

$$\frac{-241{,}75-707{,}81}{20000}=\frac{B-70781}{N_2-30000} \quad \text{(Eq. 3)}$$

$$B_{N_2}=\frac{(-241{,}75-707{,}81)\cdot(N_2-30000)}{20000}+707{,}81 \quad \text{(Eq. 4)}$$

$$y=A_{N_2}\cdot x+B_{N_2} \quad \text{(Eq. 5)}$$

$$y=(0{,}00288\cdot N_2+1486{,}7)\cdot x+(-0{,}097478\cdot N_2+2132{,}15) \quad \text{(Eq. 6)}$$

Table 10: Input data for finding the coefficients of dependence of gas quantity for own needs on nitrogen separation in a small and large cycle of cooling system

		A_{N_2}		B_{N_2}
y	=	-330,84	+	898,82
		-437,48		1072,7
		-540,52		1241,1

$$y=(-0{,}010484\cdot N_2-16{,}32)\cdot x+(0{,}017114\cdot N_2+385{,}4) \qquad \text{(Eq. 7)}$$

Based on the analysis of performance indicators of LNG plant operation modes by results of calculation experiment, it was obtained that at flow rates above 5500 Nm^3/h, the gas cooling system fails to cope with the load, i.e., works unstable and with a large error in the required values. For example, the maximum error of the individual functional curves for a plant load at 5,500 Nm^3/h was 9.00 %, while at higher loads it was up to 40.91 %. At the same time, despite the small errors in the partial functional relationships at the maximum plant capacity loading considered, it was concluded that the process flow diagram has little flexibility for a possible reduction in plant capacity.

It was found that the amount of system recycles and gas for own needs significantly depend on the performance of the plant itself and in terms of power consumption and versatility of the average approximating line also show that the operating mode of the plant at the capacity of 5500 Nm^3/h is the most optimal.

The thesis assessed the technological efficiency of a range of possible LNG plant control modes, depending on potential customers' request. It has been revealed that the most flexible has an LNG plant consisting of 6 lines of 5500 Nm^3/h (shown in Table 7) which can provide both the minimum possible loading of the plant and the maximum exceeding required one by 30 %.

In addition, the overlap of the different volume ranges at the selected feed gas capacity provides the possibility to use different variations of the nitrogen cycle volumes, resulting in greater optimization of operating costs and flexibility of plant control.

Chapter 3. Improvement of equipment in the gas reduction unit

In this chapter various variations of the reduction unit design, i.e. the calculation scheme with turbo expander unit and supersonic separator, are considered. Additionally, a comparative analysis of the two liquefaction units is given.

These improvements are based on the plant from the last chapter and the change is only in the reduction unit. However, there is a slight change in the layout of the other equipment in the circuit. This change is since the turbo expander unit does not tolerate a liquid phase for the gas reduction process, as it has a turbine designed for gas only. The situation is similar to the 3S-separator. A supersonic separator loses a lot in efficiency and capacity if a liquid phase is present. That is why this equipment had to be moved in front of the second heat exchanger, right after the separator, which separates the liquid fraction of heavy hydrocarbons and prepares the gas phase for passing through the reduction unit.

3.1. Calculation of an operating plant with turbine expander

This section calculates a liquefied natural gas plant using a turbo expander unit. This scheme is a logical continuation of the previous study. However, due to some disadvantages, it was necessary to change the location of the turboexpander unit relative to the throttle in the previous version of the plant.

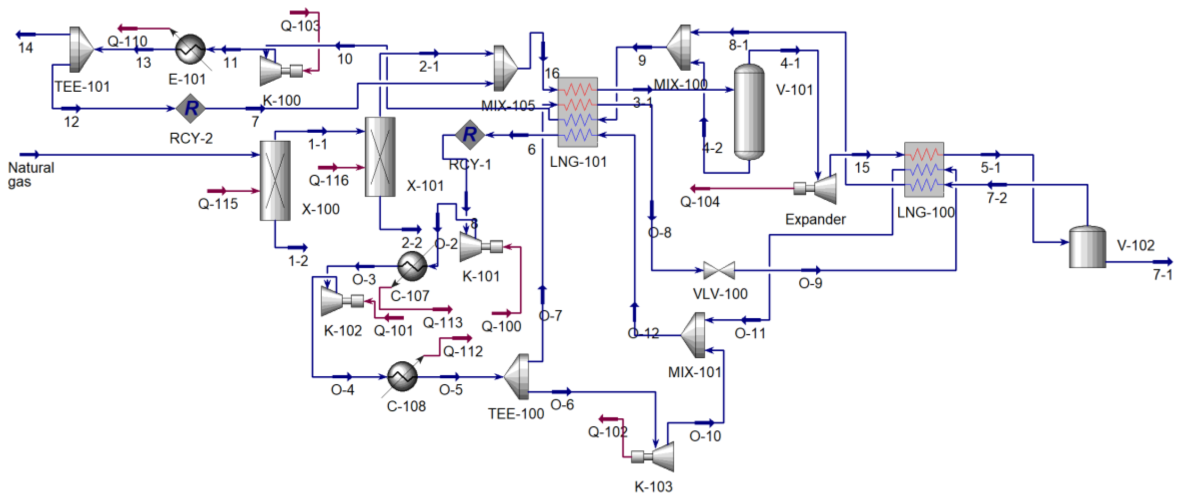


Figure 30: Basic schematic diagram of a natural gas liquefaction plant with a turboexpander

This unit's basic principle of operation is based primarily on the pretreatment of gas from the trunk pipeline. The gas enters the dehydration unit and the purification unit. The dehydration unit is required to remove all moisture from the gas to avoid hydrate formation. The purification unit is required to remove CO2 from the gas. After that, prepared gas enters the heat exchanger, where gas cooling takes place. This cooling is performed by cold energy from the nitrogen cycle. Gas after the first heat exchanger enters the separator. In the separator the heavy hydrocarbon liquid phase is separated from the light gaseous phase. Heavy hydrocarbons escape into the return flow, where it mixes with the gas phase from the tank, forming a recycle flow of the system. After the separator, gas under high pressure (from the main pipeline - 7.5 MPa) enters the turbo expander unit. In it the gas is strongly cooled due to a sharp pressure change. After that the gas-liquid phase enters the second heat exchanger, where almost complete gas liquefaction takes place and enters the tank (storage) for further transportation.

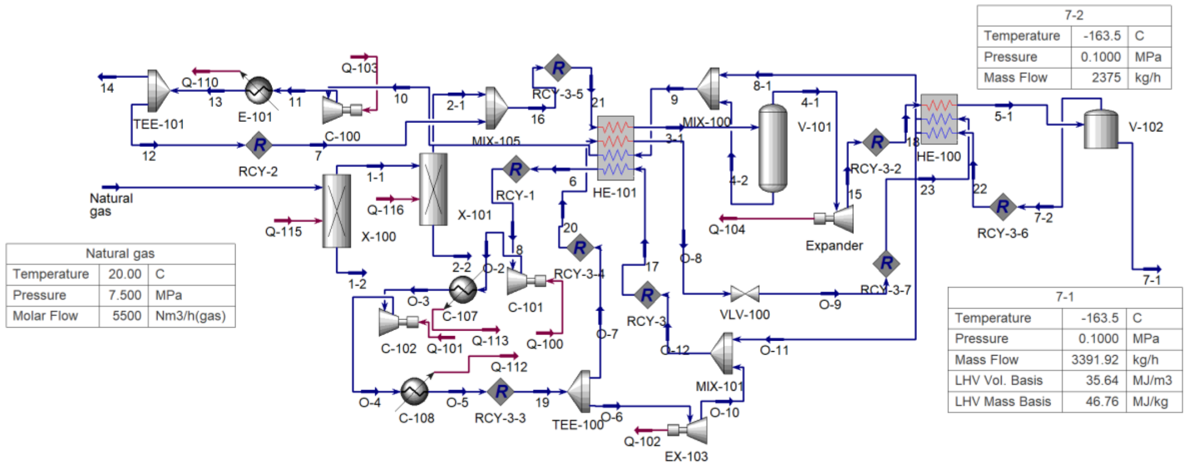


Figure 31: Basic liquefaction plant layout with initial and output data for LNG

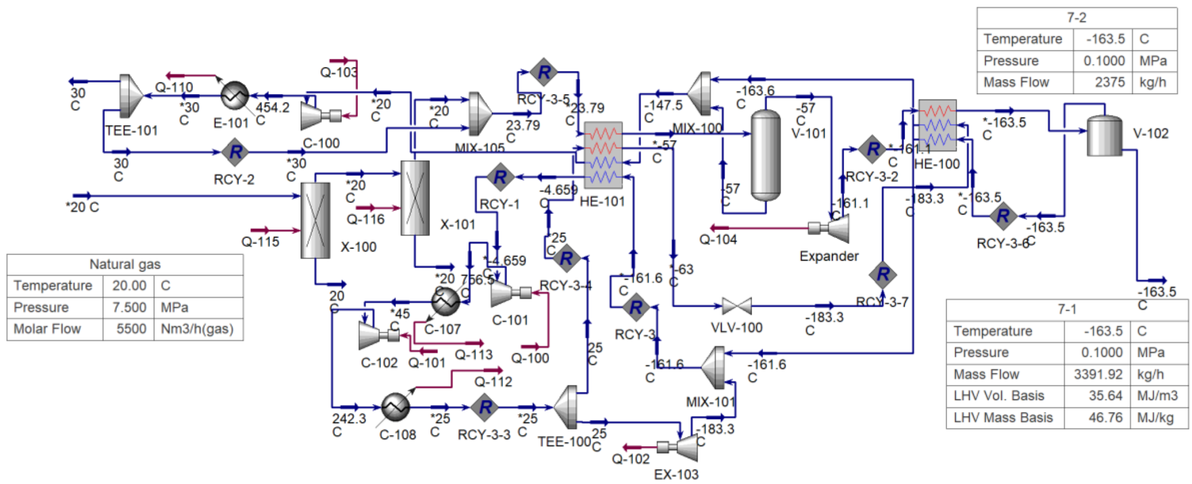


Figure 32: A gas liquefaction plant based on a turbo expander unit with temperature distribution throughout the system

Using Aspen HYSYS software, liquefied natural gas at -163.5 degrees Celsius is obtained at the plant outlet. This temperature allows unpressurized LNG storage in cryogenic tanks with forced venting of the vapor cap. It is also a cold LNG because its temperature is less than -140 degrees Celsius (Warm LNG).

The advantages of using a turbo expander in this application include the generation of electricity by harnessing the energy of the gas flowing through the turbine, rotating the electric motor. Here, the electric motor performs two functions simultaneously. The first is to start the turbo expander into operation. The second is the generation of electricity for the plant's own needs during operation.

3.2. Calculation of an operating plant with a supersonic separator

This section deals with the supersonic separator (3S separator). The working group has been working on this topic. I was involved in this working group to solve a certain number of tasks. I solved the following tasks: the construction of the computational profile in Excel, compilation of the temperature calculation profile for the supersonic nozzle, the selection of diameters of the inlet and outlet nozzles, as well as the comparative analysis of the Excel calculation model and simulation in ANSYS Fluent. This supersonic separator is designed specifically for the operating conditions of a natural gas liquefaction plant with the parameters used in the design scheme. In other words, all boundary and initial conditions for the design natural gas liquefaction plant used in this master's thesis.

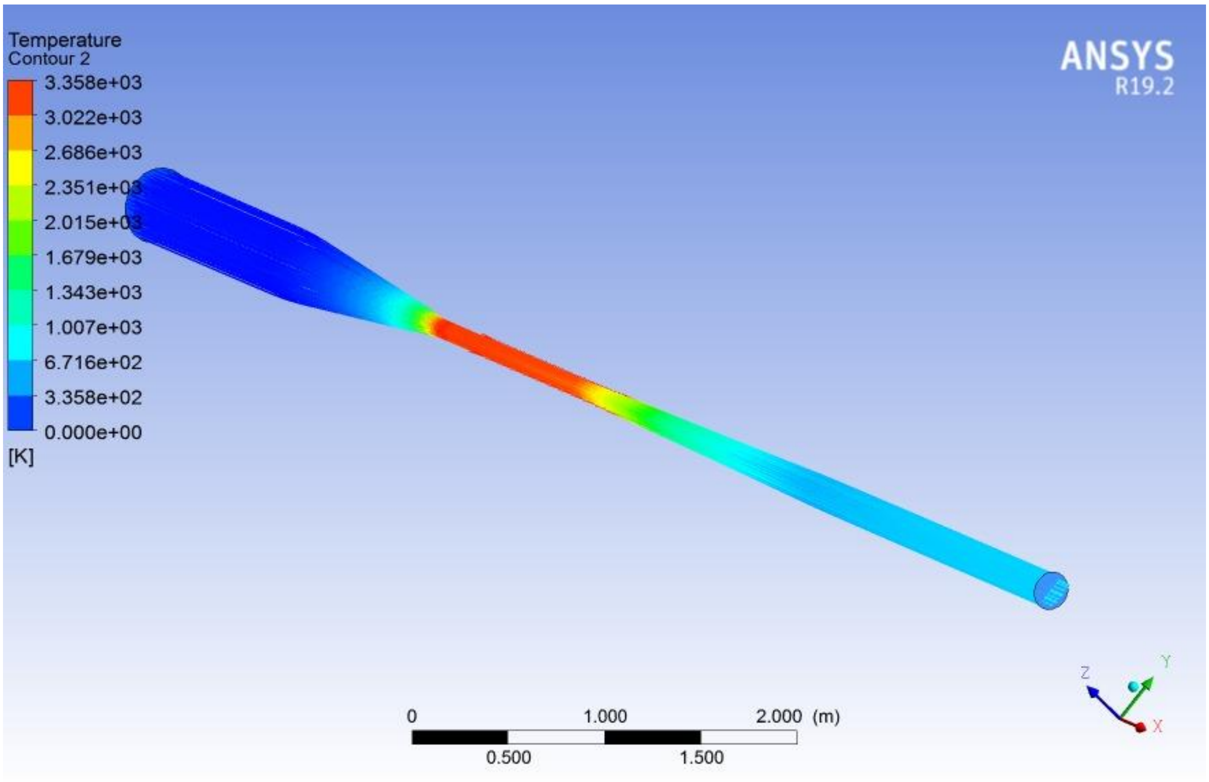


Figure 33: The temperature profile of a 3S-separator (incorrect)

In this case, the 3S-separator frame was modeled in the ANSYS software package while constructing the temperature profile of the separator. During the temperature analysis, an incorrect grid layout was selected. Also, during model preparation, the gas flow rate inside the separator was incorrectly selected, which also resulted in wrong output data and, consequently, incorrect temperature profile distributions. It can be concluded that a more in-depth analysis of the model, selection of the grid partitioning, and finer adjustments are required to obtain stable and correct results.

As can be seen from the second model (Figure 34), the analysis of the above weaknesses of the past model eventually provided the necessary data to modernize the current model. However, in doing so, several assumptions were made. Furthermore, elaborate model requires further analysis and a more detailed design.

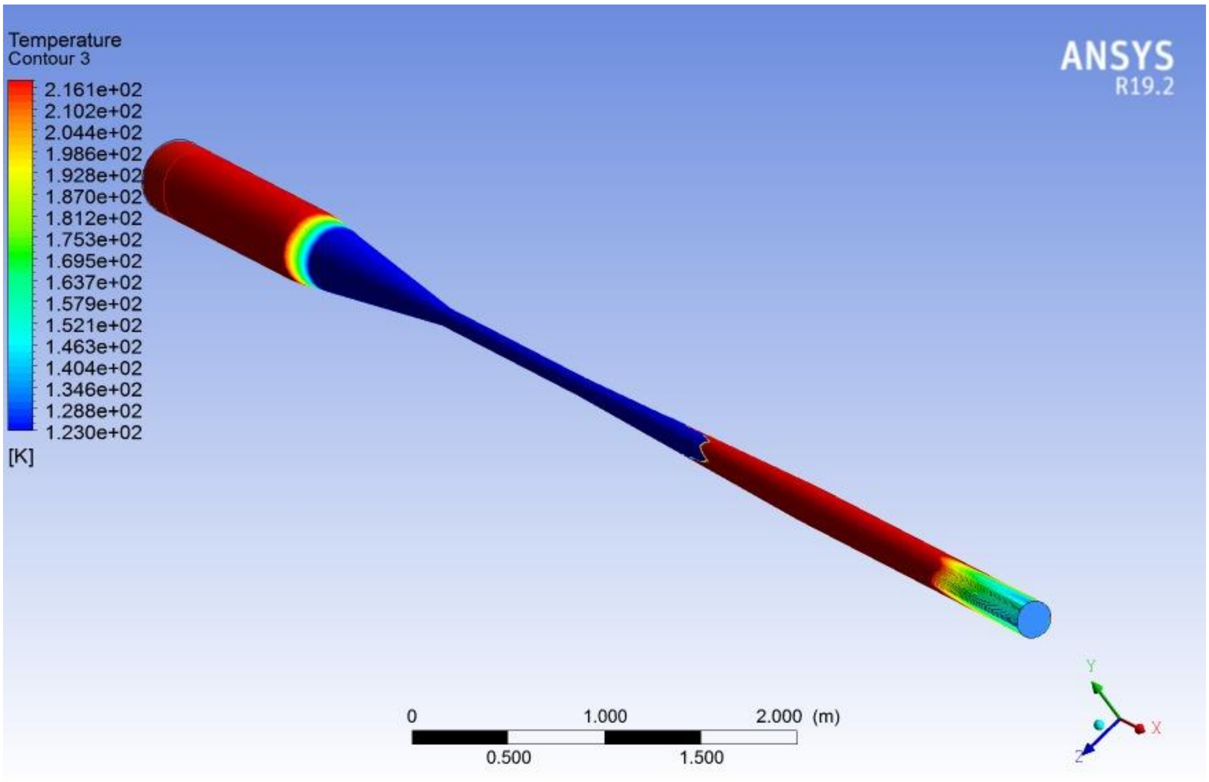


Figure 34: The temperature profile of a 3S-separator (correct)

In addition, it is possible to observe that the temperature profile is correct for the designed temperature differential in the gas reduction unit for LNG production. The temperature at the inlet of the 3S-separator can be seen to be brought up to -55°C. This heating is since kinetic energy is converted back into potential one. The temperature then gradually decreases due to the Joule-Thompson effect in the nozzle, after overcoming supersonic speed and the subsequent expansion of the gas helps the temperature rise to the inlet temperature. However, as the gas is further travels through the diffuser of the supersonic separator, the temperature decreases due to the Joule-Thompson effect, and at the outlet of the diffuser portion of the separator -150°C is obtained, which is a sufficient temperature to further cool the gas for its liquefaction through the second heat exchanger.

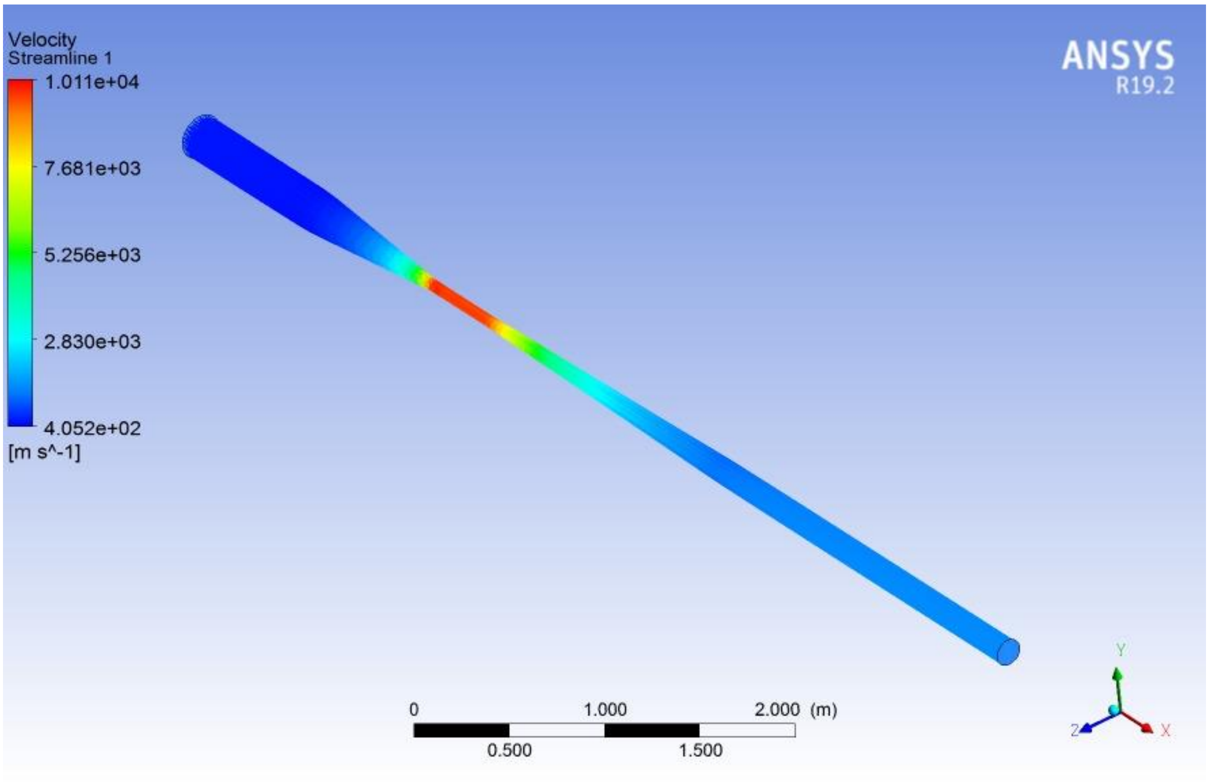


Figure 35: Velocity distribution profile inside the supersonic separator

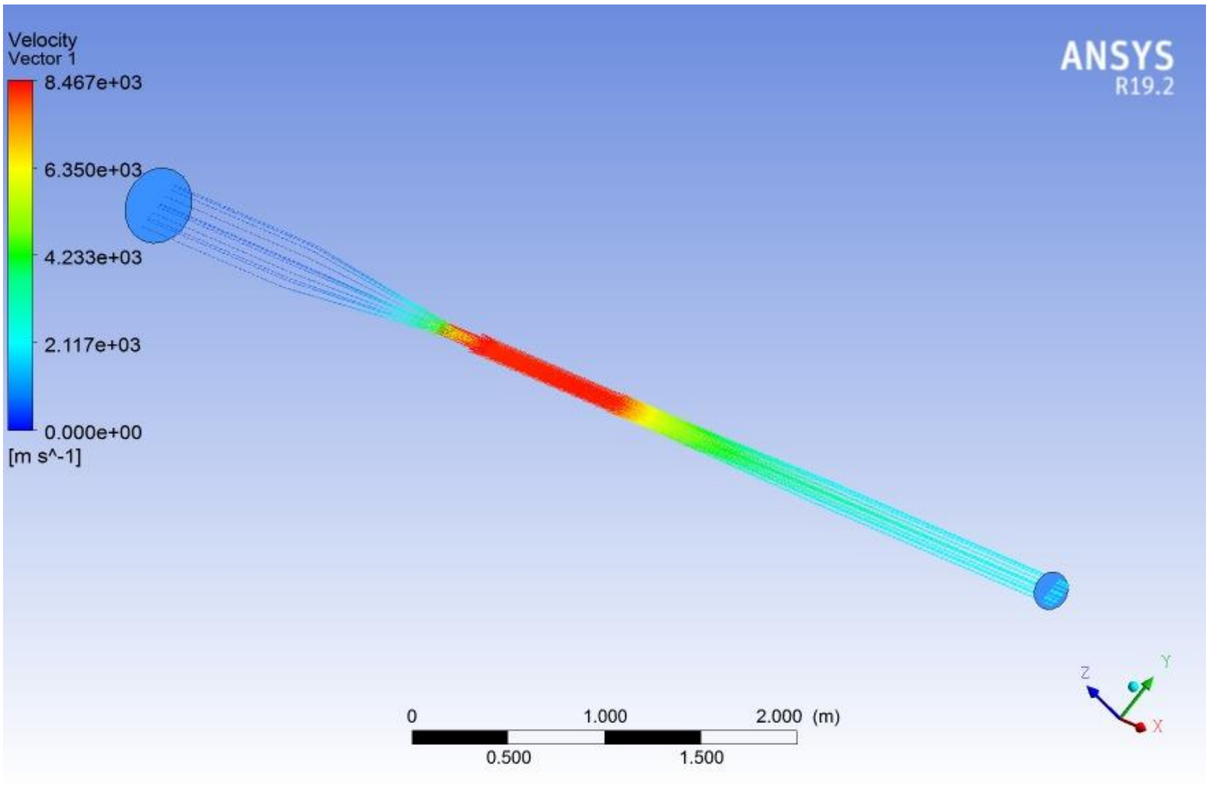


Figure 36: Vector distribution of gas velocity inside a supersonic separator

As can be seen from the velocity distribution inside the separator, it can be seen how the 3S-separator works. It is at the junction of the confuser (nozzle) and diffuser, at the narrowest passage diameter, that the transition to supersonic gas velocity occurs.

The uniform velocity distribution due to the contraction and further expansion of the gas inside the separator is displayed in different colors. This confirms that the model is correct and that the supersonic separator in the ANSYS software package is functioning properly.

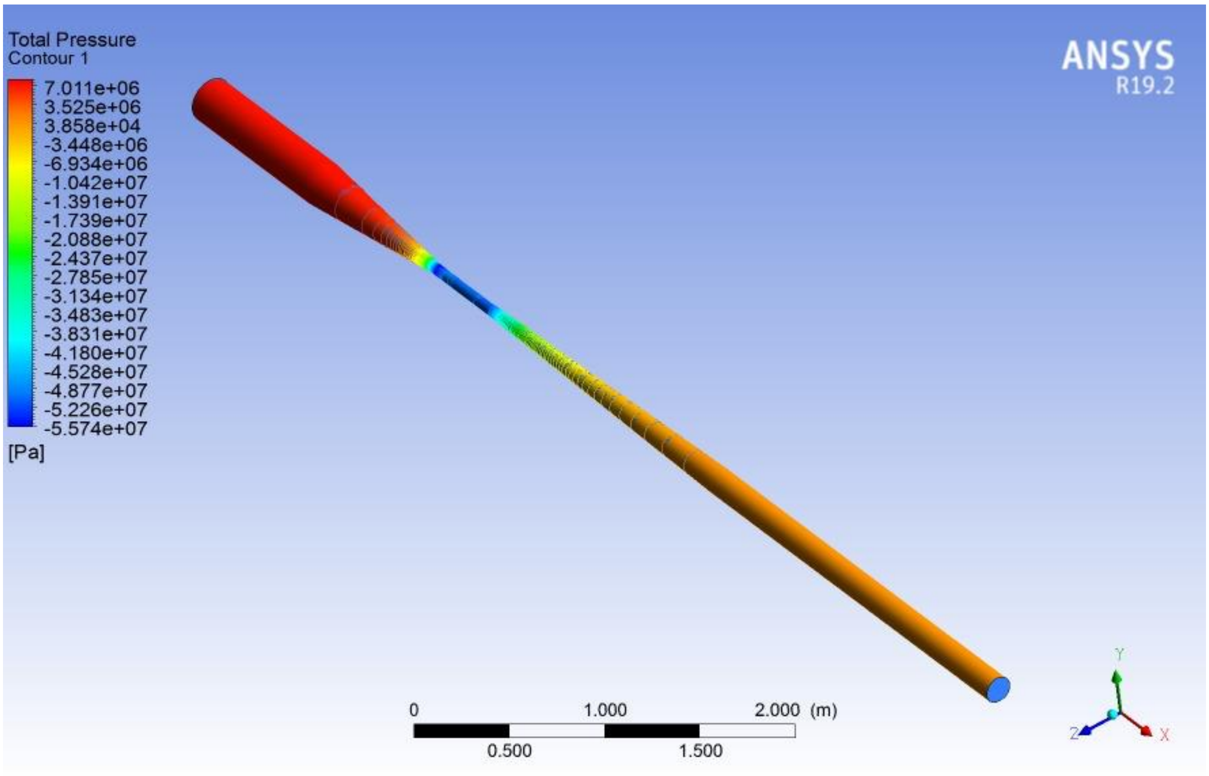


Figure 37: Pressure distribution profile inside the 3S-separator

Gas at 7.5 MPa enters the 3S separator. This gas comes from the main gas pipeline at this pressure. As can be seen from the distribution profile in the supersonic section, decompression is created due to high flow velocity and transition from kinetic energy back to potential energy and due to the Joule-Thompson effect. Due to this further liquefaction of the gas also occurs.

Profile of 3S Separator (Appendix A) is a calculated geometric profile of the supersonic separator, which was constructed and calculated as an analytical model.

Graph (Appendix B) shows the 3S-separator temperature profile, which was obtained by manual calculation of the separator. As can be seen the temperature discrepancy with the simulated Laval nozzle in the ANSYS software is negligible. Literally within the margin of error, as the discrepancy is only 8 degrees Celsius. This means that the model is correct.

3.3. Comparative analysis of an operating plant with improved equipment

In this thesis, a basic combined natural gas liquefaction scheme based on the throttle effect is considered. Since an improvement of the equipment is being considered, the efficiency of the equipment in question (turbo expander and supersonic separator) must be evaluated compared to the throttle type.

In this section comparison of all three LNG methods on several criteria is provided:
- The outlet temperature, i.e. the temperature difference due to the pressure drop.
- The efficiency of the pressure reduction unit
- The economic benefit (but this criterion will be considered in the next chapter of this thesis)

Effectiveness of temperature differential due to the Joule-Thompson effect upstream and downstream of the gas reduction unit is presented in the Table 11.

Table 11: Temperature differential on the various equipment in the reduction unit

Name of equipment	Inlet temperature, K	Outlet temperature, K	Difference, K
Throttle	193	111	82
Turbo expander	216	112	104
3S-separator	218	112.5	105.5

As can be seen from the summary results and the temperature difference, the choke is the worst performer. So, selecting new equipment and finding out the efficiency was the right decision. It is now necessary to decide which equipment is better: the turbo expander or the supersonic separator. From the temperature differences, the supersonic separator is slightly more efficient than the expander. However, the difference is only 1.5 Kelvin. Which can only be an error in the calculation and software calculation. But let’s assume that the 3S-separator is more efficient. Therefore, the supersonic separator has the best performance according to this criterion.

If you consider all three installations, the most inefficient would be the choke, as it has no additional effect other than pressure relief and temperature reduction without phase separation. After the choke a gas-liquid mixture comes out, i.e. after the choke, an additional separator has to be put to separate gas from LNG.

We also get gas-liquid mixture when using turbo expander and it also requires additional equipment after it. However, this turbo expander unit additionally generates power for its own needs for the whole natural gas liquefaction complex.

With the supersonic separator, the gas liquefaction process takes place simultaneously as well as the separation of the gas-liquid mixture, which does not require any additional equipment downstream.

Consequently, an economic analysis is also needed to show the most efficient type of equipment to be used at a natural gas liquefaction complex.

Chapter 4. An economic evaluation of the natural gas liquefaction complex

In this chapter we will look at the economics of each type of equipment, and separately take a closer look at the current and new equipment in the reduction unit. We will also carry out an economic analysis of projects with different equipment to find the most suitable design for a natural gas liquefaction plant.

Calculation of the LNG plant specific energy consumption shown in Eq. 8

$$E = \frac{P}{G} \qquad \text{(Eq. 8)}$$

Where E - specific energy consumption of LNG plant (kWh/ton), P - total energy consumption (±270000 kW), G - mass flow rate of produced LNG (t/h).

The estimated cost of annual LNG production shown in Eq. 9

$$C = s \cdot G \cdot T \cdot E \qquad \text{(Eq. 9)}$$

Where C - estimated cost of LNG production per year ($), s - average cost of 1 kWh ($0.015), G - mass flow of LNG produced (3.4 t/h), T - working hours per year without overhaul and commissioning (8400 hours).

Based on Equation 8, the specific energy consumption in the original scheme is

$$E = \frac{270000}{3{,}4 \cdot 30 \cdot 24} = 110\ kW \cdot h/ton$$

Estimated annual cost for the original scheme

$$C_o = 0.015 \cdot 3{,}4 \cdot 8400 \cdot 110 = 472500\ \$\ per\ year$$

Table 12: Specific Energy Consumption and Estimated Cost of LNG for the Scheme

Parameter	**Value**
Specific energy consumption, *E*, [kWh/ton]	
Turboexpander	110
3S-separator	118
Throttle	86
Estimated annual cost of LNG production, C_c, [$]	
Turboexpander	472500
3S-separator	505512
Throttle	368424

Apparently, the estimated cost of LNG annual production increases by $104076 and $137088 after modernization to the expander and 3S separator, respectively.

Additional costs will also arise for new units, such as Expander, 3S-Separator and extra separators. Valve costs are assumed to be neglected because they are relatively small. Approximate installation costs are shown in Table 13.

Table 13: Costs of all equipment of basic scheme

Equipment	**Quantity**	**Cost per piece, [$]**
Compressor	3	1.200.000
Throttle	2	6.500
Adsorbers (Dehydration unit)	3	300.000
Adsorbers (Purification unit)	4	560.000
Na-A zeolite	150 tons	1,9 per kilo
Na-X zeolite	200 tons	2,3 per kilo
Cryogenic tanks	4	300.000
Separator	2	80.000
Heat Exchange Equipment	2	130.000
Regasification plants	3	50.000
Total for the basic scheme		**9.508.000,00**

Table 14: Costs of new Equipment in the plant

Equipment	**Quantity**	**Cost, [$]**
Separator	3	80000
3S-Separator	1	300000
Turbo expander	1	3000000
Total for 3S-separator		**9.795.000,00**
Total for Turboexpander		**12.495.000,00**

Estimated annual income from LNG production is calculated according to Equation 10

$$R = T \cdot (G \cdot p_{LNG}) \qquad \text{(Eq. 10)}$$

Where R - annual income [$], T - annual operating hours without overhaul and commissioning (8400 hours), G - mass flow rate of produced LNG (3500 kg/h), p_{LNG} - average LNG market price (480$/ton). Price of gas according to statistic on 28th of February 2021: Gas TTF: 6,8 $/mmBTU (lhv)= 318 $/ton and liquefaction costs have to be added to gas price. Today values between 3 and 5 $/mmBTU for liquefaction might be used.[58]

[58] LNG as marine fuel - DNV. https://www.dnv.com/maritime/insights/topics/lng-as-marine-fuel/current-price-development-oil-and-gas.html

Estimated annual income from LNG production (Eq. 10)

$$R_c = 8400 \cdot \frac{(3500 \cdot 480)}{1000} = 14112000\$$$

The molar flow of produced LNG in the original scheme is 3400 kg/h and the average market price of LNG 318$/ton.

Estimated annual income from the original scheme according to Eq. 10

$$R_o = 8400 \cdot \frac{(3400 \cdot 318)}{1000} = 9082080\$$$

The estimated actual income during the first year of the basic scheme using Eq. 11

$$R_{real} = R_c - C_c - X \qquad \text{(Eq. 11)}$$

Where *X* is the total cost of additional equipment ($).

$$R_{real} = 14112000 - 505512 - 9795000 = 3811488\$$$

In the second year of operation estimated real income for the original scheme (Eq. 12)

$$R_{real} = R - C \qquad \text{(Eq. 12)}$$

Estimated real income from the second year of operation of this scheme

$$R_{real} = 3811488 - 505512 = 3305976\$$$

The results of all the calculation schemes for the three years ahead are presented in Table 15.

Table 15: Real Annual Revenue of all three schemes

Year of Operation	**First year, [$]**	**Second year, [$]**	**Third year, [$]**	**Sum, [$]**	**Project payback, years**
Original Scheme	4235576	3867152	3867152	11969880	2.88
Scheme with 3S-separator	3811488	3305976	3305976	10423440	2.93
Scheme with Turboexpander	3289000	2344000	2344000	7977000	3.57

After analysis of data presented in Table 15, it can be concluded that by the end of the second year of operation, the scheme with a supersonic separator will have recovered additional expenses.

Considering the cost of all equipment when building a new natural gas liquefaction plant, the most cost-effective plant is the one with the 3S-Separator design. However, given all the previous findings, it is safe to say that the supersonic separator has significant advantages over the throttle. At the same time, the cost of the projects in these two designs does not differ much (about $300.000). Although famous for installation in this type of plant, the turboexpander is much more expensive than the other designs. This suggests that this project is not feasible compared to the others. As a result, the best option would be a natural gas liquefaction plant with a 3S-separator. This project is more economical and more productive and efficient in terms of gas reduction and liquefaction to produce the final LNG product.

Conclusion

During the analysis of the global LNG production market, it was revealed that the main factors in the choice of LNG production technology are process efficiency, reliability of design solutions, simplicity of plant maintenance, possibility of modular design and low investment costs. Moreover, it is necessary to minimize operating costs. As a result of the analysis, a scheme based on an external isolated nitrogen cycle, by analogy with the technological schemes of APCI and JSC Cryogenmash, was selected as the basic one for further research.

Based on the analysis of performance indicators of LNG production regimes according to the results of the calculation experiment, it was obtained that at a flow rate over 5500 Nm^3/h, the gas cooling system cannot cope with the load. Which means that it does not work stably and with a large margin of error in the required values. So, the maximum error of private functional dependences for installation loading at 5500 Nm^3/h has made 9,00 % when on the big loadings on productivity to 40,91 %. Thus, despite of small errors of private functional dependences at the maximum considered load of the plant on productivity, it has been concluded that the technological scheme has small flexibility at possible decrease in productivity of the plant.

In this study, the technological efficiency of several possible regulation modes of the LNG plant depending on potential customers' requests has been evaluated. It was found that the LNG plant which consists of 6 lines of 5500 Nm^3/h has the greatest flexibility. Thus, it can provide both the minimum possible loading of the plant, and the maximum exceeding the required one by 30 %. In addition, the intersection of different volume ranges at the selected feed gas capacity provides the possibility of using different variations of the nitrogen cycle volume, which allows for more significant optimization of operating costs and flexibility of the plant regime control.

Turboexpander unit and 3S separator have the same potential, which is at the limit of calculation error. When TEU is reduced from the nitrogen cycle, the plant capacity is not reduced and remains at the same level. However, the number of dynamic equipment changes, thereby unloading the calculation scheme. Creating a 3S separator for these operating conditions will require a unique material that is cold resistant and at the same time not brittle. This is no opportunity to do that nowadays or it would be too hard and expensive to create such cryogenic material.

The economic analysis revealed that the most economically viable liquefaction plant is one using a supersonic separator. However, it does not consider the cost of materials and fabrication of this unit for use in cryogenic conditions. Consequently, further analysis is needed for this type of equipment. The unit with throttle and turboexpander is more expensive in design for LNG complex.

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Abbreviations

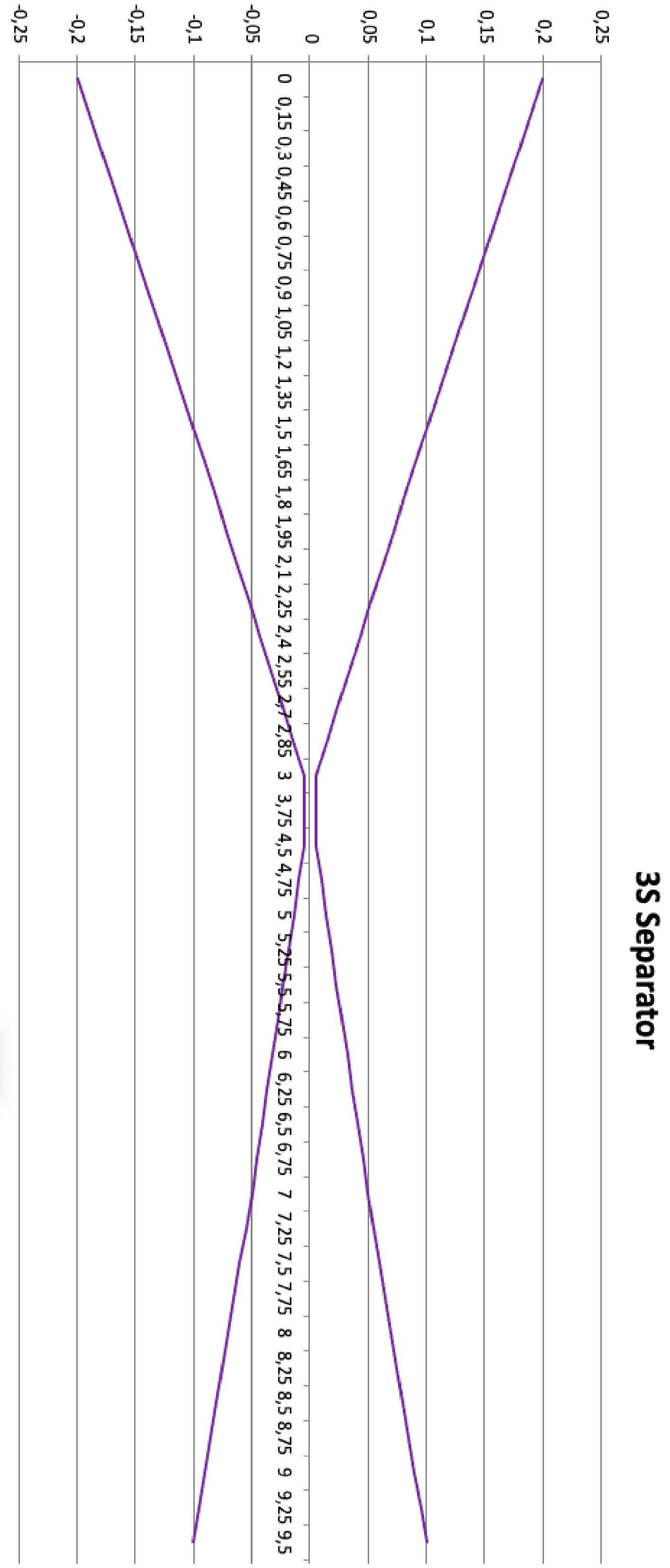
3S separator	Supersonic separator
APCI	Air Products and Chemicals Inc.
bcm	Billion cubic meters
C	Compressor
CNG	Compressed natural gas
DEA	Diethanolamine absorption
Eq	Equation
GCPU	Gas Condensate Processing Units
H	Flow splitter
HE	Heat exchancher
LNG	Liquified Natural Gas
LPG	Liqified Petroleum Gas
LTS	Low-temperature separation
mmBTU	million British Thermal Unit
NGV	Natural gas vehicle
S	Separator
TEE	Flow splitter
TEU	Turboexpander unit
TGE	Tractebel Gas Engineering
TH	Throttle

Nomenclature

C	Estimated cost of LNG production per year [$]
c_p	Mass heat capacity [J/kg/℃]
E	Estimated specific energy consumption of liquefaction plant [kWh/ton]
G	Mass flow [kg/h]
M	Mass flow [kg/h]
n	ratio of nitrogen cycle flow volumes
Q	Mass flow of liquefied gas [kg/h]
P	Power [J/s]
p_{LNG}	Average market price of LNG [$/ton]
R	Annual revenue [$]
s	Average cost of 1 kWh [$]
T	Number of working hours in year without workovers and commissioning operations [h]

Appendixes

Appendix A. Geometric profile of a 3S-separator



Appendix B. The calculated temperature profile of a 3S-separator