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Improvement of the industrial extraction method of methane homologues

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Absorption processes are one of the main ones in the preparation of hydrocarbon gases. However, in connection with a decrease in the pressure of the input gas, a number of facilities for the preparation of industrial products cannot provide the necessary technological regime and therefore require reconstruction. A typical example is the installation of low-temperature absorption at the main facilities «Solokha» of the gas industry management «Poltavagasvydobuvannya» where there is a decrease in the efficiency of the separation process of natural and oil gases. According to the results of analytical studies, to eliminate the indicated shortcoming, a variant of the reconstruction of the ULTA MF «Solokha» is proposed, which involves the use of a scheme with preliminary saturation of the regenerated absorbent. This will increase the extraction rate of components by 21.96%.

Keywords: low-temperature absorption, a wide fraction of light hydrocarbons, pre-saturation of the absorbent

Удосконалення методу промислового вилучення гомологів метану

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Процеси абсорбції є одними з основних при підготовці природного і супутнього нафтового газів. Застосування схем низькотемпературної абсорбції дає змогу забезпечити високе вилучення компонентів С3+ з вуглеводневих газів при порівняно помірному охолодженні технологічних потоків. Однак, у зв’язку зі зниженням тиску вхідного газу, ряд об’єктів підготовки промислової продукції не можуть забезпечити необхідний технологічний режим і тому потребують реконструкції. Типовим прикладом є установка низькотемпературної абсорбції (УНТА) на головних спорудах (ГС) «Солоха» газопромислового управління «Полтавагазвидобування». При роботі установки абсорбції під середнім і високим тиском разом з пропаном і вищими вуглеводнями абсорбентом поглинається також значна кількість метану і етану. У результаті розвиток процесу може лімітуватися за деякими компонентами внаслідок термодинамічної рівноваги. Отже, поглинання небажаних компонентів в абсорбері спричиняє підвищення середньої температури абсорбції та несприятливого формування профілю температур по висоті апарата. Наслідком цього є зниження ефективності процесу розділення природних і нафтових газів. За результатами аналітичних досліджень для усунення вказаного недоліку запропоновано варіант реконструкції УНТА ГС «Солоха», який передбачає застосування схеми з попереднім насиченням регенерованого абсорбенту. В такій схемі контакт сирого газу і насиченого абсорбенту здійснюється в холодильнику сирого газу при нижчій температурі, ніж в абсорбері. Попереднє насичення регенерованого абсорбенту сухим газом дозволяє підвищити глибину вилучення цільових компонентів з газу в абсорбері. У цьому холодильнику одночасно конденсується частина важких вуглеводнів, що призводить до зниження тепла абсорбції в абсорбері. Завдяки попередньому насиченню абсорбенту метаном, в самому абсорбері відбувається вилучення з газу в основному цільових вуглеводнів. Застосування вузла насичення регенерованого абсорбенту інертними компонентами дозволить мінімізувати наслідки зниження тиску вхідного газу і підвищити рівень вилучення компонентів широкої фракції легких вуглеводнів (ШФЛВ) на 21,96 % (в т. ч. пропану на 22,05 %, бутанів на 30,75 %). При цьому рівень вилучення досягне проектного.

Ключові слова: низькотемпературна абсорбція, широка фракція легких вуглеводнів, попереднє насичення абсорбенту

Introduction

Absorption processes are one of the main ones in the preparation of natural and associated petroleum gases. Absorption schemes, in addition to general nodes (modules) of separation, compression and drying of gas, have absorption and desorption modules [1].

The use of LTA schemes makes it possible to ensure a high extraction of propane from petroleum gases with relatively moderate cooling of process flows: on LTA installations, to extract 90-95% of propane, it is enough to have a refrigeration cycle with an isotherm of minus 30-38 °C [2]. At low-temperature condensation (LTC) installations, this requires an isotherm of minus 80-85 °C [3, 4].

Light absorbents (molecular weight 80-140) are used at LTA installations, their specific consumption is usually no more than 1-1.5 l/m^3. In combination with other measures, this made it possible to increase the degree of extraction of commercial products at the gas processing plant (GPP) and reduce the specific costs of gas processing by 25-50% (compared to the costs of conventional oil absorption plants) [5-8].

However, during absorption, the absorbent and the gas are heated due to the released heat, and the temperature of the absorbent is higher than the temperature of the gas, since the heat is directly transferred to the absorbent, and the gas is heated only due to the heat exchange between it and the absorbent [8, 9].

Peculiarities of the technological regime of the ILTA of the main facilities (MF) «Solokha»

The gas at the outlet of the ILTA installation must meet the dew point requirements for water and hydrocarbons. However, the pressure of the inlet stream is currently at the lower level permissible for the implementation of the process at the required level. It will be quite logical to envisage the option of its reconstruction with (preliminary saturation of the absorbent with an inert gas).

At ILTA (MF) «Solokha», pre-dried gas under a pressure of 2.2-2.4 MPa and at a temperature of minus 11.6 ºC enters two absorbers working in parallel (130,000 m^3/h). Although according to the regulations, this pressure should have been 4 MPa [10]. The ILTA installation allows to increase the extraction rate of C_{3+} higher hydrocarbons in the form of a broad fraction of light hydrocarbons (BFLH) by 80% (63 thousand tons per year, although the absolute value of propane extraction is only 27% of its content in raw gas).

The existing sheme of low-temperature absorption at the main facilities «Solokha» JSC «Ukrgazvydobuvannya» is shown in Fig. 1. In Tables 1 and 2. the composition of the input gas and adsorbent of ILTA at the main facilities «Solokha», respectively, is given.

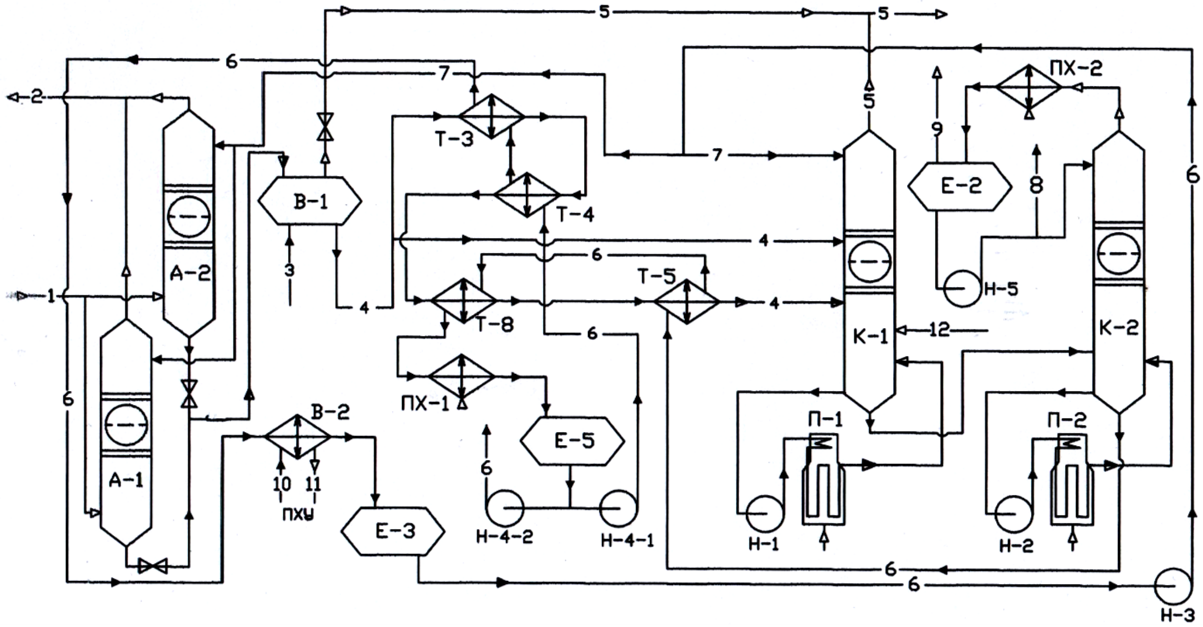

Figure 1 – **Scheme of low-temperature absorption of MF «Solokha»** [10]**:**
A-1, A-2 - absorbers; K-1 – absorption-evaporation column; K-2 – desorber; B-1 – weathering agent; T-3, T-4, T-5, T-8 – recuperative heat exchangers; P-1, P-2 – ovens; E-3, E-5 – collection tanks; E-2 – irrigation capacity; H-1, H-2, H-3, H-4-1, H-4-2, H-5 – pumps; B-2 – propane vaporizer of ACI; AC-1, AC-2 – air coolers; 1 – raw gas; 2 – dry gas; 3 – feeding with an absorbent; 4 – saturated absorbent; 5 – deethanization and weathering gas for own needs; 6 – regenerated absorbent; 8 – BFLH; 9 – blowing gas to the torch; 10 – liquid propane; 11 – propane vapors; 12 – refueling with the vapor phase from the absorbent preparation unit

Table 1 – Characteristics of the input and output gas of ILTA (%) [10]

Components	Raw natural gas in ILTA	Commodity dry gas in ILTA	Low pressure for own	Propane technical
N_2	1.112	1.582	0.287	
CH_4	88.67	89.475	59.71	
CO_2	2.477	2.396	5.923	
C_2H_6	5.037	5.125	31.62	1.232
C_3H_8	1.791	1.087	2.043	89.70
Iso-butane	0.201	0.09	0.127	5.265
n-butan	0.421	0.168	0.179	3.756
neopentane	0.005	0.02	0.001	0.006
iso-pentan	0.087	0.03	0.02	0.031
n-pentan	0.082	0.023	0.015	0.012
Fract. to 60 °C	0	0.002	0.007	10^{-4}
Fractю 60-80 °C	0	0.002	0.012	-
Fract. 80-100 °C	0.089	0.009	0.010	-
Fract. 100-120°C	0.015	0.007	0.027	-
Fract. 120-150°C	0.017	0.003	0.013	-
Fract. 150-160°C	-	0.0003	0.001	-
Fract. 160-180°C	-	0.0001	$7 \cdot 10^{-4}$	-
Fract. 180-200°C	-	$2 \cdot 10^{-5}$	$2 \cdot 10^{-4}$	-
Fract. 200-230°C	-		$2 \cdot 10^{-5}$	-
M_{aver}, kg/mol	18.54	18.08	22.95	45.2
Density, kg/m^3	0.746	0.728	0.924	503.8

Table 2 – Characteristics of absorbents of the ILTA installation [10]

Components	Regenerov. absorbent	Saturated absorbent	Absorbent with IPA (light)
N_2	-	1.25	$6 \cdot 10^{-5}$
CH_4	-	0.002	0.006
CO_2	-	0.004	0.007
C_2H_6	$6 \cdot 10^{-5}$	0.59	0.1
C_3H_8	0.04	7.64	0.85
Iso-butane	0.13	1.98	0.99
n-butan	0.57	3.98	2.15
neopentane	0.01	0.04	0.02
iso-pentan	1.07	1.52	2.36
n-pentan	1.39	1.62	2.99
Fract. to 60 °C	0.83	0.86	4.49
Fract. 60-80 °C	2.03	1.86	6.68
Fract. 80-100 °C	4.87	4.25	9.57
Fract. 100-120 °C	21.36	18.41	17.79
Fract. 120-150 °C	32.62	27.71	26.87
Fract. 150-160 °C	8.08	6.82	6.94
Fract. 160-180 °C	10.07	8.48	10.5
Fract. 180-200 °C	6.79	5.71	6.82
Fract. 200-230 °C	3.88	3.25	0.83
Fract. 230-250 °C	2.83	2.37	0.02
Fract. 250-270 °C	3.44	2.88	0.01
Fract. 270-290 °C	0.01	0.008	0
Fract. 290-320 °C	0.008	1.26	0
Fract. 320-250 °C	0.003	0.002	0
Fract. > 350 °C	0.0007	0.004	0
$M_{aver,}$ kg/mol	121.03	81.43	110.88
Density kg/m^3	755.11	623.49	716.89

Insufficient pressure, as well as the limited capabilities of the propane refrigeration plant for cooling the absorbent and gas complicate the process of absorption of C_{3+} higher in the absorbers and further affect the parameters of the processes of deethanization and regeneration of the absorbent. As practice shows, the disadvantages of the absorption process under such conditions are especially noticeable when the ambient temperature rises above 20-25 °C. At the same time, the temperature of the gas at the entrance to the LTA rises to minus 18 – minus 15 °C, the temperature of the regenerated absorbent when fed to the absorber to minus 10 – minus 8 °C. During this period, the production indicators of BFLH decrease by an average of 30% – from 120-140 to 90-100 t/day (although the design capacity of the installation is 200 t/day).

From Table 3 [10], which presents a verification calculation of the absorption of ILTA MF «Solokha» (theoretically possible), namely from the liquid plate enthalpy (absorption occurs in the liquid, and therefore the main release of energy) and the average temperature on each of the eight theoretical plates (the temperature difference between the first and second and second and third plates is 1.7 °C, while between the third and fourth, fourth and fifth plates is only 1.2 °C), it can be concluded that the majority of heat is in the upper part of the absorber is released during the absorption of methane and ethane.

At a given pressure and average temperature on the plates, the value of the phase equilibrium constant shifts towards the opposite process – desorption (Fig. 2).

The level of extraction of propane, as the main target component of the process (39% according to the regulations of the installation and 27% in real conditions), which can be seen from the difference in its content in the gas at the entrance and at the exit from the installation (Table 4), is hardly acceptable from the point of view of profitability in general for absorption processes and even more so for low-temperature absorption with significant costs for obtaining cold.

Table 3 – Verification calculation of absorption (one absorber), pressure 2.2 MPa [10]

№ theore-tical plate	Tempera-Ture °C	Flows, kmol/h		Enthalpy, kcal/kmol		Flows, kg/h		Density, kg/m^3	
		gas	liquid	gas	liquid	gas	liquid	gas	liquid
1	-5.9	2586.5	256.43	2025.62	-4617.29	47166.5	27332.09	19.32	761.53
2	-7.6	2638.4	263.51	22010.98	-4593.87	48498.56	27576.14	19.66	760.12
3	-9.3	2645.5	268.26	1995.25	-4605.64	48743.76	27737.64	19.88	759.73
4	-10.7	2650.3	272.46	1981.9	-4613.07	48907.3	27884.51	20.05	759.36
5	-11.9	2654.6	276.88	1970.34	-4610.16	49056.77	28047.08	20.21	758.77
6	-13.1	2659.2	282.56	1959.41	-4590.83	49222.14	28270.35	20.35	757.69
7	-14.5	2664.9	292.32	1947.22	-4533.86	49447.88	28685.97	20.57	755.35
8	-16.5	2674.7	321.42	1929.43	-4328.68	49865.15	30074.02	20.90	747.77

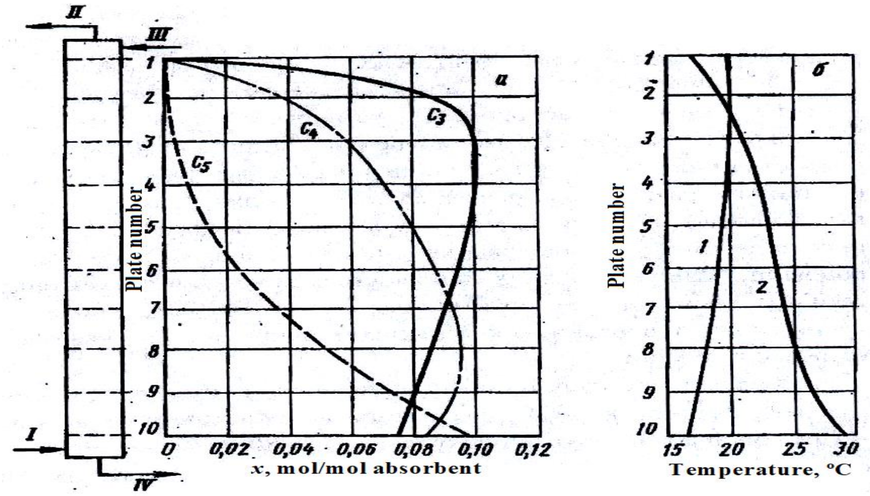

Figure 2 – Dynamics of changes in the height of the absorption apparatus [5]:
a – the content of saturated components in the absorbent; b – flow temperatures; And – raw gas; II – dry gas; III – regenerated absorbent; IV – saturated absorbent; 1 – gas; 2 – absorbent.

Table 4 – Material balance of ILTA MF «Solokha» with a gas productivity of 130,000 m^3/h before reconstruction

Compositions	gas inlet		withdrawal	Dry gas output		Withdrawal of BFLH		
	% mol.	g / m^3	% mol.	% mol.	g/m^3	g/m^3	t/h	t/day
C_3	1.791	43.0	35.56	1.14	27.71	15.29	1.9875	47.70
$\sum C_4$	0.701	17.7	55.86	0.16	7.81	9.89	1.2858	30.86
$\sum C_5$	0.179	5,40	100	0	0	5.40	0.7021	16.85
$\sum C_{6+}$	0.05	1.88	100	0	0	1.88	0.2438	5.85
$\sum CO_2, CH_4, C_2H_6$	-	-	-	-	-	-	-	3.48
$\sum$	2.72	67.98	47.74	1.30	35.52	32.46	4.2192	104.74

Review of the research sources and publications

Currently, the following technological schemes of absorption plants are used:

– absorption with partial evaporation of light hydrocarbons;

– absorption with recirculation of non-condensed gases;

– two-stage absorption;

– absorption with preliminary saturation of the regenerated absorbent.

Absorption with partial evaporation of light hydrocarbons involves heating the bottom of the column in order to partially evaporate methane (and sometimes ethane). At the same time, partial desorption of propane and heavier hydrocarbons occurs. Devices in which this process is implemented are called absorption-evaporation columns (AECs). At present, technological schemes of almost all absorption plants include AEC.

During the operation of the absorption unit under medium and high pressure, along with propane and higher hydrocarbons, the absorbent also absorbs a significant amount of methane and ethane. This complicates the desorption scheme. Due to the high pressure of the saturated vapor of the products at the top of the column, their condensation is difficult, since low temperatures are required.

At a number of gas processing plants, schemes with recirculation of residual gases. However, it is difficult to determine the optimal operating parameters of such an installation. On the one hand, it is desirable to carry out demethanization of the saturated absorbent under high pressure in order to reduce the load on the compression of the residual gas, on the other hand, increasing the pressure increases the metal capacity of the column and heat consumption for the regeneration of the absorbent.

In the LTA process, a number of issues arise when using a light absorbent: the removal of the absorber with gas increases both due to its dissolution in the gas and due to removal in the form of small droplets. To reduce the losses of the absorber with dry gas, two-stage absorption schemes are used: a light absorbent is fed to the first stage, and a relatively heavy one to the second stage of absorption. Multi-stage absorption schemes with different temperatures and pressures at individual stages are also used.

It is known that the process of absorption of hydrocarbon gases takes place with the release of heat – the greatest exothermic effect is observed in the upper and lower parts of the absorber, because the main mass of methane and ethane is absorbed at the top, and butane and heavier hydrocarbons are absorbed at the bottom. During the processing of oil gas of medium «fatness» ($C_{3+\text{higher}}$ = 300 g/m^3), more methane and ethane are absorbed in the absorber than propane and heavier hydrocarbons (in moles). And this means that the removal of unwanted components (methane and ethane) causes more heat release than the absorption of high molecular target hydrocarbons, since methane and ethane have higher heats of absorption at the operating pressures of the processes. The profile of the change in propane concentration, for example, is formed by the height of the apparatus so that sometimes in the middle part of the absorber propane desorption begins from the absorbent flowing from the higher plates (Fig. 2).

Accordingly, under these conditions, the development of the process may be limited by some components due to thermodynamic equilibrium. Therefore, probably, increasing the number of real plates in the absorber (more than 25-30) does not contribute to increasing the efficiency of the process. Therefore, in the conditions of the adiabatic regime, the absorption of unwanted components in the absorber causes an increase in the average temperature of absorption and an unfavorable formation of the temperature profile along the height of the device and, as a result, is one of the reasons for reducing the efficiency of the separation process of natural and petroleum gases.

The analysis of the temperature distribution along the height of the absorbers at different installations showed that the intensity of the heating of the absorbent is greater in the upper and lower parts of the apparatus, since the main amount of methane and ethane is absorbed at the top of the column, while the dissolution of butanes and pentanes occurs on the lower plates. Therefore, it is advisable to remove the maximum amount of heat of the dissolution process in intermediate refrigerators installed at the top and bottom of the absorber. However, schemes with intermediate refrigerators have a number of disadvantages: the presence of blind plates in the absorber, the difficulty of accurately choosing the place of introduction of the cooled absorbent, low heat transfer coefficients.

To eliminate the indicated shortcomings, it is possible to use schemes with preliminary degasification of raw gas and saturation of the regenerated absorbent. Presaturation of the regenerated absorbent with dry gas makes it possible to increase the depth of extraction of the target components from the gas in the absorber, since in this scheme the contact of raw gas and saturated absorbent is carried out in the refrigerator of raw gas at a lower temperature than in the absorber. In this refrigerator, at the same time, a part of heavy hydrocarbons condenses, which leads to a decrease in the heat of absorption in the absorber. Due to the pre-saturation of the absorbent with methane, mainly target hydrocarbons are extracted from the gas in the absorber itself. On the basis of calculations, it was found that the lower the content of heavy hydrocarbons in raw gas, the greater the negative impact of residual components in the regenerated absorbent on the depth of their extraction [8, 9].

Increasing the absorption pressure reduces the negative impact of residual components. For example, at pressures of 1.5 and 3 MPa and 15 °C, the presence of 1% (wt.) propane and butane in the regenerated absorbent reduces the extraction of propane in the first case by 15%, and in the second by 10%, butane by 5 and 3, respectively. 5% in comparison with the mode in which complete evaporation of the absorbent would be carried out [11, 12].

First of all, three pre-saturation schemes are used. The first option (Fig. 3) – the regenerated absorbent is mixed with dry gas of the absorption-evaporation column 4 and enters the propane evaporator 5 together with it. As a result of the contact of these flows, the regenerated absorbent is saturated with light hydrocarbons with the simultaneous removal of heat of absorption. After that, the saturated (ballast) regenerated absorbent is separated from the free gas in the separator 6 and fed to the upper plate of the absorber and AEC.

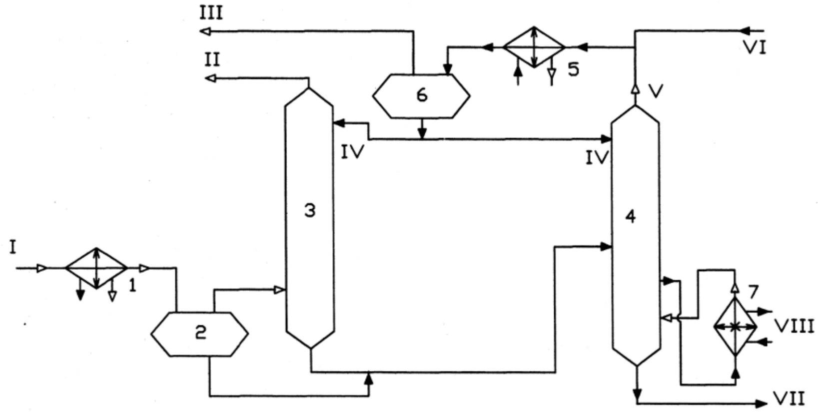

Figure 3 – Schematic diagram of the absorption unit with pre-saturation of the regenerated absorbent with dry AEC gas (option I):
1, 5 – propane evaporators; 2, 6 – separators; 3 – absorber; 4 – AEC; 7 – reboiler; And – raw gas;
II – dry gas; III – dry gas after preliminary saturation of the regenerated absorbent;
|IV – regenerated absorbent after previous saturation; V – dry gas AEC;
VI – regenerated absorbent; VII – deethanized saturated absorbent; VIII – coolant

The second option (Fig. 4) – the regenerated absorbent is mixed with dry AEC gas 4 and enters the propane evaporator 5 together with it. As a result of the contact of these streams, the absorbent is saturated with light hydrocarbons with the simultaneous removal of the heat of absorption. After separating this mixture in the separator 6, the saturated absorbent is divided into two streams – one is directed to the upper plate of the AEC, the other is mixed with the dry gas of the absorber and enters the propane evaporator 7. As a result, the absorbent is additionally saturated with light hydrocarbons. After the evaporator 7, the mixture of gas and absorbent enters the separator 8, from which the saturated regenerated absorbent is fed to the upper plate of the absorber 3, and the dry gas is sent to consumers.

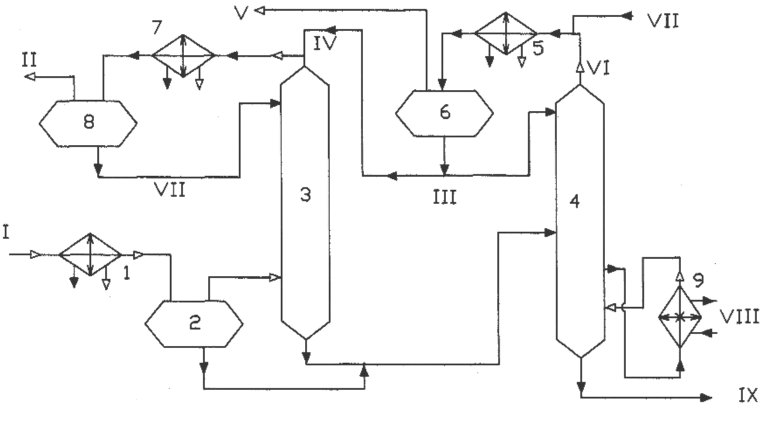

Figure 4 – Basic technological scheme of the absorption unit with preliminary saturation of the regenerated absorbent with dry gas of the absorber and dry gas of AEC (II variant):
1, 5, 7 – propane evaporators; 2, 6, 8 – separators; C – absorber; 4 – AEC; 9 – reboiler; I - raw gas;
II – dry gas after preliminary saturation of the regenerated absorbent;
III – regenerated absorbent after the pre-saturation node; IV, V, VI – dry gas;
VII – regenerated absorbent; VIII – coolant; IX – deethanized saturated absorbent

The third option (Fig. 5) – one stream of regenerated absorbent is saturated with light hydrocarbons as a result of mixing with dry gas of absorber C, after cooling in propane evaporator 7 and separation from gas in separator 8, it is fed to the upper plate of absorber 3. The second stream of regenerated absorbent is saturated with light hydrocarbons as a result of mixing with dry gas of the absorption-evaporation column 4 and after cooling in the propane evaporator 5 and separation from the gas in the separator 6 is fed to the upper plate of the absorption-evaporation column 4.

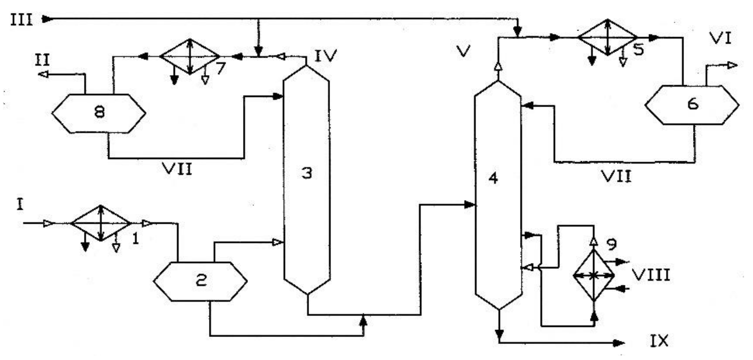

Figure 5 – Basic technological scheme of the absorption unit with preliminary saturation of the regenerated absorbent with dry gas of the absorber and dry gas of AEC (III variant):
1,5, 7 – propane evaporators; 2, 6, 8 – separators; C – absorber; 4 – AEC; 9 – reboiler; I – raw gas;
II – dry gas of the absorber after preliminary saturation of the regenerated absorbent;
III – regenerated absorbent; IV, V – dry gas; VI – dry AEC gas after pre-saturation
of the regenerated absorbent; VII – regenerated absorbent after previous saturation;
VIII – coolant IX – deethanized saturated absorbent

The installation uses two absorbents: light - with a molecular weight of 100 and heavy - with a molecular weight of 140.

In fig. 6 shows the technological diagram of the LTA installation with the absorbent pre-saturation unit of the gas processing plant designed for the extraction of ethane and heavier hydrocarbons from natural gas (Alvin, USA). [8]. A gasoline fraction with a molecular weight of 100 is used as an absorbent.

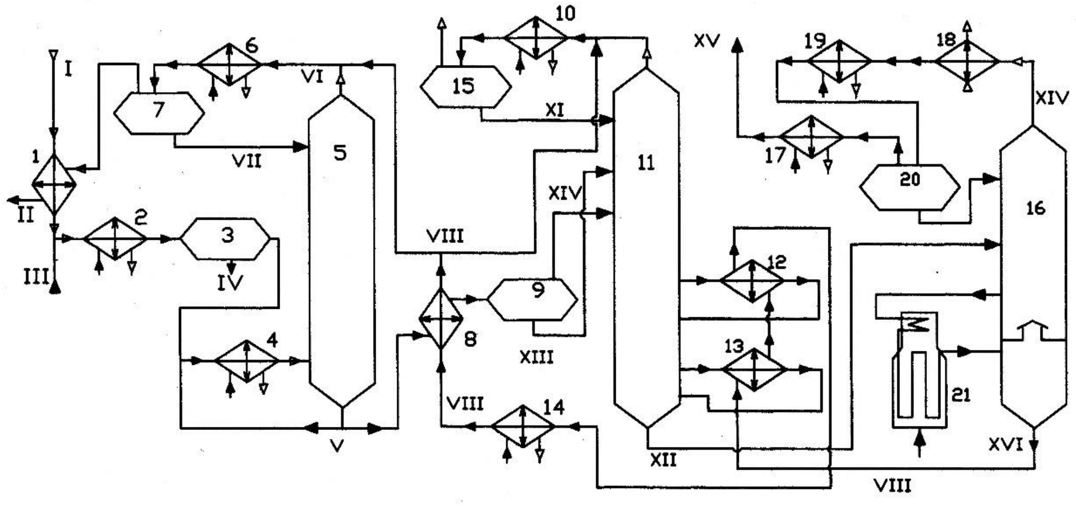

Figure 6 – Technological scheme of the LTA gas processing plant, designed for the extraction of C^{2+} HIGHER hydrocarbons from natural gas:
1, 8, 12, 13 – recuperative heat exchangers; 2, 4, 6, 10, 14, 17, 19 – propane vaporizers; Z, 7, 15 – separators; 5 – absorber; 9 – evaporator-separator; 11 – AEC; 16 – desorber; 18 – air cooler; 20 – reflux capacity; 21 – stove; 1 – raw gas; 11 – dry gas of the absorber after the assembly of pre-saturation of the regenerated absorbent with light hydrocarbons; II I – ethylene glycol solution; IV – hydrated ethylene glycol; V – saturated absorbent; VI – dry gas; VII – regenerated absorbent saturated with light hydrocarbons; VIII – regenerated absorbent; IX – dry gas; X – dry gas of AEC after pre-saturation of the regenerated absorbent with light hydrocarbons; XI – regenerated absorbent saturated with light hydrocarbons; XII – demethanized saturated absorbent; XIII – saturated partially degassed absorbent; XIV – gas; XV is a broad fraction of C_{2+} higher hydrocarbons

Definition of unsolved aspects of the problem

However, in connection with a decrease in the pressure of the input gas, a number of facilities for the preparation of industrial products cannot provide the necessary technological regime and therefore require reconstruction. A typical example is the installation of low-temperature absorption (ILTA) at the main facilities (MF) «Solokha» of the gas industry management JSC «Ukrgazvydobuvannya». During the operation of the absorption unit under medium and high pressure, along with propane and higher hydrocarbons, a significant amount of methane and ethane is also absorbed by the absorbent. As a result, the development of the process may be limited by some components due to thermodynamic equilibrium. Therefore, the absorption of unwanted components in the absorber causes an increase in the average absorption temperature and an unfavorable formation of the temperature profile along the height of the device. The consequence of this is a decrease in the efficiency of the process of separation of natural and oil gases

Problem statement

Therefore, the extraction of light hydrocarbons from the gas by a regenerated absorbent outside the absorber with the simultaneous removal of heat of absorption makes it possible to reduce the release of heat in the upper part of the apparatus and, as a result, to increase the degree of extraction of target hydrocarbons from the gas or to reduce the specific consumption of the absorbent with an unchanged degree of extraction of commercial products.

Calculation studies of the described schemes in relation to the conditions of a typical installation showed that with «average» gas fat content and low flow temperature (minus 37 °C), the extraction of propane is ensured in the first case by 76%, in the second by 83% and in the third by 91% (removal of butanes and heavier hydrocarbons was in all cases about 100%). Experience shows the possibility of reducing the specific absorbent consumption due to this measure by 20%.

Thus, the extraction of light hydrocarbons from the gas by a regenerated absorbent outside the absorber with the simultaneous removal of the heat of absorption allows to reduce the release of heat in the upper part of the apparatus and, as a result, to increase the degree of extraction of target hydrocarbons from the gas or to reduce the specific consumption of the absorbent with an unchanged degree of extraction of commercial products.

Basic material and results

Taking into account the fact that the existing LTA installation lacks technological solutions for removing excess heat along the height of the absorber or preventing its release in the absorber due to the pre-saturation of the regenerated absorbent with inert components (methane, ethane) and removal of excess heat before it is fed to the absorber, it should be optimized first of all, the temperature regime of the absorber. Based on the availability of a resource of inert components (self-needs gas consisting of the deethanization gas of the K-1 column and the weathering gas of the aerator B-1 (Fig. 1)), it is advisable to supplement the ILTA scheme with a node for saturating the regenerated absorbent with gas for own needs.

Figure 7 shows the ILTA scheme supplemented with the node of saturation of the regenerated absorbent.

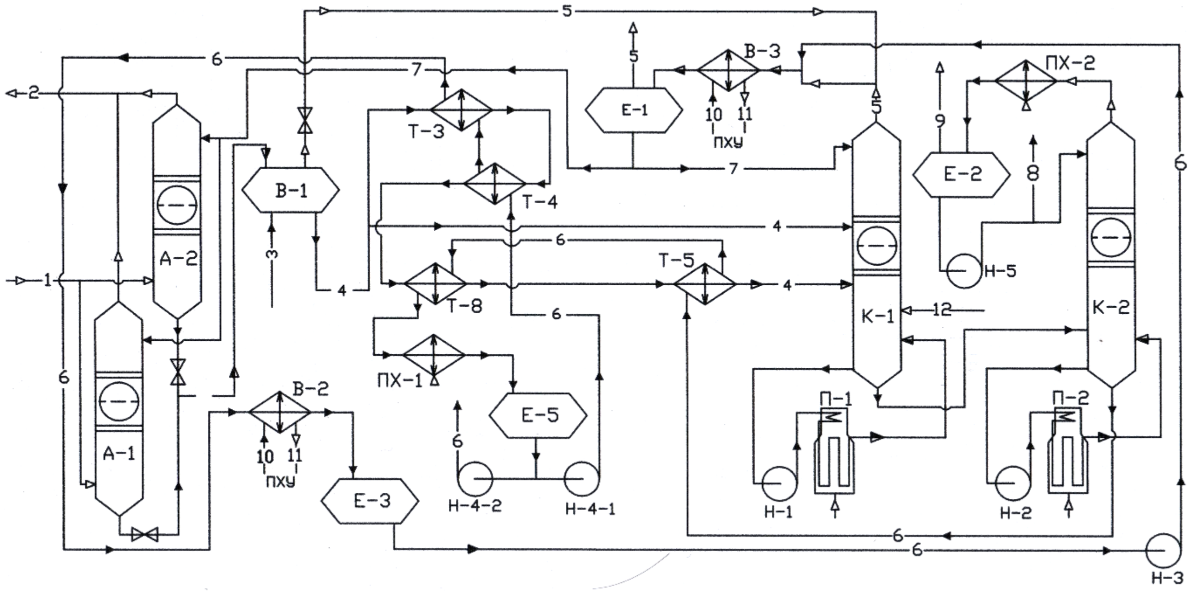

Figure 7. – Scheme of the ILTA with the node of saturation of the regenerated absorbent:
A-1, A-2 – absorbers; K-1 – absorption-evaporation column; K-2 – desorber; B-1 – weathering agent; T-3, T-4, T-5, T-8 – recuperative heat exchangers; P-1, P-2 – ovens; E-3, E-5 – collection tanks; E-2 – irrigation capacity; E-1 – separator; H-1, H-2, H-3, H-4-1, H-4-2, H-5 – pumps; B-2, B-3 – propane vaporizers of ACI; AC-1, AC-2 – air coolers; 1 – raw gas; 2 – dry gas; 3 – feeding with an absorbent; 4 – saturated absorbent; 5 – deethanization and weathering gas for own needs; 6 – regenerated absorbent; 7 – pre-saturated absorbent; 8 – BFLH; 9 – blowing on a torch; 10 – liquid propane; 11 – propane vapors; 12 – refueling with the vapor phase from the absorbent preparation unit

The process is implemented as follows. Gas from the LTS installation with a pressure of 2.2 MPa and a temperature of minus 21 °C in the amount of 130,000 m^3/h enters the lower part of two absorbers A-1 and A-2, irrigated with regenerated and cooled to a temperature of minus 11.6 °C absorbent.

The absorption process is accompanied by the release of heat (heating of the absorbent from minus 11.6 °C to minus 10 °C, and the gas from minus 21 to 5.9 °C).

The absorbent, saturated with BFLH, with a temperature of minus 10 °C and a pressure of 2.2 MPa is throttled from the absorbers to 1.6 MPa in the air conditioner B-1. As a result of the throttle effect, the absorbent temperature drops to minus 16.9 °C.

In the weathering device B-1, weathering gases (methane, ethane) are desorbed from the absorbent, which, through the regulating valve, are sent to the gas collector for its own needs. The degassed absorbent is sent from it in two streams to the K-1 deethanization column as feed. One stream with a temperature of minus 16.9°C (about 25% of the total feed) is supplied as a «cold» feed to the 30 plate of the column. The second flow is heated in heat exchangers T-3, T-4, T-8 and T-5 to a temperature of 200 °C by the flow of hot regenerated absorbent coming from the cube of column K-2 and sent as «hot» feed to plate 36 of column K-1.

As irrigation, regenerated, pre-saturated and cooled to a temperature of minus 11.6 °C absorbent is supplied to the first plate from the separator of the pre-saturation unit of the regenerated absorbent, maintaining the temperature of the top of the column to 10 °C. The heating of the liquid in the cube up to 170 °C is ensured by the circulation of the cube liquid according to the scheme: «cube column K-1 – pump H-1 – condensate heating furnace P-1 – cube column K-1».

In the K-1 column, light components (methane, ethane) are separated from the absorbent. The deethanization gases from the top of the K-1 column together with the weathering gas from the B-1 weatherer pass through the regenerated absorbent saturation node into the fuel gas system.

The deethanized absorbent with a temperature of 185°C and a pressure of 1.6 MPa is fed from the bottom of the K-1 column as feed to the K-2 absorbent regeneration column on a 36 plate.

In the K-2 column, the cubic residue of the K-1 column is separated into the regenerated absorbent and the HFLV extracted from the gas, which comes out in the form of a vapor phase from the upper part of the column, condenses in the AC-2 air coolers and enters the reflux tank at a temperature of 45-50°C E-2.

The pressure of 1.4 MPa in the K-2 column is maintained by changing the degree of condensation of BFLH in the air cooler AC-2.

From the reflex tank E-2, a part of broad fraction of light hydrocarbons (BFLH) is supplied in the form of irrigation to the first plate K-2, the rest is removed from the installation as a target product. The regenerated absorbent from the bottom of the column K-2 is successively cooled in the heat exchangers T-5 and T-8 by the flow of saturated absorbent and in the air cooling apparatus X-1 to a temperature of 50 °C and enters the tank E-5.

The regenerated absorbent from tank E-5 is sequentially cooled in the pipe space of heat exchangers T-4 and T-3 with saturated absorbent, in the pipe space of evaporator B-2 with boiling propane (to a temperature of minus 11.6 °C) and enters the buffer tank E-3.

The cooled regenerated absorbent from the E-3 tank is pumped by the H-3 pump through the pre-saturation unit, which consists of the B-3 propane evaporator and the E-1 separator. This is where it is saturated with deethanization gas and weathering, removal of heat of absorption (methane and ethane) and separation of the gas from the absorbent. After the saturation node, the gas enters the gas system of its own needs, and the pre-saturated absorbent with a temperature of minus 11.6 °C is divided into two streams, the first flows into absorbers A-1 and A-2, the second – into the deethanization column – as irrigation.

Absorption calculation

At the moment, the supply of regenerated absorbent to ILTA is limited by the capacity of the ACI and is 511 kmol/h. In the variant of pre-saturation of the regenerated absorbent, its amount increases from 511 to 584 kmol/h (by 14.4%) (Table 5). At the same time, absorption will receive 468 kmol/h instead of 409 kmol/h before reconstruction. Due to the reduction of the average absorption temperature to the regulatory level (minus 15 °C), the installation will reach the design level of extraction of BFLH (from 100 – 110 t/day to 150 t/day) (Table 6).

Conclusions

The application of the saturation of regenerated absorbent with inert components at the low-temperature absorption instalation of MF «Solokha» will allow to minimize the consequences of reducing the pressure of the inlet gas and increase the level of extraction of the components of BFLH by 21.96 % (including propane by 22.05 %, butanes by 30.75 %). At the same time, the extraction level will reach the design level.

The proposed version of the reconstruction of the installation involves supplementing the existing technological scheme with standard equipment, analogues of which are already used at the enterprise (propane evaporator, separator and mixing unit). Deethanization and weathering gases of the saturated absorbent act as raw materials for pre-saturation of the regenerated absorbent. Their consumption in conditions critical for the implementation of low-temperature absorption (reduced pressure and elevated temperature) increases 1.5-2 times.

Table 5 – Calculation of the saturation node of the regenerated absorbent (under existing conditions)

Component	*M*	Regenerated absorbent, T=-11.6 °C				Weathering gas К-1: T=-5.6 °C, P=1.6 MPa			
		mol/ mol	kg/ kmol	kmol/ h	kg/h	mol/ mol	kg/ kmol	kmol/ h	kg/h
N_2	28.01	0	0	0	0	0.001	0.036	0.152	4.258
CH_4	16.04	0	0	0	0	0.55	8.795	65.613	1052.43
CO_2	44.01	0	0	0	0	0.08	3.571	9.709	427.293
C_2H_6	30.07	0.00003	0.001	0.0153	0.4612	0.354	10.64	42.341	1273.19
C_3H_8	44.09	0.00736	0.325	3.7630	165.91	0.010	0.429	1.163	51.277
Iso-butane	58.1	0.00809	0.470	4.1362	240.31	0.001	0.066	0.136	7.902
n-butan	58.1	0.02322	1.349	11.872	689.75	0.002	0.128	0.264	15.338
iso-pentan	72.1	0.01805	1.301	9.2285	665.38	0.001	0.047	0.077	5.552
n-pentan	72.1	0.02590	1.867	13.242	954.75	0.001	0.048	0.080	5.768
Fr. to 80°C	85.7	0.0493	4.224	25.201	2159.7	0.001	0.046	0.060	5.142
80-100 °C	98.8	0.0830	8.200	42.436	4192.7	0	0.027	0.033	3.260
100-130 °C	110.1	0.1574	17.329	80.47	8859.7	0	0.020	0.022	2.422
130-160 °C	125.9	0.1530	19.266		9850.5	0	0.006	0.005	0.630
160-180 °C	140.1	0.18348	25.706	93.809	13142.6	0	0.003	0.002	0.280
180-200 °C	152.8	0.2008	30.676	102.64	15684		0.002	0.001	0.153
200-240 °C	172.6	0.072	12.49	36.98	6383.1				
240-290 °C	205.7		2.983	7.414	1525.0				
290-330 °C	242.5	0.003	0.783	1.651	400.5				
>330 °C	305.1	0.00034	0.104		53.037				
∑	-	1.000	127.07	511.28	64967.3	1.00	23.86	119.66	2854.90

Continuation of Table 5

Component	*M*	*Ki*	*Ai*	φi	1- φi	Dry gas	Pre-saturation of the absorbent	
						V1,i	***LN,i***	
		P= 1.6 MPa, T= -10 °C				kmol/h	kmol/h	mol/mol
N_2	28.01	32.0	0.24	0.59	0.406	0.062	0.090	0.0015
CH_4	16.04	9.5	0.80	0.45	0.55	36.087	29.526	0.05049
CO_2	44.01	2.9	2.62	0.72	0.28	2.719	6.990	0.1195
C_2H_6	30.07	1.5	5.07	0.83	0.17	7.198	35.158	0.06012
C_3H_8	44.09	0.42	18.10	0.95	0.05	0.058	4.868	0.00832
Iso-butane	58.1	0.17	44.71	0.99	0.01	0.001	4.271	0.00730
n-butan	58.1	0.11	69.09	1.0	0	0	12.136	0.02075
iso-pentan	72.1	0.042	181.0	1.0	0	0	9.306	0.01591
n-pentan	72.1	0.035	217.0	1.0	0	0	13.322	0.02278
Fr. to 80 °C	85.7	0.0108	704.0	1.0	0	0	25.261	0.04319
80-100 °C	98.8	0.0039	1949.0	1.0	0	0	42.469	0.07262
100-130 °C	110.1	0.0017	4471.0	1.0	0	0	80.492	0.13764
130-160 °C	125.9	0.0007	10857.0	1.0	0	0	78.245	0.13380
160-180 °C	140.1	0.0004	19000.0	1.0	0	0	93.811	0.16041
180-200 °C	152.8	0.0002	38000.0	1.0	0	0	102.645	0.17552
200-240 °C	172.6						36.981	0.06324
240-290 °C	205.7						7.413	0.01268
290-330 °C	242.5						1.651	0.00282
>330 °C	305.1						0.174	000030
∑	-					46.12489	584.81	10

Table 6 – Calculation of the absorber operation with the absorbent saturation unit

Component	*M*	Regenerated absorbent, T=-11.6 ºC				Dry gas, T=-21 ºC, P=2.2 MPa			
		mol/ mol	kg/ kmol	kmol/ h	kg/h	mol/ mol	kg/ kmol	kmol/ h	kg/h
N_2	28.01	0	0.004	0.04	1.01	0.014	0.401	38.69	1083.8
CH_4	16.04	0.050	0.810	11.81	189.49	0.864	13.864	2337.03	37485.9
CO_2	44.01	0.012	0.526	2.80	123.10	0.031	1.349	82.87	3647.2
C_2H_6	30.07	0.060	1.808	14.07	423.01	0.059	1.774	159.53	4797.0
C_3H_8	44.09	0.008	0.367	1.95	85.87	0.022	0.988	60.57	2670.4
Iso-butane	58.1	0.007	0.424	1.71	99.28	0.003	0.145	6.76	392.7
n-butan	58.1	0.021	1.206	4.86	282.12	0.005	0.261	12.17	706.9
iso-pentan	72.1	0.016	1.147	3.72	268.45	0.001	0.072	2.70	195.0
n-pentan	72.1	0.023	1.642	5.33	384.32	0.001	0.058	2.16	156.0
Fract. to 80 ºC	85.7	0.043	3.702	10.11	866.19	0.000	0.030	0.95	81.1
80-100 ºC	98.8	0.073	7.175	16.99	1678.86	0.000	0.013	0.35	34.7
100-30ºC	110.1	0.138	15.154	32.21	3545.89	0.000	0.002	0.05	5.95
130-60ºC	125.9	0.134	16.845	31.31	3941.60				
160-80ºC	140.1	0.160	22.474	37.54	5258.70				
180-00ºC	152.8	0.176	26.819	41.07	6275.48				
200-240C	172.6	0.063	10.915	14.80	2554.03				
240-290 ºC	205.7	0.013	2.608	2.97	610.16				
290-330 ºC	242.5	0.003	0.685	0.66	160.23				
>330 ºC	305.1	0	0.091	0.07	21.22				
∑	-	1.000	114.40	233.99	26769.03	1.000	18.957	2703.83	51256.5

Continuation of Table 6

Component	*M*	***Ki*** -15,5ºC	***Ai***	**φ**i	1- **φ**i	Dry gas	Saturated absorbent	
						V1,i *k*mol/h	*LN,i* kmol/h	*G*$G_{abs..}$ kg/h
N_2	28.01	30.0	0.003	0.003	0.997	38.576	0.52	4.259
CH_4	16.04	7.0	0.013	0.013	0.987	2307.014	41.828	670.920
CO_2	44.01	2.8	0.032	0.032	0.968	80.212	5.458	240.198
C_2H_6	30.07	1.19	0.076	0.076	0.924	147.474	26119	785.40
C_3H_8	44.09	0.31	0.290	0.290	0.710	43.002	19.512	860.28
Iso-butane	58.1	0.13	0.790	1.000	0	0	8.468	492.02
n-butan	58.1	0.08	0.773	1.000	0	0	17.023	989.04
iso-pentan	72.1	0.03	0.692	1.000	0	0	6.427	463.40
n-pentan	72.1	0.025	0.605	1.000	0	0	7.493	540.28
Fraction to 80 ºC	85.7	0.008	0.306	1.000	0	0	11.054	947.29
80-100 ºC	98.8	0.0027	0.110	1.000	0	0	17.344	1713.59
100-130ºC	110.1	0.0013	0.001	1.000	0	0	32.260	3551.85
130-160ºC	125.9						31.307	394160
160-180ºC	140.1						37.535	5258.70
180-200ºC	152.8						41.070	6275.48
200-240ºC	172.6						14.797	2554.03
240-290ºC	205.7						2.966	610.16
290-330ºC	242.5						0.661	160.24
>330 ºC	305.1						0.070	21.22
∑	-	-	-	-	-	2586.96	321.54	30079.93

This actualizes the problem of its more efficient use, increases the pressure in the gas system of own needs and leads to non-productive emissions of valuable raw materials through the safety valves for combustion in the flare system. However, in the case of using these gases for pre-saturation of the absorbent, about 19% of the low-pressure gas is absorbed and returned to the process. As a result, the percentage of marketable products (BFLH and dry gas) is increasing.

Table 7- 9 shows the material balance of absorption.

Table 7 – Calculation of extraction coefficients

Component	Absorption, T=21.1 °C, P=1.58MPa			Desorption, T=153 °C, P=1.61MPa		
	***K*i at 21 °C**	***Ai***	**φ'i**	***K'*i**	***S'i***	**φ''i**
N_2	40.000	0.019	0.019	60.0	9.9740	1.0
CH_4	10.000	0.077	0.077	17.3	2.8758	1.0
CO_2	8.200	0.094	0.094	18.25	3.0338	1.0
C_2H_6	2.000	0.385	0.385	6.8	1.1304	0.88
C_3H_8	0.642	1.200	0.95	3.85	0.64	0.64
Iso-butane	0.270	2.855	1.0	2.2	0.3657	0.36571
n-butan	0.160	4.817	1.0	1.9	0.3158	0.31584
iso-pentan	0.080	9.635	1.0	1.05	0.17455	017455
n-pentan	0.060	12.846	1.0	0.93	0.1546	0.15460
Fraction up to 80 °C	0.024	32.115	1.0	0.56	0.0931	0.09309
80-100 °C	0.009	85.640	1.0	0.3	0.0499	0.0499
100-130 °C	0.005	154.152	1.0	0.186	0.0309	0.03092
130-160 °C	0.003	296.446	1.0	0.093	0.0155	0.0155
160-180 °C	0.001	7707.600	1.0	0.052	0.0086	0.0086
180-200 °C	0	2569.200	1.0	0.0295	0.0049	0.00490
200-240 °C	0	7707.600	1.0	0.0165	0.0027	0.00274
240-290 °C	0		1.0	0.0083	0.00138	0.0014
290-330 °C	0		1.0	0.0046	0.00076	0.00077
>330 °C	0		1.0	0.0022	0.00037	0.00037

Table 8 ‒ Material balance of ILTA MF «Solokha» after reconstruction with a «raw» gas productivity of 130,000 m^3/h

Component	Gas inlet in ILTA		Degree withdrawal	Entrance		Withdrawal of BFLH		
	% mol.	g / m^3	% mol.	% mol.	g/m^3	g/m^3	t/h	t/day
C_3	1.791	43.0	57.61	0.759	18.23	24.77	3.2204	77.29
$\sum C_4$	0.701	17.7	86.61	0.094	2.37	15.33	1.9925	47.82
$\sum C_5$	0.179	5.40	100	0	0	5.40	0.7021	16.85
$\sum C_{6+}$	0.05	1.88	100	0	0	1.88	0.2438	5.85
$\sum CO_2, CH_4, C_2H_6$	-	-	-	-	-	-	-	7.35
$\sum$	2.72	67.98	69.70	0.853	20.6	47.38	6.1588	155.16

Table 9 – Results of the introduction of the node of saturation of the regenerated absorbent with inert components

Component	Withdrawal of BFLH						Difference		
	after reconstruction			before reconstruction					
	g/m^3	t/day	% mol.	g/m^3	t/day	%mol.	g/m^3	t/day	% mol.
C_3	24.77	77.29	57.61	15.29	47.70	35.56	+9.48	+29.59	+22.05
$\sum C_4$	15.33	47.82	86.61	9.89	30.86	55.86	+5.44	+16.96	+30.75
$\sum C_5$	5.40	16,85	100	5,40	16,85	100	-	-	-
$\sum C_{6+}$	1.88	5.85	100	1.88	5.85	100	-	-	-
$\sum CO_2, CH_4, C_2H_6$	-	7.35	-	-	3.48	-	-	+3.87	-
$\sum$	47.38	155.16	69.70	32.46	104.74	47.74	+14.92	+50.42	+21.96

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